GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN CIRCULATING FLUIDIZED BED BOILERS: PHASE 1 – A PRELIMINARY SYSTEMS EVALUATION

FINAL REPORT

VOLUME I
EVALUATION OF ADVANCED COAL COMBUSTION & GASIFICATION POWER PLANTS WITH GREENHOUSE GAS EMISSION CONTROL

VOLUME II
BENCH-SCALE FLUIDIZED BED COMBUSTION TESTING

SUBMITTED BY

ALSTOM POWER INC.
POWER PLANT LABORATORIES
2000 DAY HILL ROAD
WINDSOR, CT 06095
(860) 688-1911

PREPARED FOR

NETL AAD DOCUMENT CONTROL BLDG. 921
US DEPARTMENT OF ENERGY
NATIONAL ENERGY TECHNOLOGY LABORATORY
P.O. BOX 10940
PITTSBURGH, PENNSYLVANIA 15236-0940
(COOPERATIVE AGREEMENTT NO. DE-FC26-01NT41146)

PHASE I REPORT SUBMITTAL DATE: MAY 15, 2003
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PPL REPORT NO. PPL-03-CT-09

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ALSTOM POWER INC.
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2000 DAY HILL ROAD
WINDSOR, CT 06095
(860) 688-1911

Project Manager: John L. Marion
Project Adviser: Carl R. Bozzuto

Principal Investigators:
Nsakala ya Nsakala
Gregory N. Liljedahl

Project Technical Expert: Herbert E. Andrus
Project Economics Expert: Ray P. Chamberland

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DISCLAIMER

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  - Richard Waryasz: For providing design and performance information on external and moving bed heat exchangers.
  - John Chiu: For providing cyclone and particulate removal system design and performance information.
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ACRONYMS AND ABBREVIATIONS

acfm	Actual cubic feet per minute
AGR	Acid Gas Removal
AGR	Air Gas Removal
ANSI	American National Standards Institute
ASFH	Air Suction Filter House
ASME	American Society of Mechanical Engineers
ASU	Air Separation Unit
ATS	Advanced Turbine System
BI	Boiler Island
BOP	Balance of Plant
BSR	Beaven Sulfur Removal
Btu	British Thermal Unit
CFB	Circulating Fluidized Bed
cfm	Cubic Feet per Minute
CGE	Cold Gas Efficiency
CHAT	Cascaded Humidified Advanced Turbine
CLC	Chemical Looping Combustion
CLG	Chemical Looping Gasification
CMB	Circulating Moving Bed
CO_2	Carbon Dioxide
COE	Cost of Electricity
CRT	Cathode Ray Tube
CS	Carbon Steel
CT	Combustion Turbine
dB	Decibel
DCA	Direct Contact Aftercooler
DCS	Distributed Control System
DGC	Dakota Gasification Company
DOE/NETL	Department of Energy/National Energy Technology Laboratory
ECBM	Enhanced Coal Bed Methane
EGR	Enhanced Gas Recovery
EOR	Enhanced Oil Recovery
EPC	Engineered, Procured and Constructed
ESP	Electrostatic Precipitator
FBC	Fluidized Bed Combustion
FD	Forced Draft
FDA	Flash Drier Absorber
FGD	Flue Gas Desulfurization
FOM	Fixed Operation & Maintenance
GI	Gasifier Island
gpm	Gallons per minute
GPS	Gas Processing System
GT	Gas Turbine
HHV	Higher Heating Value
HP	High Pressure
hp	Horse Power
hr	Hour

HRSG	Heat Recovery Steam Generator
HT	High Temperature
HTAH	High Temperature Air Heater
HTSGH	High Temperature Sweep Gas Heater
HVAC	Heating, Ventilating and Air Conditioning
Hz	Hertz
ID	Induced Draft
IGCC	Integrated Gasification Combined Cycle
in. H_2O	Inches of Water
in. Hga	Inches of Mercury, Absolute
IP	Intermediate Pressure
ISO	International Standards Organization
kV	Kilovolt
kWe	Kilowatts electric
kWh	Kilowatt-hour
lbm	Pound mass
LHV	Lower Heating Value
LLHR	Low Level Heat Recovery
LP	Low Pressure
LT	Low Temperature
LTSGH	Low Temperature Sweep Gas Heater
MBHE	Moving Bed Heat Exchanger
MCR	Maximum Continuous Rating
MDEA	Methyl Diethanolamine
MTF	Multi-use Test Facility
MWe	Megawatt electric
N_2	Nitrogen Gas
NGCC	Natural Gas Combined Cycle
NPHR	Net Plant Heat Rate
O&M	Operation & Maintenance
OTM	Oxygen Transport Membrane
PA	Primary Air
PC	Pulverized Coal
PFD	Process Flow Diagram
PFWH	Parallel Feedwater Heater
PHX	Primary Heat Exchanger
ppm	Parts per million
psia	Pound per square inch, absolute
psig	Pound per square inch, gauge
RSC	Radiant Syngas Cooler
SA	Secondary Air
TGA	Thermo-Gravimetric Analysis
TPD	Ton Per Day
TPH	Ton Per Hour
TSA	Temperature Swing Adsorption
UBC	Unburned Carbon
VOM	Variable Operation & Maintenance

EXECUTIVE SUMMARY

Background

Because fossil fuel fired power plants are among the largest and most concentrated producers of CO_2 emissions, recovery and sequestration of CO_2 from the flue gas of such plants has been identified as one of the primary means for reducing anthropogenic CO_2 emissions. In this study, ALSTOM Power Inc. (ALSTOM) has investigated several coal fired power plant configurations designed to capture CO_2 from effluent gas streams for sequestration.

Burning fossil fuels in mixtures of oxygen and recirculated flue gas (made principally of CO_2) essentially eliminates the presence of atmospheric nitrogen in the flue gas. The resulting flue gas is comprised primarily of CO_2, along with some moisture, nitrogen, oxygen, and trace gases like SO_2 and NOx. Oxygen firing in utility scale Pulverized Coal (PC) fired boilers has been shown to be a more economical method for CO_2 capture than amine scrubbing (Bozzuto, et al., 2001). Additionally, oxygen firing in Circulating Fluid Bed Boilers (CFB's) can be more economical than in PC or Stoker firing, because recirculated gas flow can be reduced significantly. Oxygen-fired PC and Stoker units require large quantities of recirculated flue gas to maintain acceptable furnace temperatures. Oxygen-fired CFB units, on the other hand, can accomplish this by additional cooling of recirculated solids. The reduced recirculated gas flow with CFB units results in significant Boiler Island cost savings.

Additionally, ALSTOM has identified several advanced/novel plant configurations, which improve the efficiency and cost of the CO_2 product cleanup and compression process. These advanced/novel concepts require long development efforts. An economic analysis indicates that the proposed oxygen-firing technology in circulating fluidized boilers could be developed and deployed economically in the near future in enhanced oil recovery (EOR) applications or enhanced gas recovery (EGR), such as coal bed methane recovery.

ALSTOM received a Cooperative Agreement from the US Department of Energy National Energy Technology Laboratory (DOE) in 2001 to carry out a project entitled "Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers." This two-phased project is in effect from August 31, 2001, to October 27, 2004. (U.S. DOE NETL Cooperative Agreement No. DE-FC26-01NT41146). This project entails a comprehensive study evaluating the technical feasibility and economics of alternate CO_2 capture technologies applied to Greenfield US coal-fired electric generation power plants. Thirteen separate but related cases (listed below), representing various levels of technology development, were evaluated as described herein. The first seven cases represent coal combustion cases in CFB type equipment. The next four cases represent Integrated Gasification Combined Cycle (IGCC) systems. The last two cases represent advanced Chemical Looping systems, which were completely paid for by ALSTOM and included herein for completeness.

Combustion Cases:

- Case-1: Air Fired Circulating Fluidized Bed (CFB) without CO_2 Capture (Base Case for Comparison to Cases 2-7)
- Case-2: Oxygen Fired CFB with CO_2 Capture
- Case-3: Oxygen Fired CFB with CO_2 Capture (sequestration only option)
- Case-4: Oxygen Fired Circulating Moving Bed (CMB) with CO_2 Capture (advanced boiler concept)
- Case-5: Air Fired CMB with CO_2 Capture utilizing Regenerative Carbonate Process

- Case-6: Oxygen Fired CMB with Oxygen Transport Membrane (OTM) and CO_2 Capture
- Case-7: Indirect Combustion of Coal via Chemical Looping and CO_2 Capture

IGCC Cases:
- Case-8: Built and Operating Present Day IGCC without CO_2 Capture (Base Case for Comparison with Case-9)
- Case-9: Built and Operating Present Day IGCC with shift reaction and CO_2 Capture
- Case-10: Commercially Offered Future IGCC without CO_2 Capture (Base Case for Comparison with Case-11)
- Case-11: Commercially Offered Future IGCC with shift reaction and CO_2 Capture

Advanced Chemical Looping Cases:
- Case-12: Indirect Gasification of Coal via Chemical Looping (Base Case for comparison to Case-13)
- Case-13: Indirect Gasification of Coal and CO_2 Capture via Chemical Looping

ALSTOM managed and performed the subject study from its US Power Plant Laboratories office in Windsor, Connecticut. Participating as sub-contractors in this effort are ABB Lummus Global, from its offices in Houston, Texas; Parsons Energy and Chemical Group, from its offices in Wyomissing, Pennsylvania; and Praxair Inc. from its offices in Tonawanda, New York. Plasma Inc. of Butte, Montana is serving as an informal consultant. The US Department of Energy National Energy Technology Laboratory provided consultation and funding. ALSTOM provided cost share to this project.

Conclusions

- There are a number of viable approaches to CO_2 capture for sequestration in solid fuel-fired power plants

- In the long term, the potential for advanced Chemical Looping with CO_2 capture appears to come closest to approaching the target economic values:
  - Calculated cost of electricity closely approaching the Base Case COE without CO_2 capture on the same basis (about 15 percent higher).
  - Avoided cost of CO_2 = $11/ton of CO_2 vs. target of $3/ton of CO_2 or $10/ton of carbon.

- Nearer term, chemical looping combustion and the high temperature regenerative carbonate process alternatives appear to be economically more attractive than IGCC for CO_2 capture.

- Oxygen-Fired CFB/CMB alternatives can be competitive under appropriate niche conditions (EOR) and are available in the near term.

- The Oxygen-Fired CFB/CMB with flue gas sequestration provides the only true “Zero Gaseous Emissions” power plant studied herein, as all flue gas was captured, and compressed for sequestration. The feasibility of sequestering the entire “dry” flue gas was confirmed by Plasma, Inc. (Schuller, 2003).

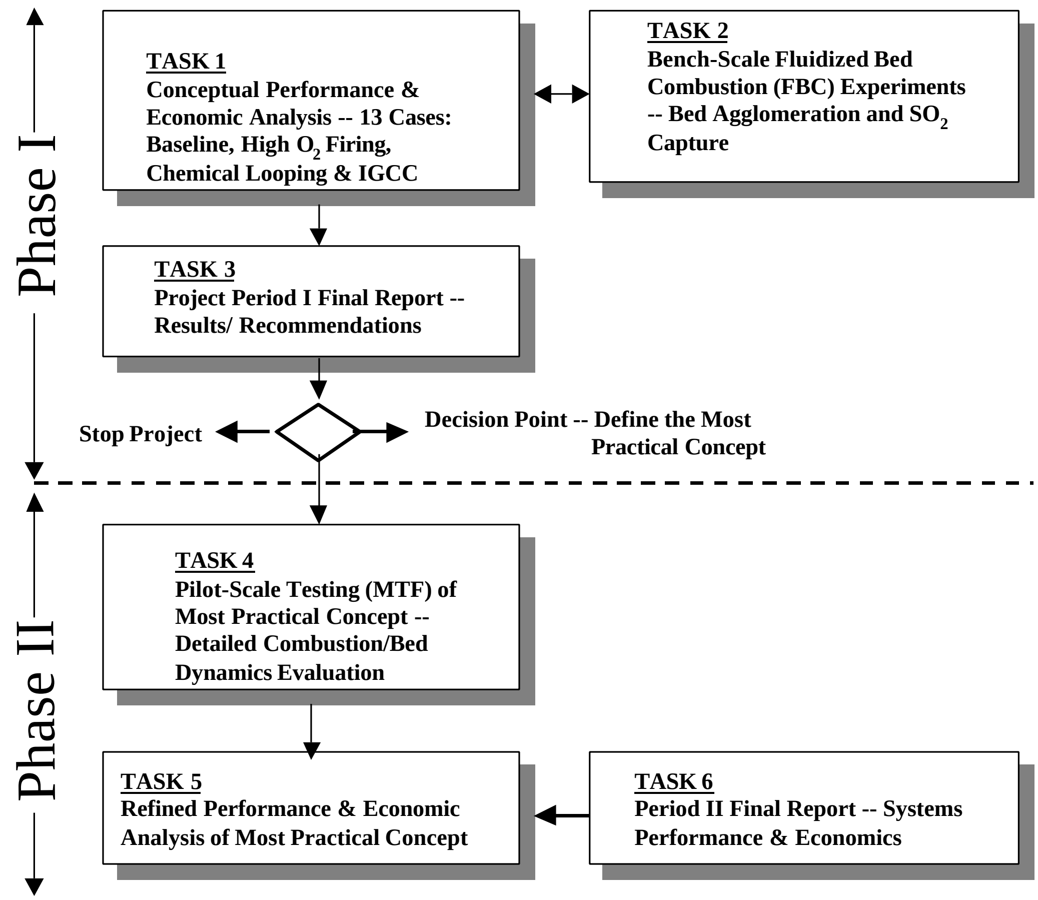
- Oxygen-Fired CFB/CMB can provide for the near term demonstration of the circulating moving bed technology, which is an enabling technology for the combustion and advanced Chemical Looping cases.

TECHNICAL SUMMARY

This technical summary provides a brief review of the overall project work scope, Phase I result summary (plant efficiency, plant investment and O&M costs, levelized cost of electricity, CO_2 mitigation costs, and CO_2 emissions), and technical conclusions.

Project Work Scope

The Work breakdown Structure of each Phase of the two-phase effort is shown in the figure below.



<u>Phase I</u>
The Phase I workscope, which is the subject of this report, is comprised of three tasks, as follows:

- Task 1: Conceptual Performance and Economic Analyses of Thirteen Study Cases
- Task 2: Bench-Scale Fluidized Bed Combustion (FBC) Testing
- Task 3: Phase I Final Report.

The major work within Phase I was Task 1. The key goals of Task 1 were to evaluate the impacts on the plant output, efficiency, and CO_2 emissions, resulting from the addition of various CO_2 capture systems to an array of CFB combustion based, IGCC based, and advanced Chemical Looping based power plants. Cost estimates were developed for these power plants including estimates for the traditional power plant equipment as well as additional non-traditional systems required to produce, extract, clean, compress and liquefy the CO_2, which could then be available for use in enhanced oil or gas recovery or sequestration. Additionally, the impact of CO_2 capture on the levelized cost of electricity (COE) and on the mitigation cost for CO_2 (\$/ton of CO_2 avoided) were also evaluated.

Task 2, Bench Scale FBC testing, supported the Task 1 case studies. The objective of Task 2 is to derive pertinent combustion performance and bed dynamic information under highly controlled operating conditions in a 4-inch fluidized bed test facility. Results from various oxy-fuel firing of three fuels, two coals and one delayed petroleum coke, are to be compared to those obtained similarly from air firing. Key Outputs include:

- Bed and ash characteristics (e.g., potential bed agglomeration/sintering)
- Gaseous emissions (NOx, SO_2 and CO)
- Desulfurization potentials
- NOx emissions reductions
- Unburned carbon (UBC) emissions

<u>Phase II</u>
The overall objective of the Phase II workscope is to generate a refined technical and economic evaluation of the case (concept) recommended, based on Phase I information, with the benefits from pilot-scale testing of the same concept. The Phase II workscope has been developed based upon the findings from Phase I and will specifically address both retrofit (moderate O_2 enrichment/ high flue gas recirculation) and Greenfield (Case 2) applications (high oxygen enrichment/low flue gas recirculation). The objective of the pilot-scale testing will be to generate detailed technical data needed to establish advanced CFB design requirements and performance when firing coals and delayed petroleum coke in O_2/CO_2 mixtures. Firing rates in the pilot test will range from 4.0 to 9.5 MM-Btu/hr. Pilot-scale testing will be performed at ALSTOM's Multi-use Test Facility (MTF), located in Windsor, Connecticut. Outputs from this testing will address key technical parameters including:

- Flue Gas Quality
- Bed Dynamics
- Heat Transfer to the Waterwalls
- Flue Gas Desulfurization
- NOx Emissions Reduction
- Other Pollutants' Emissions (N_2O and CO)
- Bed and Ash Characteristics (e.g., Potential Bed Agglomeration)

Results will be used for the design of units retrofitted for oxygen firing and for the design of new oxygen-fired CFB boilers. Results will also provide a generic performance database for other researchers. At the conclusion of Phase II, revised costs will be estimated and performance calculations will be updated for the design concept evaluated (i.e., Case 2: New Compact O_2-Fired CFB with CO_2 Capture, Purification, Compression, and Liquefaction).

Phase I Results Summary

This Section provides a brief description of the primary results from the Phase I work. Task 1 results, which represents the major part of this work, are presented first followed by Task 2 results. Finally, Technical Conclusions and Recommendations for Future Work are described.

Task 1 Results

The results for Task-1, which are described in some detail in the Volume I report, are summarized below. The summary is divided into several sections: System Descriptions, Performance Analysis Results, Plant CO_2 Emissions, Plant Costs, and Economic Evaluations.

System Descriptions:
A total of thirteen (13) Greenfield case studies, listed below, were analyzed in this evaluation. The thirteen cases were subdivided into three groups. Seven of the cases were grouped as Coal Combustion cases, four were IGCC cases, and two were advanced Chemical Looping cases. One Combustion case, two IGCC cases and one advanced Chemical Looping case were analyzed without CO_2 capture. These cases without CO_2 capture represent Base Cases for comparison with the respective CO_2 capture cases. Inclusion of the Base Cases allows accurate quantification of the impact of CO_2 capture and gas processing on plant efficiency, cost, and cost of electricity. CO_2 mitigation costs ($/Ton of CO_2 avoided) were calculated relative to the appropriate Base Case. Within each technology group, the order of the various cases roughly represents increasing levels of technology development complexity (i.e., within the combustion cases, Case-7 would require the most development and Case-2 the least).

The following provides a brief description of the thirteen cases analyzed in this study.

<u>**Combustion Cases:**</u>
Case-1: Air Fired CFB without CO_2 Capture (Base Case for Comparison to Cases 2-7)
Conventional air-fired CFB without CO_2 capture using 1,800 psig / 1,000 ^oF / 1,000 ^oF steam cycle.
<u>Implication:</u> Provides a reference point for performance and economic analyses of the CO_2 capture Cases 2-7.

Case-2: New Compact Oxygen-Fired CFB with CO_2 Capture, Purification, Compression and Liquefaction.
Nearly identical steam cycle as Case-1 (same main and reheat steam conditions - flow, temperature, and pressure) but smaller Boiler Island equipment than Case-1. Oxygen is supplied from a Cryogenic Air Seperation Unit (ASU) Plant. The CFB Boiler Island provides a concentrated CO_2 flue gas product stream to the Gas Processing System.
<u>Implication:</u> Cost savings for the Boiler Island. Cost savings for the Gas Processing System equipment as compared to amine scrubbing systems. Improved plant thermal efficiency as compared to amine based CO_2 capture systems.

Case-3: Oxygen-Fired CFB with CO_2 Capture (sequestration only option).
Same as Case-2, but uses a simplified Gas Processing System whereby the product gas stream is not purified and therefore is suitable for sequestration only.
<u>Implication:</u> Further cost savings (Gas Processing System) are realized as compared to Case-2.

Case-4: Oxygen-Fired Circulating Moving Bed (CMB) with CO_2 Capture, Purification, Compression and Liquefaction.
Same as Case-2, but uses advanced boiler design concepts.
<u>Implication:</u> Anticipated further cost savings on Boiler Island equipment are expected as compared to Case-2.

Case-5: Air-Fired CMB Boiler with CO_2 Capture Utilizing High Temperature Regenerative Carbonate Process.
Utilizes air-firing and carbonate regeneration at higher temperatures than steam cycle temperatures. All the energy rejected from the carbonate regeneration process is recovered in the steam cycle at high temperature such that there is no efficiency penalty associated with CO_2 capture for this process. Nearly pure CO_2 is removed continuously from a calciner within the Boiler Island. This CO_2 stream is then cooled and ready for compression and liquefaction.
Implication: Advanced novel boiler and CO_2 capture concept eliminates the high power requirement of the cryogenic ASU used in Cases 2, 3, and 4 and eliminates the energy penalty typically associated with CO_2 capture for significant plant thermal efficiency improvement. Significant Boiler Island cost savings are also anticipated for this case.

Case-6: The Case-4 CMB process, Integrated with Oxygen Transport Membrane (OTM) instead of a cryogenic ASU.
Utilization of an OTM is a more efficient method for oxygen production as compared to a cryogenic ASU as was used with Cases 2, 3, and 4. The CMB process also has a 2,000 ^oF solids stream available for high temperature air heating as required by the OTM.
Implication: Significant plant thermal efficiency improvement as compared to Cases 2, 3, and 4.

Case-7: Indirect Combustion of Coal via Chemical Looping.
This air-fired boiler utilizes a continuously looping solid oxygen-carrier which oxidizes the fuel into primarily H_2O and CO_2 . Simple condensation of the H_2O then yields a fairly pure CO_2 product stream for compression and liquefaction.
Implication: Advanced novel boiler concept eliminates the high power requirement of the cryogenic ASU used in Cases 2, 3, and 4 and eliminates the energy penalty typically associated with CO_2 capture for significant plant thermal efficiency improvement. Significant Boiler Island cost savings are also anticipated for this case.

IGCC Cases:

Case-8: Built and Operating Present Day IGCC plant without CO_2 Capture (Base Case for Comparison with Case-9).
IGCC without CO_2 capture utilizing Texaco pressurized (i.e., 450 psig), oxygen-blown, entrained flow gasification technology (based on Tampa IGCC design), and a single train GE- 7FA gas turbine with HRSG and 1,800 psig / 1,000 ^oF / 1,000 ^oF steam cycle.
Implication: Provides a reference point for the performance and economic analyses of Case 9.

Case-9: Built and Operating Present Day IGCC plant with shift reaction and CO_2 Capture.
Same as Case-8 but with shift reactor and CO_2 capture, compression and liquefaction system.
Implication: Provides direct comparison with Case-8 to isolate the impact of CO_2 capture for present day IGCC plants.

Case-10: Commercially Offered Future IGCC plant without CO_2 Capture (Base Case for Comparison with Case-11).
IGCC without CO_2 capture utilizing Texaco high pressure (i.e., 950 psig), oxygen-blown, entrained flow, quench gasification technology with syngas expander (based on Eastman Chemical Company Acetic Anhydride design), and a single train GE-7FA gas turbine with HRSG and 1,800 psig / 1,000 ^oF / 1,000 ^oF steam cycle.
Implication: Plant cost improvement as compared to Case 8.

Case-11: Commercially Offered Future IGCC plant with shift reaction and CO_2 Capture. Same as Case-10 but with shift reactor and CO_2 capture, compression and liquefaction system.
Implication: Provides direct comparison with Case-10 to isolate the impact of CO_2 capture for commercially offered future IGCC plants.

Advanced Chemical Looping Cases:
Case-12: Indirect Gasification of Coal via Chemical Looping (Base Case for comparison to Case-13)
Advanced Chemical Looping without CO_2 capture utilizing a single train GE-7FA gas turbine with HRSG and 1,800 psig / 1,000 ^{o}F / 1,000 ^{o}F steam cycle.
Implication: Provides a point of reference for the performance and economic analyses of Case 13.

Case-13: Indirect Gasification of Coal and CO_2 Capture via Chemical Looping.
Same as Case-12 but with CO_2 capture, compression and liquefaction system.
Implication: Provides direct comparison with Case-12 to isolate the impact of CO_2 capture for advanced Chemical Looping based future plants.

Common Parameters and Assumptions for the Case Studies:
All plants were designed for the identical coal and limestone analyses, ambient conditions, site conditions, etc. such that each case study provides results which are directly comparable, on a common basis, to all other cases analyzed within this work. The ambient conditions used for all material and energy balances were based on the standard American Boiler Manufacturers Association (ABMA) atmospheric conditions (i.e. 80 ^{o}F, 14.7 psia, 60 percent relative humidity).

The steam cycle represents another common thread among the cases. It is nearly identical for all the combustion cases differing only in the arrangement of the low level heat recovery systems or small process steam extractions in some cases (Case-5 and Case-7). The steam turbine for the combustion cases is a nominal 210 MWe single reheat machine with steam conditions of 1,800 psig 1,000 ^{o}F / 1,000 ^{o}F and a condenser pressure of 3.0 in. Hga. The main steam flow is identical for all the combustion cases. The reheat steam flow is also identical for all the combustion cases except for a slight increase in Case-6. Six extraction feedwater heaters are used and the final feedwater temperature is 470 ^{o}F.

The steam cycles utilized for the IGCC and advanced Chemical Looping cases use the same steam conditions and expansion line as for the combustion cases but with somewhat different steam flows and process steam extractions as required by the respective gasifier, gas turbine and heat recovery arrangement combination. Additionally, no extraction feedwater heaters are used due to the low level heat recovery requirements of these combined cycles.

The CO_2 capture systems were designed for a minimum of 90 percent CO_2 capture. The Dakota Gasification Company's CO_2 specification (DGC WebPages, 2001) for EOR, given in the following table, was used as the basis for the CO_2 capture system design.

Dakota Gasification Project's CO_2 Specification for EOR

Constituant	Units	Value
CO_2	vol. %	96.0
H_2S	vol. %	0.9
CH_4	vol. %	0.7
C_2 + HC's	vol. %	2.3
CO	vol. %	0.1
N_2	vppm	< 300
H_2O	vppm	< 20
O_2	vppm	< 50

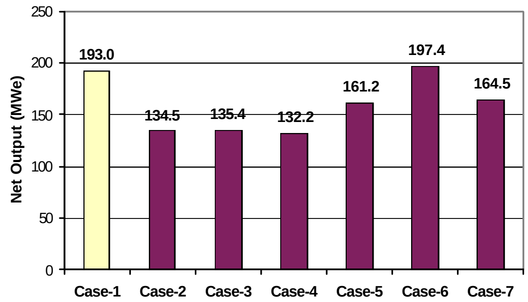
Performance Analysis Results:
The performance results for the seven combustion cases are presented first followed by the performance results for the four IGCC and two advanced Chemical Looping cases.

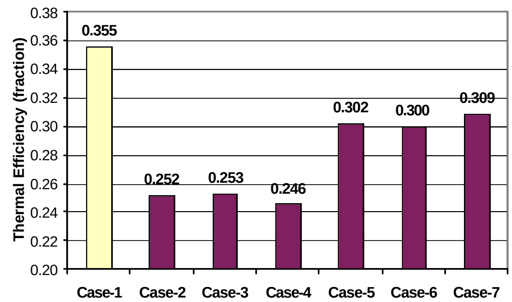
<u>Combustion Cases:</u>
The table shown below summarizes the performance differences between the combustion cases. The primary design constraint among all cases was the supply of an equivalent main steam flow to the steam turbine.

		CFB Air Fired (Case 1)	CFB Cryogenic O_2 Fired (Case 2)	CFB Cryogenic O_2 Fired (Case 3)	CMB Cryogenic O_2 Fired (Case 4)	CMB Air Fired High Temp Carb Proc (Case 5)	CMB with OTM O_2 Fired (Case 6)	CMB Chemical Looping (Case 7)
Auxiliary Power Summary								
Traditional Power Plant Auxiliary Power	(kW)	16007	10983	10687	12888	18887	14570	12833
Air Separation Unit or Fuel Compressor	(kW)	n/a	37505	37505	37800	n/a	n/a	n/a
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a	n/a	n/a	n/a	110920	n/a
CO_2 Removal System Auxiliary Power	(kW)	n/a	26905	26364	27200	22878	33434	25453
Total Auxiliary Power	(kW)	16007	75393	74556	77888	41765	158923	38287
	(frac. of Gen. Output)	0.077	0.359	0.355	0.371	0.206	0.196	0.189
Output and Efficiency								
Main Steam Flow	(lbm/hr)	1400555	1400555	1400555	1400555	1400555	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8256	8256	8275	8397	8758	8404
OTM System Expander Generator Output	(kW)	n/a	n/a	n/a	n/a	n/a	122659	n/a
Gas Turbine Generator Output		n/a	n/a	n/a	n/a	n/a	n/a	n/a
Steam Turbine Generator Output	(kW)	209041	209907	209907	210056	202949	233699	202770
Net Plant Output	(kW)	193034	134514	135351	132168	161183	197435	164483
	(frac. of Case-1 Net Output)	1.00	0.70	0.70	0.68	0.84	1.02	0.85
Boiler Efficiency (HHV)[1]	(fraction)	0.8946	0.9412	0.9412	0.9366	0.9217	0.9404	0.9242
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855	1806	1806	1820	1815	2242	1810
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	16.5	20.6	16.6	7.9	4.8	7.9
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855	1822	1826	1836	1822	2247	1818
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	13546	13492	13894	11307	11380	11051
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.2520	0.2530	0.2456	0.3019	0.2999	0.3088
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.71	0.71	0.69	0.85	0.84	0.87

[1] Boiler Heat Output / (Qcoal-HHV + Qcredits)
[2] Required for GPS Desiccant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4

As shown in the above table and the figure on the left below, significant reductions in Net Plant Output are incurred as a result of the CO_2 capture systems. The Base Case (Case-1), which does not include CO_2 capture, has a net output of about 193 MWe. The net output for the CO_2 capture cases ranges from about 132 – 198 MWe. It should be pointed out that Case-6 (the OTM Case), which has a net output of about 198 MWe, has a significantly higher coal heat input than the other cases (about 21-24 percent higher), as shown in the above table. The additional coal heat input for Case-6 is a result of high temperature air heating duty as required by the OTM. These output reductions are a direct result of increases in auxiliary power as required by the Gas Processing Systems and oxygen production systems.





Net plant efficiency is reduced from about 35.5 percent (HHV basis) for the Base Case (Case-1) to a range of about 25 - 31 percent for the CO_2 capture cases as shown by the figure on the right. These efficiencies represent energy penalties ranging from 13 – 30 percent. The higher efficiency reductions shown for Cases 2, 3, and 4 are due to significant auxiliary power increases, relative to the other capture cases, as a result of using cryogenic Air Separation Units.

Case 7, the chemical looping combustion case, shows the highest net plant thermal efficiency at about 30.9 percent. For this case the efficiency reduction is almost entirely due to the power required for the compression and liquefaction of the captured CO_2. For this case there is essentially no energy penalty associated with the capture of CO_2 other than the energy required to recirculate the solids between the oxidizer and reducer vessels.

Case 5, the regenerative carbonate case, was found to be second best with a thermal efficiency of about 30.2 percent. For Case-5 there is also essentially no energy penalty associated with the capture of CO_2 because the separation of CO_2 and the regeneration of the sorbent occurs at high temperature above the Rankine cycle thermodynamic working fluid temperature. The efficiency reduction for this case is also almost entirely due to the power required for the compression and liquefaction of the captured CO_2. The efficiency difference between Case-5 and Case-7 is primarily due to higher draft losses within the Boiler Island for Case-5.

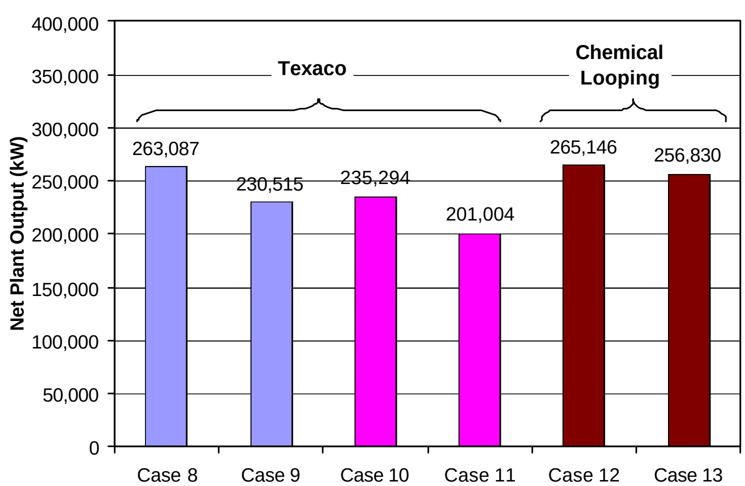
Case-6, which is similar in technical approach to Case-4 except the oxygen is produced from an integrated Oxygen Transport Membrane (OTM) system as opposed to a cryogenic ASU. Case 6 also realized a relatively good thermal efficiency of about 30.0 percent. This is nearly as efficient as Cases 5 and 7. Again, the main efficiency penalty is due to compression and liquefaction of the CO_2.

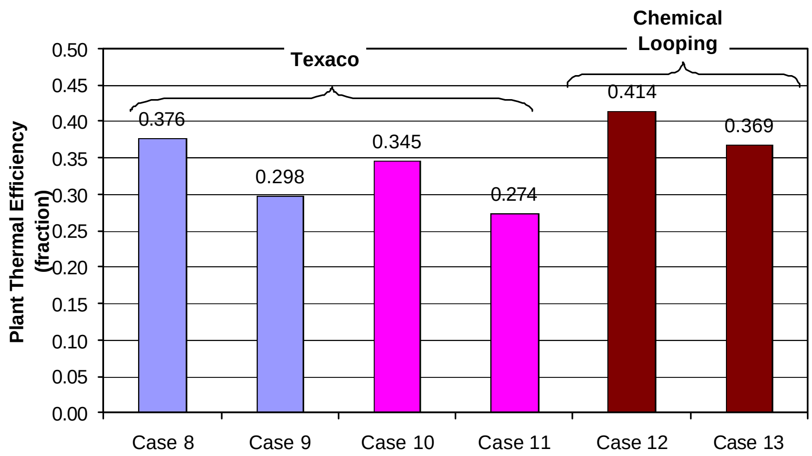
<u>IGCC and Advanced Chemical Looping Cases:</u>
Each of the four IGCC and two advanced Chemical Looping cases were designed with a single train GE-7FA gas turbine, HRSG and 1,800 psig / 1,000 ^oF / 1,000 ^oF steam cycle. The following table and two figures compare the net power outputs and thermal efficiency (HHV basis) among the six cases.

CO_2 capture Cases 9, 11 and 13 all incur significant power output degradation as compared to their Base Case counterparts (Cases 8, 10, and 12), due to the heavy demands of auxiliary power for gas processing which includes CO_2 compression. The efficiency differences among these cases are a reflection of the differences in gasification processes, CO_2 capture processes, and auxiliary power requirements.

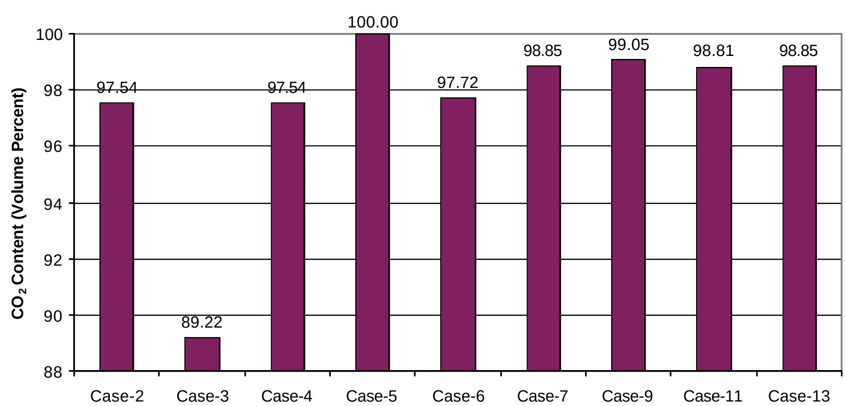
The advanced Chemical Looping cases (Cases 12 and 13) were found to be more efficient both with and without CO_2 capture (36.9 and 41.4 percent HHV, respectively) than the comparable Texaco based IGCC cases. Case-12 was 13 and 23 percent more efficient than Cases 8 and 10 respectively, while Case 13 was 28 and 38 percent more efficient than Cases 9 and 11 respectively.

		Texaco Built and Operating IGCC		Texaco Commercially Offered IGCC		Chemical Looping Gasification	
	(Units)	w/o CO_2 Removal (Case 8)	w/ CO_2 Removal (Case 9)	w/o CO_2 Removal (Case 10)	w/ CO_2 Removal (Case 11)	w/o CO_2 Removal (Case 12)	w/ CO_2 Removal (Case 13)
Power Generator Outputs							
Gas Turbine Power	(kW)	187,150	187,150	187,150	187,150	197,000	197,000
Sweet Gas Expander Power	(kW)	0	0	6,650	6,570	0	0
Steam Turbine Power	(kW)	113,717	112,318	97,924	96,550	104,990	117,379
Gross Plant Power	(kW)	300,867	299,468	291,724	290,270	301,990	314,379
Key Auxiliary Power Listing							
ASU Auxiliaries	(kW)	20,680	22,911	20,080	23,302	0	0
Fuel Compressor	(kW)					29,200	13,080
Oxygen Compressor	(kW)	9,150	10,137	10,570	10,990	0	0
CO_2 Compressor	(kW)	0	27,105	0	25,644	0	35,469
Balance of Auxiliaries	(kW)	7,950	8,800	25,780	29,330	7,644	8,999
Total Auxilary Power	(kW)	37,780	68,953	56,430	89,266	36,844	57,548
Auxiliary Power, % of Gross	(kW)	12.6	23.0	19.3	30.8	12.2	18.3
Net Plant Power	(kW)	263,087	230,515	235,294	201,004	265,146	256,830
Coal Feed Rate	(lbm/hr)	215,454	238,694	210,010	225,822	197,428	213,582
Gasifier Oxygen (95% pure)	(lbm/hr)	183,333	204,167	174,309	187,431	0	0
Thermal Input (HHV)	(kW-thermal)	699,073	774,479	681,410	732,714	640,756	696,012
Net Plant Thermal Efficiency (HHV)	(percent)	37.6	29.8	34.5	27.4	41.4	36.9
Net Plant Heat Rate (HHV)	(Btu/kWhr)	9,069	11,467	9,884	12,441	8,248	9,249





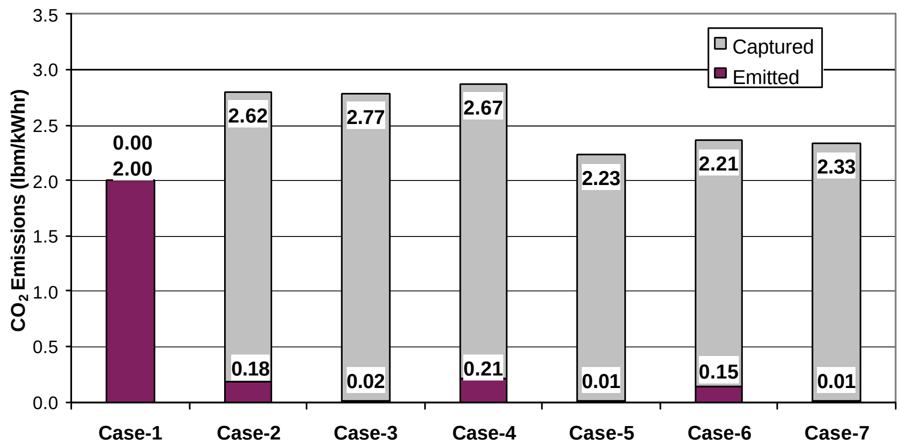
Plant CO_2 Emissions:
Recovery of CO_2 ranged from 90 percent to near 100 percent for these cases. Product purity was greater than 97.5 percent CO_2 by volume for all cases except Case-3, which was about 89.2 percent as shown in the following figure. Case-3 uses a simplified Gas Processing System whereby the product gas stream is not purified, as was done in similar oxygen fired Cases 2, 4, and 6. Therefore, the product gas for Case-3 is suitable for sequestration only.



The CO_2 emissions results for the seven combustion cases are presented first followed by the emissions results for the four IGCC and two advanced Chemical Looping cases.

<u>Combustion Cases:</u>
Specific carbon dioxide emissions for the combustion cases were reduced from about 2.0 lbm/kWh for the Base Case to between 0.01– 0.21 lbm/kWh for the study cases as shown in the following table and figure. Recovery of CO_2 ranged from 93 percent to near 100 percent.

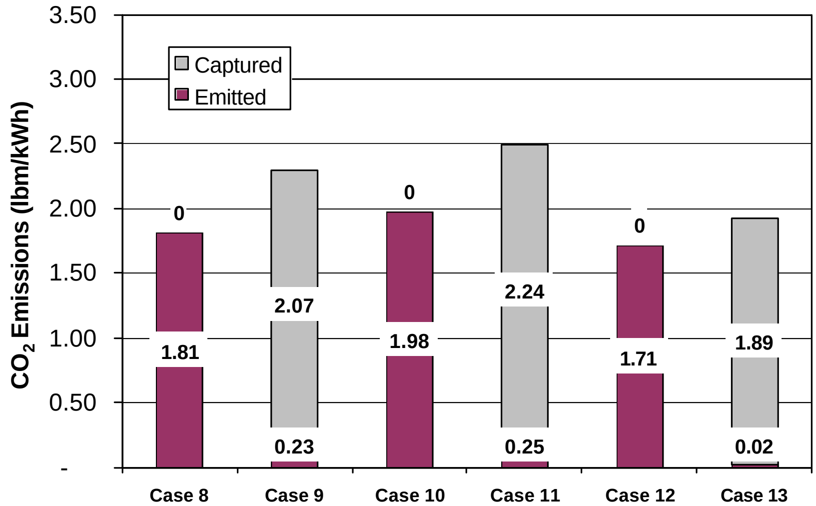
		CFB Air Fired (Case 1)	CFB Cryogenic O_2 Fired (Case 2)	CFB Cryogenic O_2 Fired (Case 3)	CMB Cryogenic O_2 Fired (Case 4)	CMB Air Fired High Temp Carb Proc (Case 5)	CMB with OTM O_2 Fired (Case 6)	CMB Chemical Looping (Case 7)
CO_2 Emissions								
CO_2 Produced	(lbm/hr)	385427	376995	377466	379959	359997	466301	384453
CO_2 Captured	(lbm/hr)	0	352377	375095	352380	359030	437084	383420
Fraction of CO2 Captured	(fraction)	0.00	0.93	0.99	0.93	1.00	0.94	1.00
CO_2 Emitted	(lbm/hr)	385427	24618	2371	27579	967	29217	1033
Specific CO_2 Emissions	(lbm/kwhr)	2.00	0.18	0.02	0.21	0.01	0.15	0.01
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.09	0.01	0.10	0.00	0.07	0.00
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.81	1.98	1.79	1.99	1.85	1.99



IGCC and Advanced Chemical Looping Cases:

The table and figure below compare overall CO_2 emissions on a normalized basis (lbm/kWh) among the four IGCC and two advanced Chemical Looping cases. Specific carbon dioxide emissions for these cases were reduced from about 1.71 - 1.98 lbm/kWh for the Base Cases to between 0.02 - 0.25 lbm/kWh for the study cases as shown in the following table and figure. Recovery of CO_2 ranged from about 90 percent to near 100 percent.

		Texaco Built and Operating IGCC		Texaco Commercially Offered IGCC		Chemical Looping Gasification	
	(Units)	w/o CO_2 Removal (Case 8)	w/ CO_2 Removal (Case 9)	w/o CO_2 Removal (Case 10)	w/ CO_2 Removal (Case 11)	w/o CO_2 Removal (Case 12)	w/ CO_2 Removal (Case 13)
CO_2 Emissions							
CO_2 Produced	(lbm/hr)	477,093	528,791	464,940	500,275	454,321	492,600
CO_2 Captured	(lbm/hr)	0	476,042	0	450,379	0	486,572
CO_2 Fraction Captured	(frac)	0.00	0.90	0.00	0.90	0.00	0.99
Specific CO_2 Captured	(lbm/kWhr)	0.00	2.07	0.00	2.24	0.00	1.89
CO_2 Emitted	(lbm/hr)	477,093	52,749	464,940	49,896	454,321	6,028
Specific CO_2 Emissions	(lbm/kWhr)	1.81	0.23	1.98	0.25	1.71	0.02
Normalized Specific CO_2 Emissions (Relative to Base)	(frac)	1.00	0.13	1.00	0.13	1.00	0.01
Avoided CO_2 Emissions (Compared to Base Case)	(lbm/kWhr)	0.00	1.58	0.00	1.73	0.00	1.69

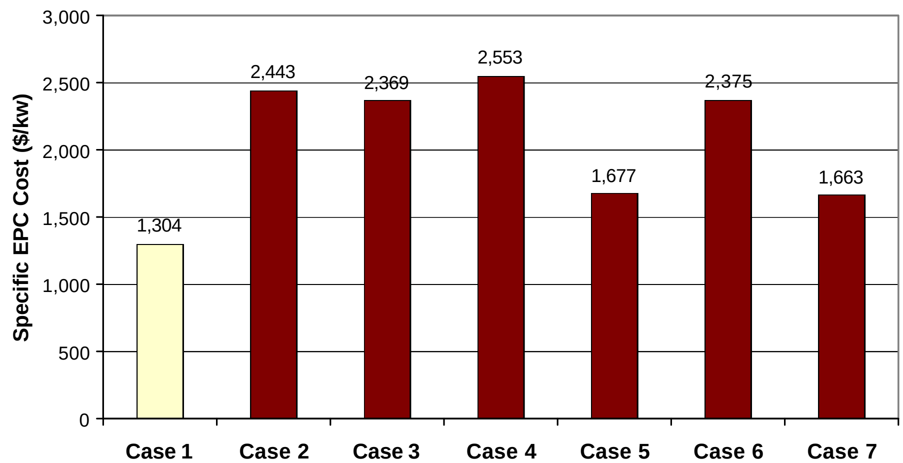


Plant Costs:
The plant investment costs including engineering, procurement, and construction (EPC basis) are presented for the seven combustion cases followed by the cost results for the four IGCC and two advanced Chemical Looping cases. The boundary limit for these estimates includes the complete plant facility within the "fence line". It includes the coal receiving and water supply systems and terminates at the high-voltage side of the main power transformers. Also, for the cases with CO_2 capture, the boundary terminates at the outlet flange of the CO_2 product pipe.

Operating and maintenance cost results are also shown. All costs shown are in July 2003 US dollars for Greenfield plants assumed to be constructed on a common site in the Gulf Coast region of southeastern Texas. All plants were designed for the identical coal analysis, limestone analysis, ambient conditions etc. All construction costs were developed using Gulf Coast non-union labor rates.

<u>Combustion Cases:</u>
The plant investment costs for the combustion cases are shown in the following table and graph. The plant investment cost for the Base Case without CO_2 capture was 1,304 \$/kW. The plant investment cost range for the remaining cases (Cases 2 - 7) with CO_2 capture was from about 1,660 to 2,550 \$/kW. Case 7 (Chemical Looping Combustion) was found to be the lowest cost of the combustion based capture cases (1,663 \$/kW) followed closely by Case-5, the Regenerative Carbonate Process, at 1,677 \$/kW. Cases 2, 3, and 4, all variants of the cryogenic based oxygen fired process, were found to have significantly higher costs (2,370 – 2,550 \$/kW). Case-3, which used a simplified Gas Processing System (drying and compression only), showed a savings of about 74 \$/kW or about 3 percent as compared to Case-2. Case 6 (oxygen fired via an advanced OTM system) was slightly less costly than the comparable cryogenic case at about 2,375 \$/kW, a savings of about 7 percent as compared to Case 4.

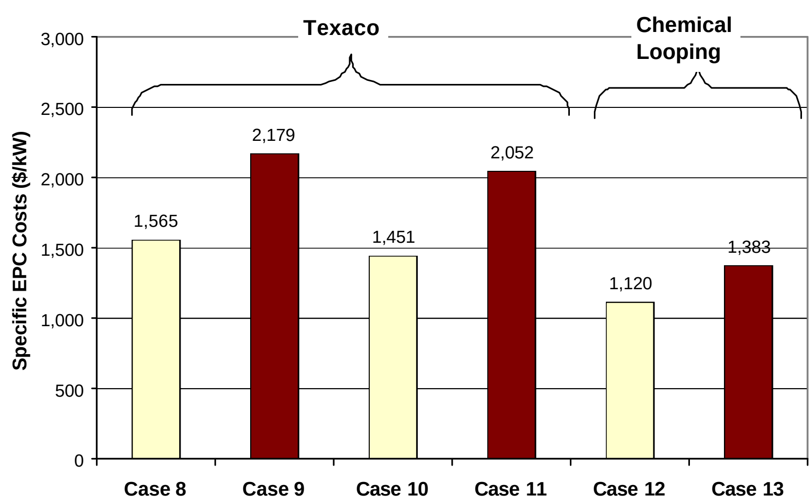
Study Case	Net Plant Output, kW	Total Investment Cost, EPC Basis	
		$x1000	$/kW
Case 1, Air-fired CFB w/o CO_2 Capture	193,037	251,804	1,304
Case 2, O_2-Fired CFB w/ASU & CO_2 Capture	134,514	328,589	2,443
Case 3, O_2-Fired CFB w/ASU & Flue Gas Sequestration	135,351	320,638	2,369
Case 4, O_2-Fired CMB w/ASU & CO_2 Capture	132,168	337,402	2,553
Case 5, Air-Fired CFB w/Carbonate Reg. Process & CO_2 Capture	161,184	270,232	1,677
Case 6, O_2-Fired CMB w/OTM & CO_2 Capture	197,435	468,919	2,375
Case 7, CMB Chemical Looping Combustion w/CO_2 Capture	164,484	273,568	1,663



Operating and maintenance (O&M) costs were calculated for all systems for the combustion cases. Total O&M costs for the capture cases ranged from about 1.2 to 1.8 cents/kWh while the Base Case was about 0.8 cents/kWh.

IGCC and Advanced Chemical Looping Cases:
The plant investment costs for the four (4) IGCC and two (2) advanced Chemical Looping cases are shown in the following table and graph. The plant investment cost (EPC basis) for the Texaco Base Cases (Cases 8 and 10) without CO_2 capture was 1,565 and 1,451 \$/kW. The plant investment costs for the corresponding cases (Cases 9 and 11) with CO_2 capture was 2,179 to 2,052 \$/kW respectively. Case 13 (advanced Chemical Looping) was found to be the lowest cost of the capture cases (1,383 \$/kW) as compared to Case 12 without CO_2 capture at 1,120 \$/kW.

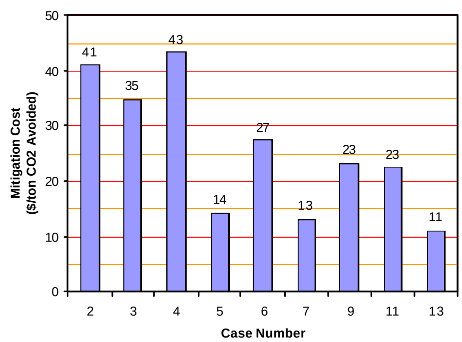
Study Case	Net Plant Output, kW	Total Investment Cost, EPC Basis	
		$x1000	$/kW
Case 8, Built & Operating IGCC w/o CO_2 Capture	263,087	411,731	1,565
Case 9, Built & Operating IGCC w/ CO_2 Capture	230,515	502,330	2,179
Case 10, Commercially Offered IGCC w/o CO_2 Capture	235,294	341,468	1,451
Case 11, Commercially Offered IGCC w/CO_2 Capture	201,004	412,377	2,052
Case 12, Chemical Looping Gasification w/o CO_2 Capture	265,146	296,991	1,120
Case 13, Chemical Looping Gasification w/ CO_2 Capture	256,830	355,132	1,383

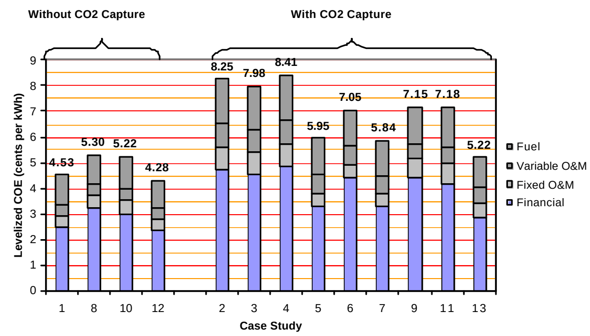


Total O&M costs for the Base Cases ranged from about 0.7 to 1.0 cents/kWh while the CO_2 capture cases ranged from about 1.0 to 1.4 cents/kWh.

Economic Evaluations:
The following two figures summarize the economic results for all thirteen cases in this study. The figure on the left shows CO_2 mitigation costs (\$/Ton of CO_2 avoided) and the figure on the right shows levelized cost of electricity for all cases with CO_2 capture cases shown on the right and cases without CO_2 capture on the left.

For cases with CO_2 capture, Case 13, advanced Chemical Looping, represents the best of the cases studied based on both levelized COE and CO_2 mitigation cost evaluation criteria. Case 7, Chemical Looping combustion, and Case 5, the regenerative carbonate process, were about 12 and 14 percent higher than Case 13 with respect to levelized COE. These three cases showed significant COE advantages as compared to all other capture cases in this study.



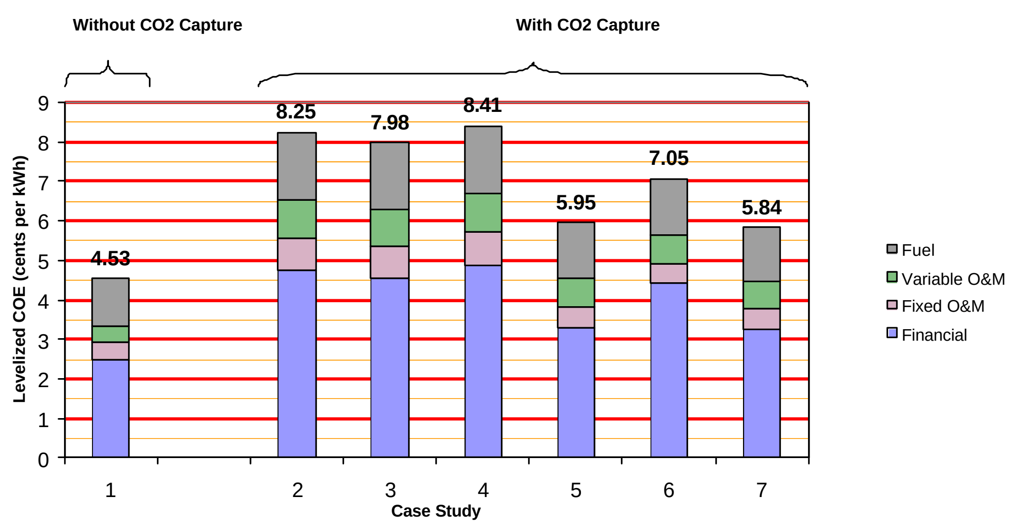


The following sections compare the economic results (levelized cost of electricity and CO_2 mitigation cost) in two groups. The results for the seven (7) combustion cases are presented first followed by the economic results for the four (4) IGCC and two (2) advanced Chemical Looping cases shown in the second group.

Combustion Cases:

The following table and figure summarize the economic analysis results for the seven combustion cases.

Study Case	Levelized Cost of Electricity (c/kWh)					Incremental COE (c/kWh)
	Capital	Fixed O&M	Variable O&M	Fuel	Total	
Without CO_2 Capture						
Case 1, Air-fired CFB w/o CO_2 Capture	2.49	0.42	0.41	1.20	4.53	0.00
With CO_2 Capture						
Case 2, O_2-Fired CFB w/ASU & CO_2 Capture	4.73	0.85	0.95	1.72	8.25	3.72
Case 3, O_2-Fired CFB w/ASU & Flue Gas Sequestration	4.53	0.85	0.91	1.69	7.98	3.45
Case 4, O_2-Fired CMB w/ASU & CO_2 Capture	4.86	0.85	0.96	1.74	8.41	3.88
Case 5, Air-Fired CFB w/Carbonate Reg. Process & CO_2 Capture	3.29	0.51	0.73	1.41	5.95	1.42
Case 6, O_2-Fired CMB w/OTM & CO_2 Capture	5.06	0.47	0.73	1.42	7.69	3.16
Case 7, CMB Chemical Looping Combustion w/CO_2 Capture	3.26	0.50	0.70	1.38	5.84	1.32



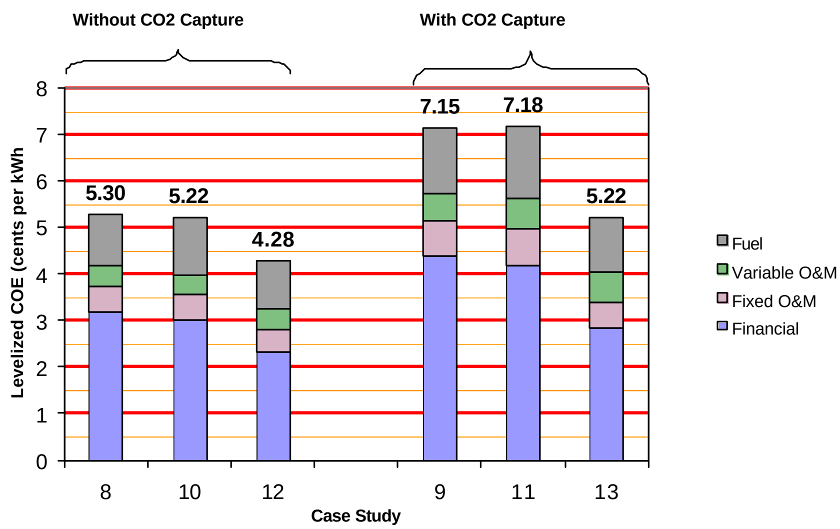
Case-7 (Chemical Looping Combustion) was found to be the best alternative of the six combustion based capture concepts studied based on levelized Cost of Electricity (COE) evaluation criteria. Case-5 (High Temperature Carbonate Regeneration) is only slightly worse (about 2 percent) than Case-7. Case-7 was found to have an incremental COE value of 1.32 cents/kWh as compared to Case-1 (about a 29 percent increase) and a CO_2 mitigation cost of 13 $/Ton avoided. COE values for the cryogenic based cases (Cases 2, 3, and 4) were significantly higher than Case-7 (by about 40 percent). Case-6, which used an OTM for oxygen production, was in between, about 20 percent higher than Case-7 and about 20 percent better than Cases 2, 3 and 4.

Case-3, which uses a simplified Gas Processing System (no purification) and produces a product suitable for sequestration only, showed a relatively insignificant improvement in COE of about 3 percent as compared to Case-2 where product purification was used.

IGCC and Advanced Chemical Looping Cases:

The following table and figures summarize the economic analysis results for the four (4) IGCC and two (2) advanced Chemical Looping cases.

Study Case	Levelized Cost of Electricity (c/kWh)					Incremental COE (c/kWh)
	Financial	Fixed O&M	Variable O&M	Fuel	Total	
Without CO2 Capture						
Case 8, Built & Operating IGCC w/o CO2 Capture	3.20	0.55	0.42	1.13	5.30	
Case 10, Commercially Offered IGCC w/o CO2 Capture	3.00	0.57	0.42	1.24	5.22	
Case 12, Chemical Lopping Gasification w/o CO2 Capture	2.34	0.47	0.44	1.03	4.28	
With CO2 Capture						
Case 9, Built & Operating IGCC w/ CO2 Capture	4.40	0.75	0.57	1.43	7.15	1.85
Case 11, Commercially Offered IGCC w/CO2 Capture	4.19	0.79	0.65	1.56	7.18	1.95
Case 13, Chemical Lopping Gasification w/ CO2 Capture	2.85	0.55	0.66	1.16	5.22	0.93



Case-13 (advanced Chemical Looping with CO_2 capture) was found to be clearly the best alternative of the three gasification CO_2 capture concepts studied based on levelized COE evaluation criteria (5.22 cents/kWh). This case was found to be about 27 percent better than the Texaco based IGCC cases. This case was found to have an incremental COE value of 0.93 cents/kWh as compared to Case 12 (advanced Chemical Looping without CO_2 capture) and a mitigation cost of about 11 $/Ton of CO_2 avoided.

Case-13 was also found to be about 11 percent better, with respect to COE, than the best of the combustion cases (Case-7; Chemical Looping Combustion; 5.84 cents/kWh).

Task 2 Results

The results for Task 2, which are described in detail in the Volume II report, are summarized below. These results were used to support the Task 1 effort.

Tests were carried out in the Thermo-Gravimetric Analysis (TGA) apparatus on the Base Case coal in environments simulating combustion in air, in 30 percent O_2 / 70 percent CO_2 medium. Clearly, these results indicate that coal combustion in this medium does not adversely affect the combustion process kinetics.

The three subject fuels are being evaluated in ALSTOM's 4-inch FBC test facility. The Base Case coal has been tested in air and O_2 / CO_2 mixtures containing from 30 percent to 70 percent O_2 (in CO_2 balance). The other two fuels (Illinois #6 coal and delayed petroleum coke) were tested in air and limited O_2/CO_2 mixtures. Additionally, all three fuels were tested in air and 30 percent O_2 / 70 percent CO_2 in the presence of a Ca/S mole ratio of 3.5. Results can be summarized as follows:

- Testing the Base Case CFB coal in O_2/CO_2 mediums containing up to 70 percent O_2, for example, caused bed temperature rises of up to about 250 ^oF. Nevertheless, it was possible to avoid bed slagging/de-fluidization problems as long as the bed was well fluidized.

- The added emission benefits offered by oxy-fuel firing over air firing in circulating Fluidized Bed Boilers (CFB's) are:
  - CO_2 in the flue gas is highly concentrated (~90 percent vs.~15 percent), thus making the processing of this stream to achieve the required CO_2 purity for EOR application relatively cheaper.
  - Typically low NOx emissions in combustion-staged air-fired CFB's are further reduced primarily due to elimination of thermal NOx.
  - SO_2 emissions reductions of up to 90 percent with sorbent utilization should not to be negatively impacted. Furthermore, ALSTOM has a commercial product called "Flash Drier Absorbent (FDA)," which has been successfully demonstrated in the pilot-scale Multi-use Test Facility (MTF) to reduce SO2 emissions by as much as 99 percent.

- Burning the three fuels in high O_2 combustion mediums improved overall fuel combustion efficiencies, which, correspondingly, improved carbon loss.

- The addition of limestone to the combustion process to control sulfur dioxide emission did not adversely impact the overall combustion efficiency of each fuel.

- The test conditions used in the FBC facility are much more aggressive than those encountered in commercial CFB's (e.g. furnace outlet O_2 concentrations: 13-51 percent vs. ~3 percent; superficial gas velocity: ~2-3 ft/sec. vs. ~18 ft/sec.). Hence, it is preliminarily concluded that the choice of 70 percent O_2/30 percent recycled flue gas (i.e., ~CO_2) as a combustion medium for study Case 2 (New Compact O_2-Fired FCB, see Section 2.2, Volume I) was feasible.

Technical Conclusions

In a region without carbon constraints, whereby only power generation is considered, the following techno-economic conclusions can be drawn based on the results from this study:

- With respect to efficiency:
  - The Texaco IGCC power plant technology is more efficient than the air-fired CFB technology with subcritical steam conditions (37.6 vs. 35.5 percent, HHV). This is due, principally, to the fact that the IGCC technology takes advantage of both elevated pressure operation and combined cycle principles.
  - The advanced Chemical Looping concept (Case 12), being developed by ALSTOM, is more efficient than the Texaco IGCC (41.4 vs. 37.6 percent, HHV). This is due to several factors the most important being that the advanced Chemical Looping takes advantage of separation of oxygen from air via the chemical looping process rather than production of oxygen from a relatively energy intensive cryogenic air separation unit.

  - The Texaco IGCC efficiency values reported in this study are lower than values reported by Parsons (Holt, 2000) primarily because Parsons used H-Class gas turbines whereas this study used F-Class gas turbines. Additionally, Parsons used ambient conditions of (63 ^{o}F dry bulb and 14.4 psia), whereas ALSTOM used ABMA ambient conditions (80 ^{o}F dry bulb and 14.7 psia) in this study. Also, Parsons and ALSTOM used condenser pressures of 2.0, and 3.0 in. Hga, respectively.

- With respect to investment costs and levelized costs of electricity (COE):
  - The Texaco IGCC power plant technology is about 20 percent more capital-intensive than the air-fired CFB technology (> 250 $/kW). This is due to the fact that an IGCC power plant syngas cleanup system is very complex, comprised of many specialized components operating at elevated pressures.
  - The air-fired CFB plant produces electricity about 15 percent cheaper in terms of COE than the Texaco IGCC plant (4.5 vs. 5.3 cents/kWh)
  - The advanced Chemical Looping plant concept shows the potential to provide electricity at a lower cost than both air-fired CFB and IGCC plants (4.3 vs 4.5 & 5.3 cents/kWh). It is noted that the advanced Chemical Looping plant concept is in a very early stage of development. Hence, the investment costs and COE estimates for this concept must be considered preliminary.

In a carbon-constrained region, whereby both power generation and carbon capture are considered, the following techno-economic conclusions can be drawn based on the results from this study:

- With respect to efficiency:
  - All options reduce power plant efficiencies compared to respective baseline plants without CO_2 capture, as discussed below.
  - The advanced Chemical Looping plant concept is the most efficient of all cases studied herein (36.9 percent vs. 24.6-30.9 percent, HHV basis).
  - The efficiencies of the advanced combustion cases (CMB w/OTM, CFB w/Carbonate Regeneration Process, and Chemical Looping Combustion [CLC] CMB) fall in the range that is marginally higher than that of the Texaco IGCC plant (30-31 vs. 29.8 percent, HHV).
  - The Texaco IGCC power plant technology is more efficient than cryogenic oxygen-fired CFB/CMB power plant technology (29.8 vs. 24.6-25.3 percent, HHV). These results are equivalent to energy penalties, compared to their

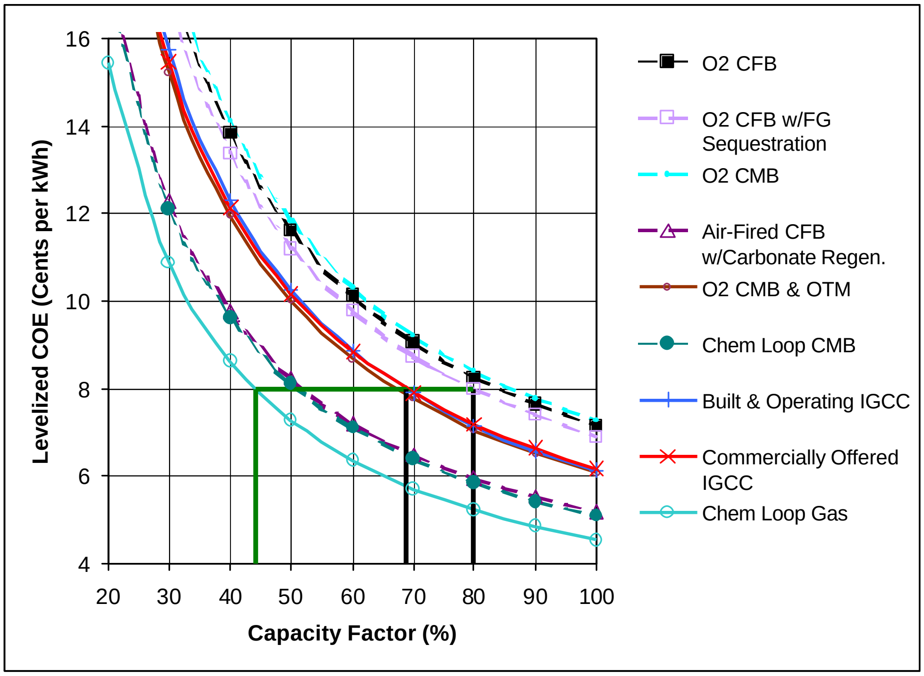
respective reference plants, of 21 percent for the IGCC and 28-31 percent for the cryogenic CFB plants. This IGCC efficiency advantage is, principally, due to the fact that the CO_2 product is compressed from 50 psig to 2,000 psig for the IGCC and from atmospheric pressure to 2,000 psig for the CFB's. Also, the CFB plants require more oxygen per unit of coal fired than the Texaco IGCC plants, resulting in additional ASU power requirements for the cryogenic CFB plants.
  - The use of oxy-fuel firing to produce a "for Sequestration only" flue gas yields only marginal benefit from a plant thermal efficiency standpoint (25.3 vs. 25.2 percent, HHV). This is due to the fact that the compression step, which is the most energy-intensive in flue gas processing, cannot be avoided. However, this plant provides the only true "zero gaseous emissions" plant in this study.

- With respect to investment costs and levelized costs of electricity (COE):
  - The advanced Chemical Looping plant concept is the least costly of all the concepts considered: its EPC (engineered, procured, and constructed) capital cost and levelized cost of electricity are 1,380 $/kW and 5.2 cents/kWh, respectively. This cost of electricity for advanced Chemical Looping is nearly equivalent to new coal-fired plants offered today without CO_2 capture (i.e., Case 1 @ 4.5 cents/kWh).
  - The Carbonate Regeneration Process and Chemical Looping Combustion CMB are less capital intensive than the built and operating or commercially offered IGCC's (1,680-1,660 vs. 2,180-2,050 $/kW). Hence, their levelized costs of electricity are correspondingly lower (5.9 - 5.8 vs. 7.2 cents/kWh).
  - The cryogenic oxygen-fired CFB or CMB plants are more capital intensive ($/kW) than the built and operating or commercially offered IGCC's, because they incur high energy penalties thus reducing plant output and increasing specific costs ($/kW); by comparison (2,350-2,550 vs. 2,180-2,050 $/kW). Consequently, their levelized costs of electricity are, correspondingly, higher (8.3 - 8.4 vs. 7.2 cents/kWh).
  - The capital investment of an oxy-fuel fired plant designed to produce "for Sequestration only flue gas" is approximately 3 percent lower than that of an oxy-fuel fired CFB designed to produce a CO_2 product suitable for EOR application (2,370 vs. 2,440 $/kW). Consequently, its levelized cost of electricity is, correspondingly, about 3 percent lower (8.0 vs. 8.3 cents/kWh).

The figure below is a plot of COE vs. Capacity Factor for all the technologies evaluated. This plot was obtained by keeping all the economic assumptions given in Table 4.1.1 at their "Base" values, and varying only the Capacity Factor. Overall, these results indicate the following:

- One of the lessons learned from this figure is that with CO_2 capture, the cycle advantages of IGCC over oxygen fired combustion systems may overcome its disadvantages in terms of capital cost and availability found in a no CO_2 capture comparison and may well provide marketplace incentive to move this technology into the mainstream market.

- While IGCC with CO_2 capture offers potential advantages over cryogenic oxygen based combustion systems with CO_2 capture, the advantages of chemical looping over IGCC is even greater. Taken as a group, a possible road map emerges which shows the oxygen based combustion system as a short-term solution, and a strong economic incentive for the development of Chemical Looping. Each step has significant advantages over the prior step.



- The chart above demonstrates the importance of availability to an operating plant. In the example shown, an O_2-fired CFB with flue gas sequestration and 80 percent availability has the same COE as an IGCC plant with CO_2 capture and 70 percent availability. Industry practice for CFB's has been to have one planned two-week outage every twelve to eighteen months, Current requirements for IGCC plants suggest two planed three-week outages per year. The difference in planned outage times equalizes the COE of these two cases.

- Another observation is that an advanced Chemical Looping Plant would have the same COE at 43 percent availability. Such a figure provides considerable latitude in the initial stages of product introduction, where most start-up problems occur. This reduces the commercialization risk of the advanced Chemical Looping technology.

Recommendations for Future Work

It is recommended that the Department of Energy's National Energy Technology Laboratory pursue a strategy that supports technologies with short-range and long-range commercialization potentials that include the following.

- To continue the development of the circulating moving bed (CMB), O_2-fired CFB, and Chemical Looping Combustion and advanced Chemical Looping.

- <u>Short-Range Technology:</u> Oxy-fuel Fired CFB technology. Given that it would use a combination of already available enabling technologies (cryogenic O_2 production and gas processing system), it could be deployed within a five-year time horizon. The moving bed portion of the CMB technology is also utilized in the O_2-fired CFB. As shown in Section 6, this technology would be suitable for Enhanced Oil Recovery (EOR) application, with the Bakersfield Project in California being a potential first application site. This technology also would be applicable for Enhanced Coal Bed Methane (ECBM). An additional advantage of this technology is that efficiency is maintained over a wide plant size range (50 MWe and larger).

The DOE's authorization of the present project's continuation, Phase II pilot-scale testing of two coals and one delayed petroleum coke followed by a refinement of the oxy-fuel plant's design, performance and economic analyses, represents a first step toward this development. This technology is the only one capable of "zero gaseous emissions" at the present time.

- Long-Range Technology: Advanced Chemical Looping is a technology showing such promise that ALSTOM has already begun the design of a small "Proof of Concept" pilot-scale facility. Additionally, ALSTOM has responded to a DOE NETL RFP to conduct an extensive test program in this facility (DE-PS26-02NT41613-01). The Circulating Moving Bed (CMBTM) technology, being developed by ALSTOM with DOE NETL's financial support (Jukkola, et al., 2003), is short-range, enabling technology that represents a stepping-stone towards the development of the Chemical Looping technology.

1. INTRODUCTION

The greenhouse effect is created by the presence of a number of gases in the atmosphere, with CO_2 being the single largest contributor, and accounting for about 50 percent of the greenhouse phenomenon. Large quantities of CO_2 are produced from fossil fuel combustion. Previous studies (Bozzuto, et al., 2001) have shown that CO_2 capture from existing coal fired plants utilizing commercial amine based flue gas scrubbing systems would reduce plant output and efficiency by up to one-third and increase cost of electricity by about 5 cents/kWh. This project embodies technical innovation for capturing CO_2 in concert with strategies that enable by-products to become saleable product streams, which would significantly and favorably impact the plant thermal efficiency and/or cost of electricity and CO_2 capture.

One of the methods identified for CO_2 capture is to burn fossil fuels in a mixture of oxygen and recycled flue gas. Employment of this concept eliminates the presence of almost all atmospheric nitrogen in the flue gas, thereby resulting in a flue gas that is composed primarily of CO_2, along with small quantities of moisture, oxygen, nitrogen, and trace gases like SO_2 and NO_x. The combination of recycled flue gas/oxygen mixtures in concert with combustion in a circulating fluidized bed (CFB) offers unique advantages compared to alternative methods of firing fossil fuels with oxygen. Unlike pulverized coal (PC) combustion or stoker firing, fluidized bed combustion has the advantage of controlling combustion chamber temperatures by virtue of modulating the recycle rate of cooled solids. This unique feature of a fluidized bed combustor means that much higher percentages of oxygen can be used in the recycled flue gas/oxygen mixtures than would be possible in alternate firing applications. Though the primary motivation for using oxygen is to facilitate CO_2 capture, newly constructed CFB combustors will be able to capitalize on the use of high oxygen content firing. Specifically, and importantly, the use of higher oxygen contents will improve overall system thermal efficiency, and allow the design and construction of more compact, relatively less expensive CFB boilers. Economic analysis indicates that the proposed oxygen-firing technology in circulating fluidized boilers could be developed and deployed economically in the near future in an enhanced oil recovery (EOR) application.

Additionally, ALSTOM Power Inc. (ALSTOM) has identified several advanced/novel plant configurations, which further improve the efficiency and cost of the CO_2 product cleanup and compression with respect to their commercialization potentials. These advanced/novel concepts require long term development efforts. These concepts are both combustion-based and gasification-based power generation systems.

ALSTOM teamed with Parsons Energy and Chemical Group Inc., ABB Lummus Global Inc., Praxair Inc., and the US DOE NETL, to conduct a comprehensive study evaluating the technical and economic feasibility of alternate CO_2 capture technologies applied to Greenfield US coal-fired electric generation power plants. Thirteen (13) separate but related cases, representing various levels of technology development, were evaluated in a directly comparable manner as described below. The first seven cases represent coal combustion cases in CFB type equipment. The next four cases are IGCC's based on Texaco gasification processes. The final two cases are based on an advanced Chemical Looping gasification process being developed and fully paid for by ALSTOM. One CFB case, two IGCC cases and one advanced Chemical Looping gasification case were defined as "Base Case" systems without CO_2 capture for comparison purposes.

These thirteen cases are grouped and briefly described as follows:

Combustion Cases

- Case-1: Air Fired Circulating Fluidized Bed (CFB) without CO_2 Capture (Base Case for Comparison to Cases 2-7)
- Case-2: Oxygen Fired CFB with CO_2 Capture
- Case-3: Oxygen Fired CFB with CO_2 Capture (sequestration only option)
- Case-4: Oxygen Fired Circulating Moving Bed (CMB) with CO_2 Capture (advanced boiler concept)
- Case-5: Air Fired CMB with CO_2 Capture utilizing Regenerative Carbonate Process
- Case-6: Oxygen Fired CMB with Oxygen Transport Membrane (OTM) and CO_2 Capture
- Case-7: Indirect Combustion of Coal via Chemical Looping and CO_2 Capture

IGCC Cases

- Case-8: Built and Operating Present Day IGCC without CO_2 Capture (Base Case for Comparison with Case-9)
- Case-9: Built and Operating Present Day IGCC with shift reaction and CO_2 Capture
- Case-10: Commercially Offered Future IGCC without CO_2 Capture (Base Case for Comparison with Case-11)
- Case-11: Commercially Offered Future IGCC with shift reaction and CO_2 Capture

Advanced Chemical Looping Cases

- Case-12: Indirect Gasification of Coal via Chemical Looping (Base Case for comparison to Case-13)
- Case-13: Indirect Gasification of Coal and CO_2 Capture via Chemical Looping

Each of these CO_2 capture technologies was evaluated against the representative Base Case from the standpoints of performance, CO_2 emissions, impacts on power generation cost and CO_2 mitigation cost. The Base Cases represent the "business as usual" operation scenario for the plants without CO_2 capture. All of the Combustion, IGCC and advanced Chemical Looping plants in this evaluation represent domestic utility-scale power plants and all technical performance and cost results associated with these options were evaluated in a directly comparable manner.

Cost estimates were developed for all the alternative power plants which included systems required to produce, extract, clean, compress and liquefy the CO_2, which could then be available for use in enhanced oil or gas recovery or for sequestration. The boundary limit for these estimates includes the complete plant facility within the "fence line". It includes the coal receiving and water supply systems and terminates at the high-voltage side of the main power transformers. Additionally, the impact of CO_2 capture on the cost of electricity (COE) and on the mitigation cost for CO_2 ($/ton of CO_2 avoided) was also evaluated including sensitivity studies.

ALSTOM Power Inc. managed and performed the subject study from its US Power Plant Laboratories offices in Windsor, Connecticut. Participating as sub-contractors in this effort were ABB Lummus Global, from its offices in Houston, Texas, Parsons Energy and Chemical Group, from its offices in Wyomissing, Pennsylvania, and Praxair Inc. from its offices in Tonawanda, New York. Plasma Inc. of Butte, Montana served as an informal consultant. The US Department of Energy (US DOE) National Energy Technology Laboratory (NETL) provided consultation and funding. ALSTOM provided cost share to this project. Additionally, as stated above, ALSTOM paid for, in full, the costs associated with the analyses of advanced Chemical Looping Cases 12 and 13.

Workscope assignments for all team members are briefly described below:

For the combustion cases (1-7), ALSTOM developed the overall plant concept and overall thermal performance for the power plant including material and energy balances. Additionally, ALSTOM developed conceptual Boiler Island designs, specifications for balance of plant (BOP) systems and equipment, specifications for the gas processing systems (GPS), and specifications for the oxygen production equipment. Using the ALSTOM system and/or equipment specifications, Parsons estimated BOP equipment costs, Lummus developed GPS performance, designs and costs, and Praxair developed ASU or OTM system performance, designs and costs.

For the four IGCC cases (8-11), Parsons developed overall plant thermal performance and costs based on overall plant specifications provided by ALSTOM, and Lummus developed the CO_2 compression / liquefaction system performance, design and costs.

For the advanced Chemical Looping cases (12, 13) both ALSTOM and Parsons developed the overall plant thermal performance with ALSTOM providing performance for the advanced Chemical Looping processes and Parsons developing performance for the BOP (fuel compression, combustion turbine, HRSG, steam cycle). ALSTOM developed conceptual designs and estimated costs for the advanced Chemical Looping systems while Parsons estimated costs for the BOP. Lummus information was used for development of CO_2 compression / liquefaction system performance, design and costs.

ALSTOM then developed the economic analyses for all thirteen cases including economic sensitivity studies.

ALSTOM Power is a well-established global leader in the design and manufacture of power generation equipment. ABB Lummus Global is a leader in petrochemical technology and CO_2 separation technology. Parsons is a recognized leader in power plant design and engineering. Praxair is one of the leaders in the design and construction of Air Separation Units and is also a recognized leader in the development of Oxygen Transport Membrane technology.

The key goals of the study were to evaluate the impacts on the plant output, efficiency, CO_2 emissions, investment costs, cost of electricity and CO_2 mitigation costs, resulting from the addition of the CO_2 capture systems to newly constructed coal power plants. An objective of the proposed project was to determine if carbon can be recovered at an avoided cost of $10/ton (or less), using existing or newly constructed, novel CFB combustor designs while burning coal or petroleum coke in oxygen with minimal recycled flue gas, instead of air.

A near term commercial scenario is also evaluated herein (Section 6) where coal or petroleum coke would be burned and carbon dioxide would be captured and sent to an oil field for EOR. This case was evaluated with the help of the input gathered from potential commercial users of CO_2 for EOR (AERA Energy LLC, a Limited Liability Corporation for oil production in California for Exxon/Mobil and Shell Oil companies). This input, received through Plasma Inc. (an informal consultant to this project), was used to help establish the scale of the combustor to be evaluated in this project and other input for the commercial scenario. These contacts facilitated the process of acquiring necessary commercially related information needed during the execution of this project.

2. CASE STUDIES: DESIGN BASIS, PROCESS DESCRIPTIONS, EQUIPMENT, AND PLANT PERFORMANCE

A total of thirteen (13) Greenfield case studies, listed below, were analyzed in this evaluation. The thirteen cases were subdivided into three groups. Seven of the cases were grouped as coal Combustion based cases, four were IGCC cases and the remaining two were advanced Chemical Looping cases. The four IGCC cases are based on Texaco gasification processes. The final two cases are based on an advanced Chemical Looping gasification process being developed by ALSTOM. One Combustion case, two IGCC cases and one advanced Chemical Looping case were defined as "Base Case" systems without CO_2 capture for comparison purposes. These cases without CO_2 capture represent Base Cases for comparison with the remaining CO_2 capture cases such that the impact of CO_2 capture is fully understood and accurately quantified. Within each group, the numerical order of the various cases roughly represents increasing levels of technology development (i.e., within the Combustion group, Case-7 would require the most development and Case-2 the least).

This section of the report describes the design basis and the various processes used for each of the cases analyzed. The equipment used for these processes is also described. Additionally, the performance of each case is also presented in terms of the associated energy and material balances as well as various plant performance summary tables.

Brief descriptions of each case are presented below with more detailed case by case descriptions provided later in Section 2.

<u>Combustion Cases</u>

- **Case-1**: Air Fired CFB without CO_2 Capture (Base Case for Comparison)
  Conventional air-fired CFB without CO_2 capture using 1,800 psig / 1,000 ^oF / 1,000 ^oF, 3.0 in. Hga steam cycle.
  <u>Implication:</u> Provides reference point for performance & economic analyses to the CO_2 capture Cases 2-7

- **Case-2**: New Compact Oxygen-Fired CFB with CO_2 Capture, Purification, Compression and Liquefaction.
  Same steam cycle as Case-1 and nearly identical thermal input but significantly smaller boiler island equipment than Case-1. Oxygen is from a Cryogenic Air Seperation Unit (ASU) Plant. CFB Boiler Island provides a concentrated CO_2 flue gas product stream to the Gas Processing System.
  <u>Implication:</u> Cost savings for the Boiler Island. Cost savings for the Gas Processing System equipment as compared to amine scrubbing systems. Improved plant thermal efficiency as compared to amine based CO_2 capture systems.

- **Case-3**: Oxygen-Fired CFB with CO_2 Capture (sequestration only option).
  Same as Case-2, but uses a simplified Gas Processing System whereby the product gas stream is not purified and therefore is suitable for sequestration only.
  <u>Implication:</u> Further cost savings (Gas Processing System) as compared to Case-2.

- **Case-4**: Oxygen-Fired Circulating Moving Bed (CMB) with CO_2 Capture, Purification, Compression and Liquefaction.
  Same as Case-2, but uses advanced boiler design concepts.
  Implication: Anticipated cost savings (on Boiler Island Equipment) as compared to Case-2.

- **Case-5**: Air-Fired Boiler with CMB and CO_2 Capture Utilizing High Temperature Regenerative Carbonate Process.
  Utilizes air-firing and carbonate regeneration at higher temperatures than steam cycle temperatures. Thus, all the energy rejected from the carbonate regeneration process is recovered in the steam cycle at high temperature such that there is no efficiency penalty associated with CO_2 capture for this process. Nearly pure CO_2 is removed continuously from a calciner within the Boiler Island. This stream is regenerativly cooled and is then ready for compression and liquefaction.
  Implication: Advanced novel boiler and CO_2 capture concept eliminates the high power requirement of the cryogenic ASU used in Cases 2, 3, and 4 and eliminates the energy penalty typically associated with CO_2 capture for significant plant thermal efficiency improvement. Significant Boiler Island cost savings are also anticipated for this case.

- **Case-6**: The Case-4 CMB process, Integrated with Oxygen Transport Membrane (OTM) instead of cryogenic ASU.
  Utilization of an OTM is a more efficient method for O_2 production as compared to a conventional cryogenic ASU as was used with Cases 2, 3, and 4. The CMB process also has a 2,000 ^oF solids stream available for high temperature air heating as required by the OTM.
  Implication: Significant plant thermal efficiency improvement as compared to Cases 2, 3, and 4.

- **Case-7**: Indirect Combustion of Coal via Chemical Looping.
  This air-fired boiler utilizes a continuously looping solid oxygen-carrier which oxidizes the fuel into primarily CO_2 and H_2O. Simple condensation of the H_2O then yields a fairly pure CO_2 product stream for compression and liquefaction.
  Implication: Advanced novel boiler concept eliminates the high power requirement of the cryogenic ASU used in Cases 2, 3, and 4 and eliminates the energy penalty typically associated with CO_2 capture for significant plant thermal efficiency improvement. Significant Boiler Island cost savings are also anticipated for this case.

IGCC Cases

- **Case-8**: Built and Operating Present Day IGCC plant without CO_2 Capture (Base Case for Comparison).
  IGCC without CO_2 capture utilizing Texaco pressurized (i.e., 450 psig), oxygen-blown, entrained flow gasification technology (based on Tampa IGCC design), and a single train GE-7FA gas turbine with HRSG and 1,800 psig / 1,000 ^oF / 1,000 ^oF, 3.0 in. Hga steam cycle.
  Implication: Provides a reference point for the performance & economic analyses of Case 9.

- **Case-9**: Built and Operating Present Day IGCC plant with shift reaction and CO_2 Capture.
  Same as Case-8 but with water-gas shift reactor and CO_2 capture, compression and liquefaction system.
  Implication: Provides direct comparison with Case-8 to isolate the impact of CO_2 capture for present day IGCC plants.

- **Case-10**: Commercially Offered Future IGCC plant without CO_2 Capture (Base Case for Comparison with Case-11).
  IGCC without CO_2 capture utilizing Texaco high pressure(i.e., 950 psig), oxygen-blown, entrained flow, quench gasification technology with syngas expander (based on Eastman Chemical Company Acetic Anhydride design), and a single train GE-7FA gas turbine with HRSG and 1,800 psig / 1,000 ^oF / 1,000 ^oF, 3.0 in. Hga steam cycle.
  Implication: Plant thermal efficiency and cost improvement as compared to Case 8.

- **Case-11**: Commercially Offered Future IGCC plant with shift reaction and CO_2 Capture.
  Same as Case-10 but with shift reactor and CO_2 capture, compression and liquefaction system.
  Implication: Provides direct comparison with Case-10 to isolate the impact of CO_2 capture for commercially offered future IGCC plants.

Advanced Chemical Looping Cases

- **Case-12**: Indirect Gasification of Coal via Chemical Looping (Base Case for comparison to Case-13)
  Coal fired Combined Cycle Plant without CO_2 capture utilizing a advanced Chemical Looping gasification technology and a single train GE-7FA gas turbine with HRSG and 1,800 psig / 1,000 ^oF / 1,000 ^oF, 3.0 in. Hga steam cycle.
  Implication: Provides a reference point for the performance & economic analyses of Case 13.

- **Case-13**: Indirect Gasification of Coal and CO_2 Capture via Chemical Looping.
  Same as Case-12 but with CO_2 capture, compression and liquefaction system.
  Implication: Provides direct comparison with Case-12 to isolate the impact of CO_2 capture for advanced chemical looping gasification based future plants.

Common Parameters:

All plants were designed for the identical coal and limestone analyses, ambient conditions, site conditions, etc. such that each case study provides results which are directly comparable, on a common basis, to all other cases analyzed within this work. The ambient conditions used for all material and energy balances were based on the standard American Boiler Manufacturers Association (ABMA) atmospheric conditions (i.e. 80 ^oF, 14.7 psia, 60 percent relative humidity).

The steam cycle represents another common thread among all the cases. It is nearly identical for all the combustion cases differing only in the arrangement of the low level heat recovery system or small process steam extractions in some cases (Case-5 and Case-7). The steam turbine for the combustion cases is an 1,800 psig 1,000 ^oF / 1,000 ^oF single reheat machine with a main steam flow of 1,400,555 lbm/hr and a condenser pressure of 3.0 in. Hga. The cold reheat flow is 1,305,632 lbm/hr. The main steam flow is identical for all the combustion cases. The reheat steam flow is also identical for all the combustion cases except for Case-6 in which there is a slight increase as required by the low level heat recovery system. Six extraction feedwater heaters are used and the final feedwater temperature is 470 ^oF.

The steam cycles utilized for the four IGCC and two advanced Chemical Looping cases use the same steam conditions (1,800 psig 1,000 ^oF / 1,000 ^oF, 3.0 in. Hga) as for the combustion cases. However, somewhat different steam flows are required by the respective gasifier, gas turbine and heat recovery system arrangements. Additionally, there are no extraction feedwater heaters in these cases due to the large quantity of low level heat recovery required for these combined cycles.

All plants that included CO_2 capture systems were designed for a minimum of 90 percent CO_2 capture. Additionally, Table 2.0.1 shows the CO_2 product purity specification that was used as a design guideline for the Gas Processing Systems (GPS's) included in this study. It should be understood that product specifications for the CO_2 are very dependent on the individual oil field being flooded. All the GPS's in this study, except for Case-3, were designed with a goal to meet or exceed this purity specification. Case-3 was different in that the CO_2 product for this case was defined to be available for "sequestration only" and therefore, its requirements were somewhat less stringent. The requirements for Case-3 are described separately in Section 2.3. All cases (except Case-3) met or exceeded all component purity specifications listed in Table 2.0.1 except for O_2 in Cases 2, 4, and 6 and N_2 in Cases 2, 4, 5, and 6.

Table 2.0. 1: Dakota Gasification Project's CO_2 Specification for EOR

Constituant	Units	Value
CO_2	vol. %	96.0
H_2S	vol. %	0.9
CH_4	vol. %	0.7
C_2 + HC's	vol. %	2.3
CO	vol. %	0.1
N_2	vppm	< 300
H_2O	vppm	< 20
O_2	vppm	< 50

The nitrogen concentration specified in Table 2.0.1 is < 300 ppmv. It should be noted that according to Charles Fox of Kinder Morgan (Fox, 2002), this specification is very conservative as his company specifies a maximum nitrogen concentration of 4 percent (by volume) to control the minimum miscibility pressure. In Cases 2, 4, 6 and 7 the nitrogen concentration in the liquid product ranged between 9,700-11,800 ppmv. All other cases met or exceeded the nitrogen purity specifications. The exact reasoning behind the very low nitrogen specification listed in Table 2.0.1 is not clear.

A very low concentration of oxygen (< 50 ppmv) in particular is also specified in Table 2.0.1 for meeting current pipeline operating practices, presumably due to the corrosive nature of the oxygen. Hence, for Cases 2, 4, 6, and 7, whereby the final CO_2 liquid product was found to contain between 1,800-11,600 ppmv of O_2, the design of the transport pipe to an EOR site for example would have to take this characteristic under consideration.

The flue gas purification systems for these cases (2, 4, 5, and 6) utilize CO_2 refrigeration in the rectifier. In order to get to low enough temperatures to selectively let the oxygen escape without losing more CO_2, the CO_2 refrigerant would need be cool enough to solidify. If it were desired to meet the oxygen purity specification with this type of process, the loss of CO_2 from the product would be excessive, probably significantly more than 50 percent (Vogel, 2003).

Another possible way to meet the specification for oxygen would be to use a refrigerant in the condenser that will not freeze at the very cold temperatures required (ethane for example). Cooling the CO_2 stream at the top of the rectifier to a low enough temperature to almost freeze it would allow the oxygen to escape without compromising the CO_2 recovery fraction specification. Operating close to the freezing point would however inevitably cause operating problems and therefore, this process option was not chosen here.

Plant Site Design Basis and Scope:

The plants designed for this conceptual level study are all assumed to be located on a common Greenfield site, and are assumed to be operated under common conditions of fuel, limestone, utility and environmental standards. This section is intended to describe the host site conditions, which will be used as a common design basis for all these plants.

The generic plant site, which is common to all study cases, is assumed to be in the Gulf Coast region of southeastern Texas. The site consists of approximately 300 usable acres within 15 miles of a medium-sized metropolitan area, with a well-established infrastructure capable of supporting the required construction work force. The area immediately surrounding the site has a mixture of agricultural and light industrial uses. The site is served by a river of adequate quantity for use as makeup cooling water with minimal pretreatment and for the receipt of cooling system blowdown discharges.

A railroad line suitable for unit coal trains passes within 2-1/2 miles of the site boundary. A well-developed road network serves the site, capable of carrying AASHTO H-20 S-16 loads and with overhead restriction of not less than 16 feet (Interstate Standard).

The site is on relatively flat land with a maximum difference in elevation within the site of about 30 feet. The topography of the area surrounding the site is rolling hills, with elevations within 2,000 yards not more than 300 feet above the site elevation. The site is within Seismic Zone 1, as defined by the Uniform Building Code. The following list further describes the assumed site characteristics.

- The site is Greenfield with no existing improvements or facilities.
- The site is relatively clear and level with no characteristics that would cause any unusual construction problems.
- The structural strength of the soil is adequate for spread footings (no piling is required) at this site.
- No rock excavation is required on this site.
- An abundant sub-surface water supply is assumed available on this site.

The boundary limit for these plants includes the complete plant facility within the "fence line". It encompasses all equipment from the coal pile to the busbar and includes the coal receiving and water supply systems and terminates at the high-voltage side of the main power transformers. Also, for the cases with CO_2 capture the boundary includes the gas processing system and terminates at the outlet flange of the CO_2 product pipe. The scope of supply is further defined by the following list.

- Site preparation and site improvements
- Foundations, buildings, and structures required for all plant equipment and facilities
- General support facilities for administration, maintenance and storage
- Coal and limestone receiving, storage, and handling systems
- Boiler / Gasifier Island from coal feed through gas cleanup system including associated solids handling systems
- Power block, including steam turbine, heat rejection, and makeup water systems, and gas turbines (where applicable)
- Gas processing systems to produce the CO_2 product gas (where applicable)
- Oxygen supply systems (where applicable)
- Plant electrical distribution, lighting, and communication systems
- High-voltage electrical system through step-up transformer
- Instruments and controls
- Miscellaneous power plant equipment

The electrical facilities within the plant scope include all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, foundations, and standby equipment.

Additionally, the following utilities are assumed to be available at the site boundary.

- Communication lines
- Electrical power for plant construction
- Potable water and sanitary sewer connections
- Electrical transmission facilities and lines

Plant Ambient Design Conditions:

Table 2.0.2 lists ambient and other relevant characteristic assumptions for this site. The ambient conditions used for all material and energy balances were based on the standard American Boiler Manufacturers Association (ABMA) atmospheric conditions (i.e. 80 ^{o}F, 14.7 psia, 60 percent relative humidity). All steam cycles used a condenser pressure of 3.0 inches of mercury (absolute) as shown in Table 2.0.2. For equipment sizing, the maximum dry bulb temperature is 95°F, and the minimum dry bulb temperature for mechanical design is 20°F.

Table 2.0. 2: Site Characteristics

Design Parameter	Value
Elevation (ft)	500
Design Atmospheric Pressure (psia)	14.7
Design Temperature, dry bulb (^{o}F)	80
Design Temperature, wet bulb (^{o}F)	52
Design Relative Humidity (percent)	60
Design Condenser Pressure (in Hga)	3
Ash Disposal	Off Site
Water Source	River

Consumables:

Table 2.0.3 shows the design coal analyses (ultimate, proximate and Higher Heating Value). The coal is classified as a medium volatile bituminous coal. Table 2.0.4 shows the design limestone analysis used for these study cases. Additionally, a small quantity of natural gas is used in some of these plants for desiccant drying in the Gas Processing Systems and Air Separation Units. For the purpose of this study, the natural gas was assumed to be pure Methane (CH_4) with a higher Heating Value (HHV) of 23,896 Btu/lbm.

Table 2.0. 3: Design Coal Analysis (Medium Volatile Bituminous)

Ultimate Analysis		
Constituent	**(Units)**	
O2	(wt. frac.)	0.0316
N2	"	0.0146
H2O	"	0.0399
H2	"	0.0357
Carbon	"	0.6205
Sulfur	"	0.0234
Ash	"	0.2343
Total	"	1.0000
Proximate Analysis		
Constituent	**(Units)**	
Fixed Carbon	(wt. frac.)	0.5483
Volatile Matter	"	0.1775
Moisture	"	0.0399
Ash	"	0.2343
Total	"	1.0000
HHV Coal	(Btu/lbm)	**11074**

Table 2.0. 4: Design Limestone Analysis

Constituent	Weight Fraction
$CaCO_3$	0.95
Inerts	0.05

Plant Services:

The following services and support systems are available at the plant as a part of the balance-of-plant systems.

Auxiliary Power Systems:

- 7,200 V system for motors above 3,000 hp.
- 4160 V system for motors from 250 to 3,000 hp.
- 480 V system for motors from 0 to 250 hp and miscellaneous loads.
- Emergency diesel generator (480 V) to supply loads required for safe and orderly plant shutdown. Instruments and controls and other loads requiring regulated (1-percent) 208/120 Vac power are supplied from this source.

- 250 Vdc system motors and, via static inverters, uninterruptible ac power for the integrated control and monitoring system, intercommunication.
- 125 Vdc system for dc controls, emergency lighting, and critical tripping circuits including the plant shutdown system.

Cooling Water:
- Cooling water (from the cooling towers) is available at between 20 and 30 psig, 90°F maximum temperature. The water is periodically chlorinated, and pH is maintained at 6.5 to 7.5. The cooling towers receive makeup water from the river.
- Auxiliary cooling water, which uses de-mineralized water treated for corrosion control, at 60 to 80 psig and 105°F, is available for small heat loads (e.g., control oil coolers). The pH is maintained at about 8.5.

Compressed Air:
- Instrument air filtered and dried to -40° dew point at 80 to 100 psig and 110°F (maximum).
- Service air at 80 -100 psig and 110°F (maximum).

Lube Oil:
- Lube oil from the conditioning system, with particulate matter removed to 10 µm or lower.

Hydrogen and Carbon Dioxide:
- H_2 and CO_2 for generator cooling and purging from storage.

Nitrogen:
- N_2 for equipment blanketing against corrosion during shutdown and lay-up.

Raw Water:
- Filtered river water. Additional water treatment will be included for potable water, etc.

Structures and Foundations:

Structures are provided to support and permit access to all plant components requiring support to conform to the site criteria. The structure(s) are enclosed if deemed necessary to conform to the environmental conditions.

Foundations are provided for the support structures, pumps, tanks, and other plant components. A soil-bearing load of 5,000 lbm/ft^2 is used for foundation design.

2.1. Case-1: Air Fired CFB without CO_2 Capture (Base Case for Comparison to Combustion Cases)

This section describes a power plant utilizing a coal-fired, atmospheric pressure, Circulating Fluidized-Bed (CFB) steam generator and a subcritical steam plant. This plant represents "business as usual" and does not capture CO_2. The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

This case represents the Base Case for the combustion cases and was included in the study to provide a basis for comparison with all the other CFB based options which include CO_2 removal (Cases 2-7). The Base Case for this study is defined as the selected unit firing coal at full load, utilizing air as the oxidant, without capturing CO_2 from the flue gas. This represents the "business as usual" operating scenario and is used as the basis of comparison for the various Combustion based CO_2 removal options investigated in this study. Comparison of Cases 2-7 to this case indicates the impact of the various CO_2 capture options investigated.

A brief performance summary for this plant reveals the following information. The Case-1 plant produces a net plant output of 193,034 kW. The net plant heat rate and thermal efficiency are calculated to be 9,611 Btu/kWh and 35.5 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 2.00 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.2.3.

Case-1 Unit Selection and Description:

The steam generator unit design selected and analyzed in this study was based on an existing built and operating utility scale Circulating Fluidized Bed (CFB) unit firing medium volatile bituminous coal (Heretofore referred to as Base Case CFB Coal). Although the unit is an existing one, the boiler for this study was completely re-designed to conform to current design guidelines. The size of the selected unit (~210 MW gross) is typical of many large utility scale CFB units. An existing unit was selected in order to minimize the level of effort required to obtain performance, cost, and economic parameters for the Base Case. The Base Case is then used as a basis of comparison for the various Combustion-based CO_2 capture options investigated in this study. The estimated costs for this CFB were also updated and the various drawings shown in this report for this case reflect current design practices for CFB boilers.

The furnace bottom is a single cell design. The unit is configured with 2 cyclones, 2 external fluidized bed heat exchangers, 2 fluidized bed ash coolers, and a single convection pass. The two air heaters (primary and secondary) are of the tubular design. The unit selected is representative in many ways of a large number of coal fired CFB units in use today. The drum type sub-critical pressure unit is designed to generate about 1.4×10^6 lbm/hr of steam at 1,800 psig and 1,005 ^oF with reheat also to 1,005 ^oF. These represent common steam cycle operating conditions for existing utility scale CFB power generation systems of this size. Outlet steam temperature control is provided with de-superheating spray and hot solids biasing to the external fluidized bed heat exchangers.

The coal-fired CFB produces low levels of NOx by combusting the coal in the circulating bed, where temperatures are maintained in the vicinity of 1,600°F. This low temperature minimizes the formation of thermal NOx, while allowing the oxidation of carbon and the capture of sulfur by calcium to proceed to completion. The addition of dry limestone to the bed, in appropriate particle sizes, provides the calcium carbonate for sulfur capture.

The CFB steam generator provides high-pressure steam to power a single-reheat subcritical steam turbine with steam turbine conditions of 1,800 psig/1,000°F throttle, 1,000°F reheat, and 3.0 in. Hga condenser pressure.

2.1.1. Case-1 Boiler Island Process Description and Equipment

The Base Case can be described as the existing unit firing coal at full load and utilizing air as the oxidant without capturing CO_2 from the flue gas. This represents the "business as usual" operating scenario and is used as the basis of comparison for the various Combustion-based CO_2 removal options investigated in this study. The first step in the development of a Base Case was to set up a computer model of the boiler. The boiler computer model was then used for complete analysis of the Base Case.

Development of the Boiler Island Computer Model

The first step in the calculation of a Base Case was to set up a steady state performance computer model of the steam generator unit and associated equipment. This involves calculating or obtaining all the geometric information for the unit as required by the proprietary Reheat Boiler Program (RHBP). The RHBP provides an integrated, steady state performance model of the Boiler Island including the combustor, cyclones, external heat exchangers, air heaters, fans, and steam temperature control logic. The RHBP is used to size components and/or predict performance of existing components. For the Base Case, since the boiler is an existing unit and boiler island component sizes are known, the RHBP was used exclusively for calculating unit performance.

Using the Boiler Island computer model and providing it with various steam side inputs (mass flows, temperatures, pressures, etc.) from the agreed upon Maximum Continuous Rating (MCR) steam turbine material and energy balance, the model was run and performance was calculated for the Base Case. The Base Case performance summary for the overall power plant system is described in Section 2.1.3. The Boiler Island performance is defined in Section 2.1.1.1 and the steam cycle performance is provided in Section 2.1.2.1.

2.1.1.1. Process Description and Process Flow Diagrams

The simplified gas side process flow diagram for the Case-1 Boiler Island is shown in Figure 2.1.1. This process description briefly describes the function of the major equipment and systems included within the Boiler Island. Selected mass flow rates (lbm/hr) and temperatures (^{o}F) are shown on this figure. Complete data for all state points are shown in Table 2.1.1.

In this concept coal (Stream 1) is reacted with preheated air (Streams 12, 15) in the Combustor section of the Circulating Fluidized Bed (CFB) system. The combustor is a water-cooled refractory lined vessel designed to evaporate high-pressure steam. The air (Streams 12, 15, 17) is supplied from primary, secondary and fluidizing air fans. The products of combustion leaving the Combustor flow through cyclones where most of the entrained hot solids are removed and recirculated to the Combustor. By properly splitting the flow of hot recirculated solids leaving the cyclone bottom, between an uncooled stream which flows directly back to the Combustor and the External Heat Exchanger where the solids are cooled before returning to the Combustor, the temperature in the combustor can be controlled to the desired level for a wide variety of operating conditions. Exchanging heat with the power cycle working fluid cools the solids in the External Heat Exchanger.

Draining hot solids from the combustor through water-cooled fluidized bed ash coolers (Stream 18) controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash coolers is feedwater from the final extraction feedwater heater of the steam cycle.

The flue gas leaving the Cyclones (Stream 3) is cooled in heat exchanger sections located in the convection pass of the system, also by exchanging heat with the power cycle working fluid. The flue gas leaving the convection pass heat exchanger sections (Stream 5) is further cooled in the Air Heaters. The flue gas leaving the Air Heaters (Stream 6) is cleaned of fine particulate matter in a baghouse and enters the Induced Draft (ID) Fan (Stream 7). The flue gas leaving the ID Fan (Stream 8) is then discharged to the atmosphere through a stack.

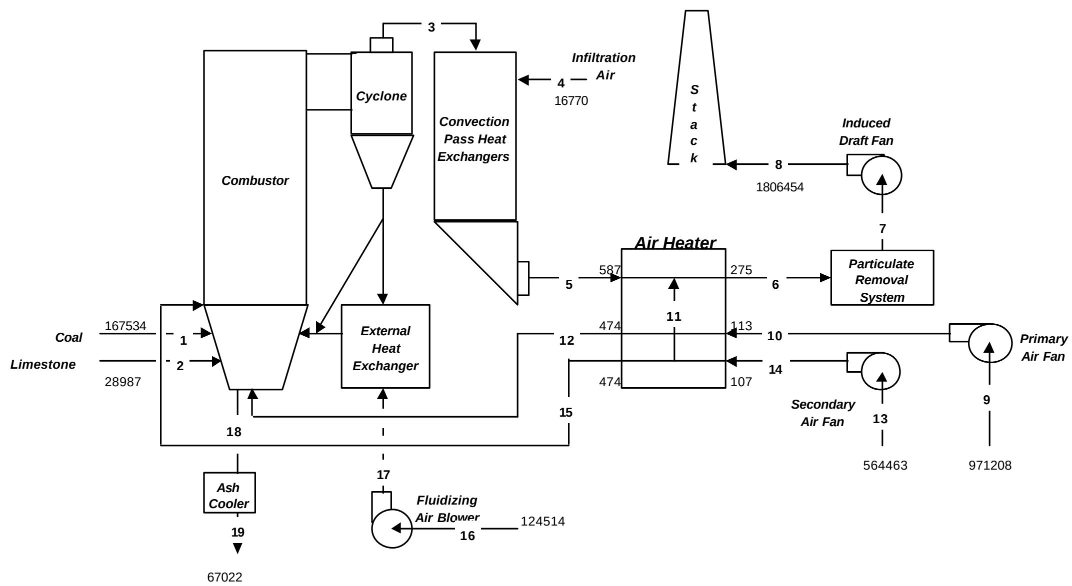


Figure 2.1. 1: Case-1 Simplified Boiler Island Gas Side Process Flow Diagram (Base Case)

2.1.1.2. Material and Energy Balance

Table 2.1.1 shows the Boiler Island material and energy balance for Case-1. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-1 simplified PFD for the Boiler Island (Figure 2.1.1).

The performance shown in Table 2.1.1 was calculated at Maximum Continuous Rating (MCR) conditions for this unit. The MCR condition is defined as high-pressure turbine inlet conditions of 1,400,555 lbm/hr, 1,815 psia, and 1,000 ^{o}F and intermediate-pressure turbine inlet conditions of 1,305,632 lbm/hr, 469 psia, and 1,000 ^{o}F. This MCR condition definition was used for the Base Case and all other CFB based cases in this study (Note: It should be understood that reheat flow was increased slightly in some cases due to required differences in low level heat recovery arrangements). The boiler for Case-1 was fired with 20 percent excess air and the resulting boiler efficiency calculated for this case was 89.48 percent (HHV basis) with an air heater exit gas temperature of 275 ^{o}F.

Table 2.1. 1: Case-1 Boiler Island Gas Side Material and Energy Balance

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10
O_2	(Lbm/hr)	5294		60726	3839	64565	64565	64565	64565	222310	222310
N_2	"	2446		1261364	12716	1274081	1274081	1274081	1274081	736467	736467
H_2O	"	6685		81384	215	81599	81599	81599	81599	12431	12431
CO_2	"	0		385427		385427	385427	385427	385427		
SO_2	"	0		783		783	783	783	783		
H_2	"	5981									
Carbon	"	103955									
Sulfur	"	3920									
CaO	"	0									
$CaSO_4$	"	0									
$CaCO_3$	"	0	27538								
Ash	"	39253	1449								
		Coal	Limestone	Flue Gas to BP	Infiltration Air	Flue Gas to AH	Flue Gas to PR	Flue Gas to ID	FGas from ID	Primary Air	Primary Air
Total Gas	(Lbm/hr)			1789684	16770	1806454	1806454	1806454	1806454	971208	971208
Total Solids	"	**167534**	28987		0	0	0	0	0		
Total Flow	"	167534	28987	1789684	16770	1806454	1806454	1806454	1806454	971208	971208
Temperature	(Deg F)	**80**	80	**1600**	80	**587**	**275**	275	291	**80**	113
Pressure	(Psia)	**14.7**	14.7	14.7	14.7	14.4	14.2	13.8	14.7	14.7	17.4
$h_{sensible}$	(Btu/lbm)	0.000	0.000	414.525	0.000	128.747	48.406	48.406	52.507	0.000	8.104
Chemical	(10^6 Btu/hr)	1855.272									
Sensible	(10^6 Btu/hr)	**0.000**	0.000	741.869	0.000	232.575	87.444	87.444	94.852	0.000	0.101
Latent	(10^6 Btu/hr)	**0.000**	0.000	85.453	0.225	85.678	85.678	85.678	85.678	13.053	13.053
Total Energy$^{(1)}$	(10^6 Btu/hr)	1855.272	0.000	827.322	0.225	318.253	173.122	173.122	180.531	13.053	13.154

Constituent	(Units)	11	12	13	14	15	16	17	18	19
O_2	(Lbm/hr)	0	222310	129206	129206	129206	28501	28501		
N_2	"	0	736467	428032	428032	428032	94419	94419		
H_2O	"	0	12431	7225	7225	7225	1594	1594		
CO_2	"									
SO_2	"									
H_2	"									
Carbon	"								2079	2079
Sulfur	"								0	
CaO	"								9258	9258
$CaSO_4$	"								14982	14982
$CaCO_3$	"								0	
Ash	"								40703	40703
		AH Lkg Air	Primary Air	Secondary Air	Secondary Air	Secondary Air	Fluidizing Air	Fluidizing Air	Ash Drain	Ash Drain
Total Gas	(Lbm/hr)	0	971208	564463	564463	564463	124514	124514		
Total Solids	"								67022	67022
Total Flow	"	0	971208	564463	564463	564463	124514	124514	67022	67022
Temperature	(Deg F)	113	474	80	107	474	80	178	1600	520
Pressure	(Psia)	17.4	17.2	14.7	16.9	16.6	14.7	23.7	14.7	14.7
$h_{sensible}$	(Btu/lbm)	8.104	96.983	0.000	6.558	96.983	0.000	23.968	407.729	95.391
Chemical	(10^6 Btu/hr)								29.301	29.301
Sensible	(10^6 Btu/hr)	0.000	94.190	0.000	3.702	54.743	0.000	2.984	27.327	6.393
Latent	(10^6 Btu/hr)	0.000	13.053	7.586	7.586	7.586	1.673	1.673	0.000	0.000
Total Energy$^{(1)}$	(10^6 Btu/hr)	0.000	107.244	7.586	11.288	62.330	1.673	4.658	56.627	35.694

Notes:
(1) Energy Basis; Chemical based on Higher Heating Value (HHV); Sensible energy above 80F; Latent based on 1050 Btu/Lbm of water vapor

2.1.1.3. Boiler Island Equipment

Figures 2.1.2 and 2.1.3 show general arrangement drawings of the Case-1 CFB boiler. The complete Equipment List for Case-1 is shown in Appendix I. Appendix II shows several drawings of the Boiler (key plan, plan view, side elevation, and various section views). The major components shown in these drawings include the combustor, fluidized bed ash coolers, fuel silos and feed system, sorbent silo and feed system, cyclones, seal pots, external fluidized bed heat exchangers (FBHE), convection pass, superheater, reheater, economizer, steam drum including circulation system, air heaters, baghouse, and draft system.

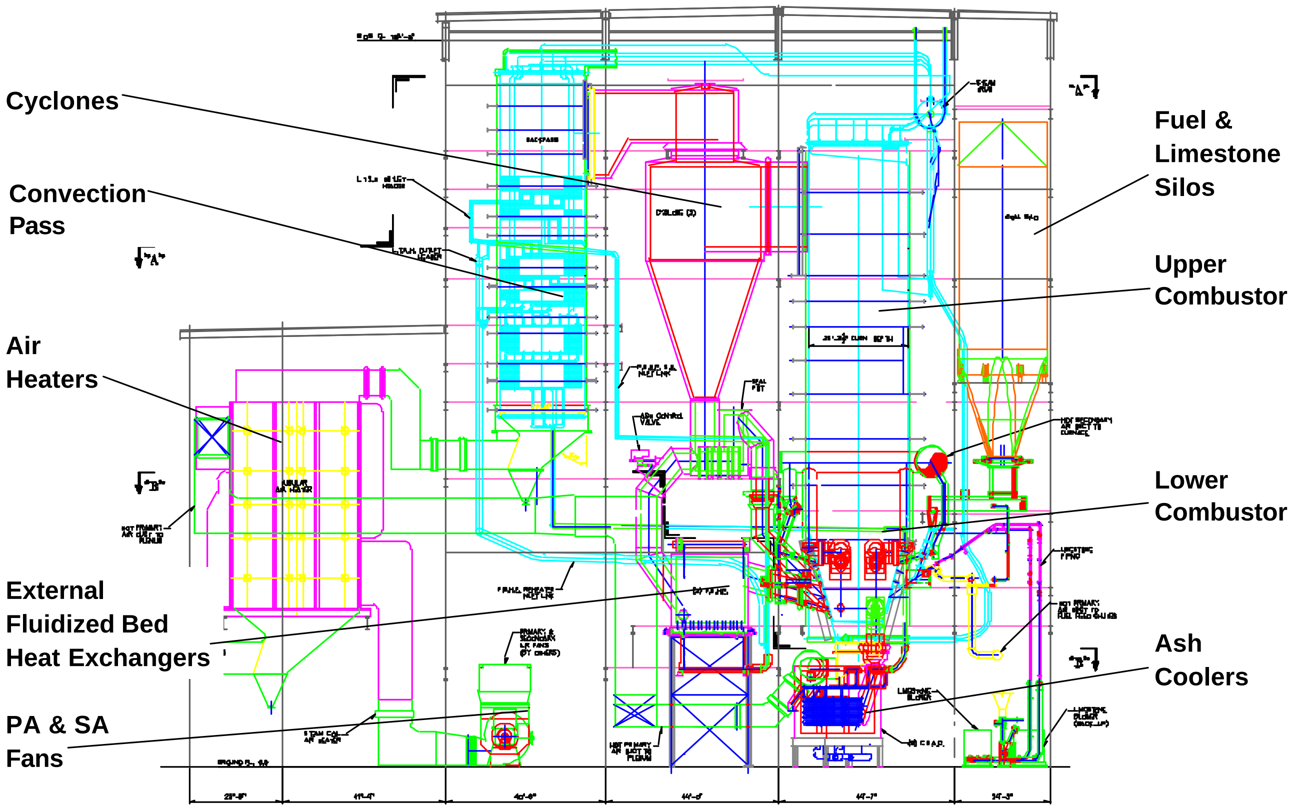


Figure 2.1. 2: Case-1 Boiler Island General Arrangement Drawing - Side Elevation of Nominal 210 MW-gross CFB Steam Generator

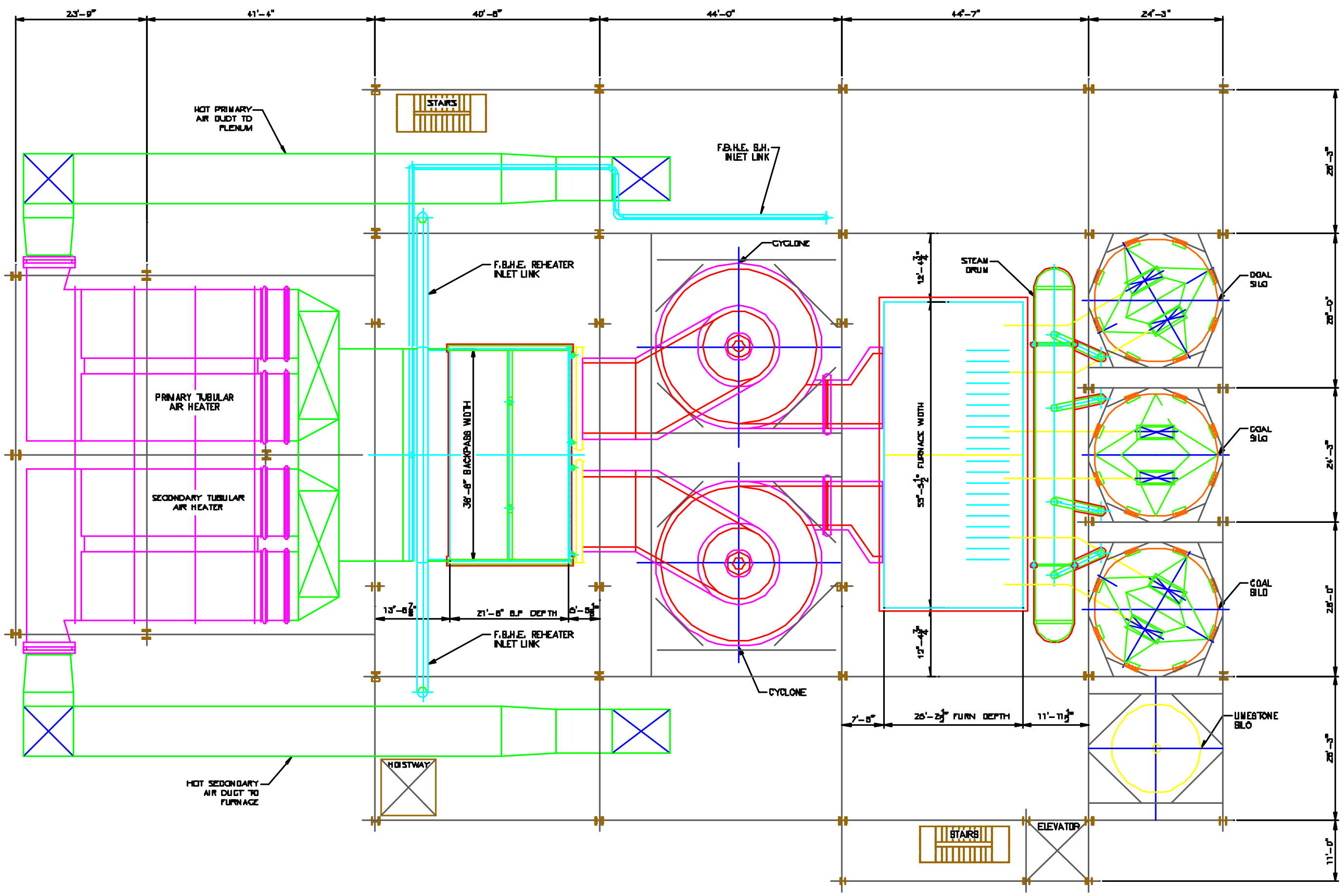


Figure 2.1. 3: Case-1 Boiler Island General Arrangement Drawing - Plan View of Nominal 210 MW-gross CFB Steam Generator

Combustor:

The single cell combustor for this unit is about 55 ft wide, 25 ft deep and 118 ft high. Crushed fuel, sorbent, and recycle solids are fed to the lower portion of the combustor. Primary air is fed to the combustor bottom through a grid plate and the secondary air is supplied higher up in the lower combustor region. The lower combustor region is a tapered rectangular section formed from fusion welded waterwall tubing with a refractory lining. Combustion occurs throughout both the lower and upper combustor, which are filled with bed material. The upper combustor section is a rectangular straight walled section formed from fusion welded waterwall tubing. Evaporator panels are included in the upper combustor area along the rear wall. The combustor walls and evaporator panels are cooled with continuous forced re-circulation of water from the steam drum. Combustor bed temperature is maintained at an optimum level for sulfur capture and combustion efficiency by balancing heat absorption in the combustor and in the FBHE.

Ash Coolers:

Draining hot solids through water-cooled ash coolers controls solids inventory in the system while recovering heat from the hot ash. Feedwater from the final extraction feedwater heater of the steam cycle is used as the cooling water source for the ash coolers. Two ash coolers are used in this case.

Fuel Feed System:

The fuel feed system transports prepared coal from the storage silos to the lower combustor. The system includes the storage silos, silo isolation valves, fuel feeders, feeder isolation valves, and fuel piping to the furnace. Three storage silos are used in this case.

Sorbent Feed System:

The limestone feed system pneumatically transports prepared limestone from the storage silo to the lower combustor. The system includes the storage silos, silo isolation valves, rotary feeders, blower, and piping from the blower to the furnace injection ports. A single limestone storage silo is used in this case.

Cyclones:

Flue gas and entrained solids exit the upper combustor and enter the cyclones. The cyclones are shaped like cylindrical cones constructed from steel plate with a multiple layer refractory lining. Vortex finders (sometimes called re-entrant throats) are included in the cyclone outlet to improve collection efficiency. Solids are separated from the flue gas in the cyclone and fall into a seal pot. Well over 99 percent of the entrained solids are captured in the cyclones. Two 28 ft diameter cyclones are used in this case.

Seal Pot:

The seal pot is a device that provides a pressure seal between the combustor, which is at relatively high pressure (~ 40 inwg at the bottom), and the cyclone that is at atmospheric pressure. The seal pot is a non-mechanical valve, which moves solids collected in the cyclones back to the combustor. The seal pot is constructed of steel plate with a multiple layer refractory lining with fluidizing nozzles located along the bottom to enhance solids flow. Some of the solids flow directly from the seal pot back to the combustor while other solids are diverted through a plug valve, through the external Fluidized Bed Heat Exchangers (FBHE), and then back to the combustor. Two seal pots located directly beneath each cyclone are used for this case.

Fluidized Bed Heat Exchangers:

The external FBHE's are compartmentalized bubbling bed heat exchangers containing immersed tube bundles, which cool the hot solids from the seal pot. The tube bundles in the FBHE's include finishing superheater, finishing reheater and evaporator sections. The tube bundles are constructed with bare tubes. Very high heat transfer rates are obtained in the FBHE's because of the high bed density. The FBHE's are constructed with water cooled fusion welded enclosure walls. Fluidizing air is provided at various locations along the bottom of the bed. The cooled solids leaving the FBHE's are returned to the combustor. Two FBHE's are used for this case.

Convection Pass:

Flue gas leaving the cyclones is ducted into the Convection Pass, which includes a low temperature superheater section, a low temperature reheater section, and an economizer section. The tube banks used in the convection pass use bare tubes. The convection pass is constructed similarly to those used for pulverized coal firing with fusion welded steam cooled enclosure walls, water cooled hanger tubes, and fly ash hoppers located at the bottom. Soot-blowers are used to keep the various heat transfer surfaces clean.

Superheater:

The superheater is divided into two major sections. Saturated steam leaving the steam drum first cools the roof and convection pass walls before supplying the horizontal low temperature superheater section also located in the convection pass. Steam leaving the low temperature superheater section first flows through the de-superheater spray stations and then to the finishing superheat section located in the external fluidized bed heat exchanger. Steam leaving the finishing superheater is piped to the high-pressure turbine where it is expanded to reheat pressure and then returned to the low temperature reheat section.

Reheater:

The reheater is also divided into two sections, a low temperature section followed by a finishing section. Steam is supplied to the low temperature reheater from the high-pressure turbine exhaust. The low temperature section is a horizontal section located in the convection pass between the low temperature superheater and the economizer. Steam leaving the low temperature reheater is piped to the de-superheating spray station and then to the finishing reheat section located in the external fluidized bed heat exchanger. Steam leaving the finishing reheater is returned to the intermediate pressure turbine where it continues its expansion through the intermediate and low-pressure turbines before being exhausted to the condenser.

Economizer:

The flue gas leaving the low temperature reheater section in the convection pass is then further cooled in an economizer section also located in the convection pass. The economizer is comprised of two banks of horizontal tubes, which heats high-pressure boiler feedwater. The water exiting the economizer tube banks then cools the convection pass hanger tubes, which support the low temperature superheater and reheater sections, before it is supplied to the steam drum. The feedwater supplying the economizer is piped from the final extraction feedwater heater and the ash coolers.

Air Heater:

Two tubular regenerative air heaters (primary and secondary) are used to cool the flue gas leaving the economizer by heating both the primary and secondary air streams prior to combustion in the furnace.

2.1.2. Case-1 Balance of Plant Equipment and Performance

The balance of plant equipment described in this section includes the steam cycle performance and equipment, the draft system equipment, the cooling system equipment, and the material handling equipment (coal, limestone, and ash). Refer to Appendix I for equipment lists and Appendix II for drawings.

2.1.2.1. Steam Cycle Performance

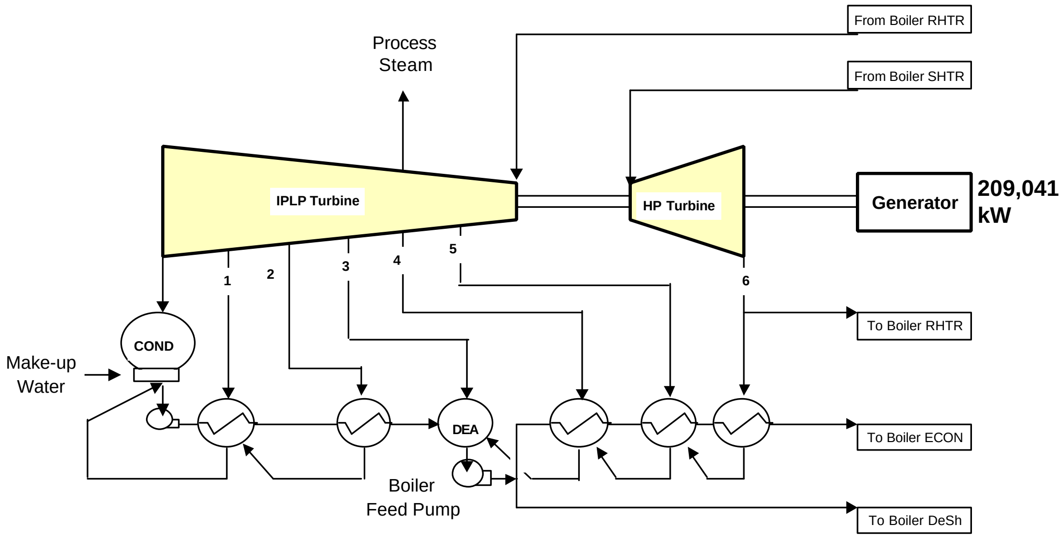
The steam cycle for the Case-1 is shown schematically in Figure 2.1.4. Figure 2.1.5 shows the associated Mollier diagram which illustrates the process on enthalpy - entropy coordinates.

The steam cycle starts at the condenser hot well, which is a receptacle for the condensed steam from the exhaust of the steam turbine. The condensate flows to the suction of the condensate pumps, which increase the pressure of the fluid by a nominal 250 psi to transport it through the piping system and enable it to enter the open contact heater, or deaerator. The condensate passes through a gland steam condenser (GSC) first, followed in series by two low-pressure feedwater heaters. The heaters successively increase the condensate temperature to a nominal 221°F by condensing and partially sub-cooling steam extracted from the LP steam turbine section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (now referred to as heater drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the condenser.

The condensate entering the deaerator is heated and stripped of noncondensable gases by contact with the steam entering the unit. The steam is condensed and, along with the heated condensate, flows by gravity to a deaerator storage tank. The boiler feedwater pumps take suction from the storage tank and increase the fluid pressure to a nominal 2200 psig. Both the condensate pump and boiler feed pump are electric motor driven. The boosted condensate flows through three more high-pressure feedwater heaters, increasing in temperature to 470°F at the entrance to the boiler economizer section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the deaerator.

Within the boiler the feedwater is evaporated and finally superheated. The high-pressure superheated steam leaving the finishing superheater (1,400,555 lbm/hr of steam at 1,815 psia and 1,000 ^{o}F) is expanded through the high-pressure turbine. Reheat steam (1,305,632 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,000 ^{o}F. These conditions (temperatures, pressures) represent common steam cycle operating conditions for existing utility scale CFB power generation systems in use today. The reheated steam expands through the intermediate and low-pressure turbines before exhausting to the condenser. The condenser pressure used for Case-1 and all other cases in this study was 3.0 in. Hga.

The steam turbine performance analysis results show the generator produces 209,057 kW output and the steam turbine heat rate is 8,146 Btu/kWh.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Steam Turbine Power Outpu	725	Boiler Heat Input	1673
Process Steam Heat Loss	0	BFP & CP Input	12
Condenser Loss	960	**Total Energy Input**	1685
Total Energy Output	1685	**In - Out**	0

Turbine Heat Rate 8147 (Btu/kwhr)

Figure 2.1. 4: Case-1 Steam Cycle Schematic and Performance

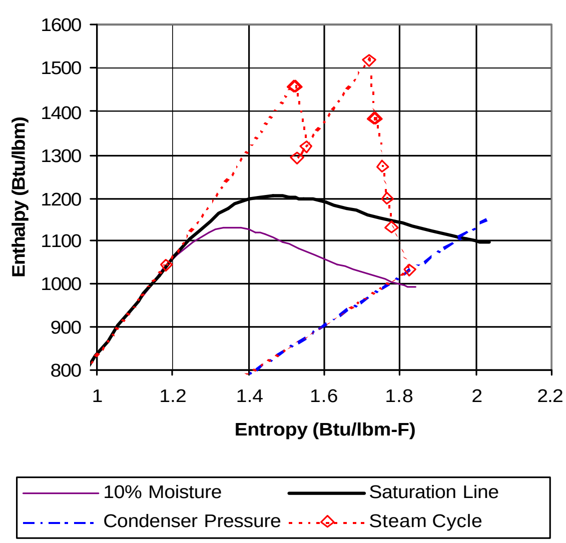


Figure 2.1. 5: Case-1 Steam Cycle Mollier Diagram

2.1.2.2. Steam Cycle Equipment

This section provides a brief description of the major equipment (steam turbine, condensate and feedwater systems) utilized for the steam cycle of this case.

Steam Turbine:

The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the boiler passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheated steam flows through the reheat stop valves and intercept valves and enters the IP section at 465 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser.

The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for economic unit cycling.

Condensate and Feedwater Systems:

The function of the condensate system is to pump condensate from the condenser hot well to the deaerator, through the gland steam condenser and the LP feedwater heaters. The system consists of one main condenser; two 50 percent capacity, motor-driven vertical condensate pumps; one gland steam condenser; two LP heaters, and one deaerator with a storage tank.

Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump feedwater from the deaerator storage tank to the boiler economizer. Two motor-driven boiler feed pumps are provided to pump feedwater through the three stages of HP feedwater heaters. Pneumatic flow control valves control the recirculation flow. In addition, the suctions of the boiler feed pumps are equipped with startup strainers, which are utilized during initial startup and following major outages or system maintenance.

2.1.2.3. Other Balance of Plant Equipment

The systems for draft, solids handling (coal, limestone, and ash), cooling, electrical, and other BOP systems are described in this section for Case-1.

Draft System:

The flue gas is moved through the boiler, baghouse and other Boiler Island equipment with the draft system. The draft system includes the primary and secondary air fans, the fluidizing air blowers, the induced draft (ID) Fan, the associated ductwork and expansion joints and the Stack, which disperses the flue gas leaving the system to the atmosphere. The induced draft, primary and secondary air fans, and fluidizing air blowers are driven with electric motors and controlled to operate the unit in a balanced draft mode with the cyclone inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

A forced draft primary air fan provides combustion air, which is split into several flow paths, as follows:

- A cold air stream flows to the fuel feeders and flows with the fuel into the furnace via a complement of fuel / air downcomers and feed spouts. This air stream provides initial fluidization of the coal mixture.
- A second hot air stream flows through a steam coil air heater followed by a regenerative air heater and flows to the fuel feed chutes.
- A third hot air stream flows through a steam coil air heater followed by a regenerative air heater; this preheated air then flows to the grate and through the fluidizing air nozzles. This air stream acts to fluidize the fuel / air mixture in the furnace and to support the initial stages of combustion. This air stream is also used for pre-mixing and firing of natural gas or No. 2 oil used for startup and warm-up.

A forced draft secondary air fan provides an air stream that is preheated in a steam coil air heater and a regenerative air preheater, and is then introduced into the furnace as secondary air.

The fluidizing air blowers provide air at higher pressure for the fluidization of the external fluidized bed heat exchangers, the fluidized bed ash coolers, and the seal pots. Air from these blowers is also used as “grease air” in several pipes used for transport of solids throughout the system

Combustion gases exit the furnace and flow through two cyclones, which separate out ash and partially burned fuel particles. These solids are recycled back to the furnace, passing through J-valves, or seal pots, located below the cyclones. The solids leaving the seal pots are then split into two streams. The first stream is uncooled and flows directly to the combustor. The second stream flows through the fluidized-bed heat exchangers where it is cooled before re-entering the furnace at the back wall.

The gas exiting the cyclones passes to the convection pass of the CFB, flowing through the low temperature reheater and low temperature superheater, and then through the economizer. The gases leaving the convection pass flow through the primary and secondary tubular air preheaters and then exit the CFB steam generator to the baghouse for particulate capture. The gases are drawn through the baghouse with the Induced Draft Fan and then are discharged to atmosphere through the Stack.

The following fans and blowers are provided with the scope of supply of the CFB steam generator:

- Primary air fan, which provides forced draft primary airflow. This fan is a centrifugal type unit, supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.1.2). The electric power required for the electric motor drive is 2,427 kW.

Table 2.1. 2: Primary Air Fan Specification

Gas Analysis			
Oxygen	(wt percent)	22.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	971208	1165450
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	17.41	
Pressure Rise	(in wg)	75.00	98

- Secondary air fan, which provides forced draft secondary airflow. This fan is a centrifugal type unit supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.1.3). The electric power required for the electric motor drive is 1,142 kW.

Table 2.1. 3: Secondary Air Fan Specification

Gas Analysis			
Oxygen	(wt percent)	22.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	564463	677356
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	16.87	
Pressure Rise	(in wg)	60.00	78

- Induced draft fan, a centrifugal unit supplied with electric motor drive and inlet damper (see Table 2.1.4). The electric power required for the electric motor drive is 2,285 kW.

Table 2.1. 4: Induced Draft Fan Specification

Gas Analysis			
Oxygen	(wt percent)	3.57	
Nitrogen	"	70.53	
Water Vapor	"	4.52	
Carbon Dioxide	"	21.34	
Sulfur Dioxide	"	0.04	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	1806454	2167745
Gas Inlet Temperature	(Deg F)	275.0	
Inlet Pressure	(psia)	13.78	
Outlet Pressure	(psia)	14.70	
Pressure Rise	(in wg)	25.50	33

- Fluidizing air blowers, centrifugal units that provide air for fluidizing and sealing the seal pots, fluidizing the external heat exchangers, fluidizing the ash coolers, and for assisting in the conveyance of cyclone bottoms through the fluidized bed heat exchangers to the furnace re-entry ports (see Table 2.1.5). The electric power required for the electric motor drive is 920 kW.

Table 2.1. 5: Fluidizing Air Blower Specification

Gas Analysis			
Oxygen	(wt percent)	22.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	124514	149417
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	23.70	
Pressure Rise	(psia)	9.00	12

Ducting and Stack:

One stack is provided with a single 19.5-foot-diameter FRP liner. The stack is constructed of reinforced concrete, with an outside diameter at the base of 70 feet. The stack is 480 feet high for adequate dispersion of criteria pollutants, to assure that ground level concentrations are within regulatory limits. Table 2.1.6 shows the stack design parameters.

Table 2.1.6: Stack Design Summary

Design Parameter	Value
Flue Gas Temperature, „ a	291
Flue Gas Flow Rate, lbm/h	1,806,454
Flue Gas Flow Rate, acfm	555,505
Particulate Loading, grains/acfm	nil

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1/4" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the three silos.

Technical Requirements and Design Basis

- Coal burn rate:
  - Maximum coal burn rate = 167,534 lbm/h = 83.8 tph plus 10 percent margin = 92 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
  - Average coal burn rate = 142,000 lbm/h = 71 tph (based on MCR rate multiplied by an 85 percent capacity factor)
  - Coal delivered to the plant by unit trains:
  - One and one-half unit trains per week at maximum burn rate
  - One unit train per week at average burn rate
  - Each unit train shall have 10,000 tons (100-ton cars) capacity
  - Unloading rate = 9 cars/hour (maximum)
  - Total unloading time per unit train = 11 hours (minimum)
  - Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
  - Reclaim rate = 300 tph
  - Storage piles with liners, run-off collection, and treatment systems:
  - Active storage = 6,600 tons (72 hours at maximum burn rate)
  - Dead storage = 50,000 tons (30 days at average burn rate)

Table 2.1. 7: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	92
Active Storage, tons	6,600
Dead Storage, tons	50000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,000 ton silo to accommodate 3 days operation.

Bottom Ash Removal:

Bottom ash, or bed drain material, constitutes approximately two-thirds of the solid waste material discharged by the CFB steam generator. This bottom ash is discharged through a complement of two bed coolers (any two of which must operate at 100 percent load on the design coal). The stripper/coolers cool the bed material to a temperature in the range of 300 ^{o}F (design coal) to a maximum of 500 ^{o}F (worst fuel) prior to discharge via rotary valves to the bed material conveying system. The steam generator scope terminates at the outlets of the rotary valves.

Fly Ash Removal:

Fly ash comprises approximately one-third of the solid waste discharged from the CFB steam generator. Approximately 8 percent of the total solids (fly ash plus bed material) is separated out in the economizer and air heater hoppers; 25 percent of the total solids is carried in the gases leaving the steam generator en route to the baghouse. Fly ash is removed from the stack gas through a baghouse filter. Particulate conditions are:

Design Specification for Particulate Removal System:

- Total solids to particulate removal system (stream 6, Figure 2.1.1) = 31,336 lbm/h
- Particle size distribution of particulate matter leaving cyclone (streams 5, 6, Figure 2.1.1), see Table 2.1.8.

Table 2.1. 8: Particle Size Distribution

% Wt. Less	Diameter (Micron, μ)
100	192
99	160
90	74
80	50
70	37
60	30
50	24
40	16
30	12
20	8
10	4
1	< 4

- Solids leaving particulate removal system (stream 7, Figure 2.1.1) meet applicable environmental regulations, see Table 2.1.9.

Table 2.1. 9: Fly Ash Removal Design Summary

Design Parameter	Value
Flue Gas Temperature, „ a	275
Flue Gas Flow Rate, lbm/h	1,806,454
Flue Gas Flow Rate, acfm	543,670
Particulate Removal, lbm/h	31,336
Particulate Loading, grains/acfm	6.724

Ash Handling:

The function of the ash handling system is to convey, prepare, store, and dispose of the fly ash and bottom ash produced on a daily basis by the boiler. The scope of the system is from the bag filter hoppers, air heater and economizer hopper collectors, and bottom ash hoppers to the truck filling stations.

The fly ash collected in the bag filter, economizer and the air heaters is conveyed to the fly ash storage silo. A pneumatic transport system using low-pressure air from a blower provides the transport mechanism for the fly ash. Fly ash is discharged through a wet unloader, which conditions the fly ash and conveys it through a telescopic unloading chute into a truck for disposal.

The bottom ash from the boiler is drained from the ash coolers, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. Ash from the fluidized-bed ash coolers is drained to a complement of screw coolers, which discharge the cooled ash to a drag chain conveyor for transport to a surge bin. The ash is pneumatically conveyed to the bottom ash silo from the surge bin.
The silos are sized for a nominal holdup capacity of 36 hours of full-load operation (1,140 tons capacity) per each. At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately 30 truckloads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.1.10: Ash Handling System Design Summary

Design Parameter	Value
Fly Ash from Baghouse, lbm/h	31,336
Hot Ash from Boiler, lbm/h	67,022
Hot Ash Temperature, „ a	1,600
Cooled Ash temperature, „ a	520
Ash Cooler Duty, MM-Btu/h	20.93

Circulating Water System:

The function of the circulating water system is to supply cooling water to condense the main turbine exhaust steam. The system consists of two 50 percent capacity vertical circulating water pumps, a multi-cell mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Condenser Analysis:

The condenser system analysis is detailed in Table 2.1.11.

Table 2.1.11: Condenser Analysis

Item	Value	Units
Pressure	3.0	in. Hga
T stm-in	115.1	°F
P stm-in	1.474	psia
H stm-in	1051.7	Btu/lbm
M drain-in	108,279	lbm/h
H drain-in	89.7	Btu/lbm
H condensate	83.0	Btu/lbm
M condensate	1,098,164	lbm/h
Q condenser	959.6	10^6 Btu/h

Waste Treatment System:

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment is housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge de-watering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 - 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water are provided. A 200,000-gallon storage tank provides a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide distributed control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes

from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Plant Layout and Plot Plan:

The Case-1 plant is arranged functionally to address the flow of material and utilities through the plant site. A plan view of the boiler, power-generating components, and overall site plan for the entire plant is shown in Appendix II.

2.1.3. Case-1 Overall Plant Performance and CO_2 Emissions

Overall plant performance and emissions for Case-1 are summarized in Table 2.1.12 and summarized below. The overall system is described previously in Section 2.1.2.

Boiler efficiency is calculated to be 89.46 percent (HHV basis). The steam cycle thermal efficiency including the boiler feed pump debit is about 41.9 percent.

Total plant auxiliary power is 16,007 kW (about 8 percent of generator output) and the net plant output is 193,034 kW.

The net plant heat rate and thermal efficiency are calculated to be 9,611 Btu/kWh and 35.5 percent respectively (HHV basis) for this case.

Carbon dioxide emissions are 385,427 lbm/hr or about 2.00 lbm/kWh on a normalized basis.

Table 2.1.12: Case-1 Overall Plant Performance Summary (Base Case)

	(Units)	CFB Air Fired (Case 1)
Auxiliary Power Listing		
Induced Draft Fan	(kW)	2285
Primary Air Fan	(kW)	2427
Secondary Air Fan	(kW)	1142
Fluidizing Air Blower	(kW)	920
Transport Air Fan	(kW)	n/a
Gas Recirculation Fan	(kW)	n/a
Coal Handling, Preperation, and Feed	(kW)	300
Limestone Handling and Feed	(kW)	200
Limestone Blower	(kW)	150
Ash Handling	(kW)	200
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400
Boiler Feed Pump	(kW)	3715
Condensate Pump	(kW)	79
Circulating Water Pump	(kW)	1400
Cooling Tower Fans	(kW)	1400
Steam Turbine Auxilliaries	(kW)	200
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719
Transformer Loss	(kW)	470
Subtotal	(kW)	16007
	(frac. of Gen. Output)	0.077
Air Separation Unit	(kW)	n/a
OTM System Compressor Auxiliary Power	(kW)	n/a
CO2 Removal System Auxiliary Power	(kW)	n/a
Total Auxilary Power	(kW)	16007
	(frac. of Gen. Output)	0.077
Output and Efficiency		
Main Steam Flow	(lbm/hr)	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147
OTM System Expander Generator Output	(kW)	
Steam Turbine Generator Output	(kW)	209041
Net Plant Output	(kW)	193034
	(frac. of Case-1 Net Output)	1.00
Simplified Boiler Efficiency (HHV)[1]	(fraction)	0.8946
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855

[1] Boiler Heat Output / Qcoal (HHV)
[2] Required for GPS Desicant Regen in Cases 2-7 and ASU in Cases 2-4

Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00
CO_2 Emissions		
CO_2 Produced	(lbm/hr)	385427
CO_2 Captured	(lbm/hr)	0
Fraction of CO2 Captured	(fraction)	0.00
CO_2 Emitted	(lbm/hr)	385427
Specific CO_2 Emissions	(lbm/kwhr)	2.00

2.2. Case–2: Oxygen Fired CFB with Moving Bed Heat Exchanger and CO_2 Capture

This section describes a power plant comprised of oxygen-fired Circulating Fluidized Bed (CFB) boiler, a Moving Bed Heat Exchanger (MBHE), a cryogenic type Air Separation Unit (ASU), and a subcritical steam plant with reheat (1,800 psia / 1,000 ^{o}F / 1,000 ^{o}F). The plant is designed to produce a flue gas having a high concentration of CO_2. This stream is then further processed in a Gas Processing System to produce a CO_2 product suitable for usage or sequestration. The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

The basic CO_2 capture concept behind Case-2 is to replace combustion air with oxygen thereby creating a high CO_2 content flue gas stream that can be further processed into a high purity CO_2 end product for various uses or sequestration. The replacement of combustion air with high purity oxygen can, as an additional benefit, significantly reduce the gas flow throughout the Boiler Island equipment. In this case a cryogenic Air Separation Unit (ASU) supplies the oxidant (99 percent pure oxygen) for the combustion of coal rather than direct utilization of ambient air as was done in Case-1. Since the size and cost of much of the equipment contained within the Boiler Island is strongly gas flow dependent (combustor, cyclones, backpass heat exchangers, air heater, fans, ductwork, baghouse, etc.) significant cost savings are anticipated for the Boiler Island of this concept as compared to the comparable air-fired CFB (Case-1).

A brief performance summary for this plant reveals the following information. The Case-2 plant produces a net plant output of about 134 MW. The net plant heat rate and thermal efficiency are calculated to be 13,546 Btu/kWh and 25.2 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 0.18 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.2.5.

Agglomeration Concerns:

When firing with oxygen in coal fired equipment it is critical to be able to control temperature in the combustor region in order to avoid potential problems. High combustor temperatures in CFB units can lead to agglomeration of the bed. With oxygen firing in a Circulating Fluid Bed (CFB), agglomeration of bed material is a major operational concern. If the ash particles are raised to a high enough temperature they will become sticky and large masses may form, requiring the unit to be shut down and cleaned out. Therefore, fuels containing ash with low fusibility temperatures are more challenging to burn than high fusibility temperature fuels.

In pulverized coal (PC) units, high furnace temperatures can cause uncontrollable slagging and or fouling which also can require the unit to be shut down to correct. In order to avoid uncontrollable slagging and fouling in oxygen fired pulverized coal and Stoker units, large quantities of recirculated flue gas are used to maintain temperatures throughout the unit at acceptable levels. The quantity of recirculated gas typically proposed for O_2 fired PC boilers in order to avoid these problems is that amount which makes the flue gas to coal flow ratio equivalent, or nearly equivalent, to that of air firing. Because of the large quantities of recirculated flue gas required, an oxygen fired pulverized coal unit is actually slightly more costly than the comparable air fired pulverized coal unit due to the additional cost of the gas recirculation system (additional ductwork, fans, and controls).

Similarly, it is possible in an oxygen fired CFB system, using the proper amount of recirculated gas, to obtain bed conditions within the combustor that are very similar to air firing. Systems can be designed to have the same flue gas flow to coal flow ratio as with air firing and the same volumetric fraction of oxygen in the flue gas leaving the bed as that of the air fired case. This scenario would represent the most conservative oxygen fired condition and is very similar to what is done in oxygen fired pulverized coal systems as described above. Because the flue gas flow to coal flow ratio would be the same as with air firing in this scenario, the size and cost of the Combustor, Cyclone, and Convection pass heat exchanger components and other equipment would also be nearly identical to the air fired case.

A potentially significant advantage for oxygen fired CFB type combustors as compared to oxygen fired pulverized coal fired or Stoker fired counterparts is the capability to reduce flue gas flow for a given coal input while maintaining combustor temperature. This capability is available to CFB units because there is an additional controllable variable available for CFB systems. Because within the CFB process (both air and oxygen fired) there is a large stream of recirculated cooled solids returning to the combustor from the external heat exchanger, the bed temperatures can theoretically be maintained at required levels in both the air and oxygen fired scenario thus avoiding any potential agglomeration. Combustor temperature can be maintained when firing with oxygen by simply cooling a greater portion of the large solids recirculation loop leaving the cyclone. Equipment designs developed for this case shows this type of system is technically feasible. Cost analysis, presented later in this report (Section 3), clearly quantifies the benefit of the reduced gas flow used for this case.

An extremely wide variety of fuels have been successfully used in ALSTOM-designed air-fired CFB units. Table 2.2.1 shows of a cross section of such fuels. They cover the complete coal rank spectrum from brown coal to anthracite, plus other waste fuels (e.g., delayed petroleum coke, anthracite culm). Table 2.2.1 demonstrates the extreme fuel-flexibility of air-fired CFB's, as they range in as-received ash contents from less than 1 percent to more than 39 percent, and higher heating values from 6,100 Btu/lbm to 12,960 Btu/lbm, and in moisture contents from 3 to 31 percent. These fuels also cover a wide range of sulfur contents, from less than 1 to more than 6 percent, indicating, correspondingly, high flexibility of CFB's in firing fuels ranging from low to high slagging potentials.

Table 2.2. 1: Range of Fuels Fired in Air Fired CFB's

Quantity	Sub-Bituminous Coal	Brown Coal	Lignite	High Volatile Bituminous Coal	High Volatile Bituminous Coal	Medium Volatile Bituminous Coal	Medium Volatile Bituminous Coal	Delayed Petroleun Coke	Anthracite Culm	Antracite
Proximate & Ultimate Analyes, Wt.%										
Moisture	11.6	24.6	31.0	6.5	2.4	4.7	5.4	14.4	14.6	3.3
Ash	25.5	24.7	14.8	11.0	15.8	14.0	13.0	0.2	33.9	39.1
Volatile Matter	40.4	26.5	24.5	36.0	35.6	18.5	17.5	10.6	6.5	4.1
Fixed Carbon (Diff.)	22.6	24.2	29.7	46.5	46.2	62.8	64.1	74.8	45.0	53.5
Hydrogen	2.9	3.0	3.0	4.7	4.4	3.7	3.6	3.0	0.8	0.3
Carbon	43.3	35.0	40.6	65.5	66.1	70.2	74.2	74.2	45.7	52.9
Nitrogen	1.0	0.5	0.7	1.2	1.2	1.5	1.5	1.0	0.6	0.2
Sulfur	3.5	1.1	0.7	0.6	4.3	1.8	1.7	6.3	0.6	0.6
Oxygen (Diff.)	12.3	11.1	9.2	10.6	5.8	4.1	0.5	0.9	3.8	3.6
HHV, Btu/lb, As Received	5785	6121	6856	11675	12067	12221	12256	12963	7000	8152
Carbon, DAF	**68.7**	**69.1**	**74.9**	**79.3**	**80.8**	**86.4**	**90.9**	**86.9**	**88.7**	**91.8**
Volatile Matter, DAF	**64.1**	**52.3**	**45.2**	**43.6**	**43.5**	**22.8**	**21.5**	**12.4**	**12.6**	**7.1**

DAF = Dry-Ash-Free

Bed temperatures in these units are maintained at acceptable values by recirculation of cooled solids from the external heat exchanger to the combustor. Additionally, heat is transferred from the gas-solids mixture to the combustor waterwalls and other heat exchange surfaces located in the combustor. Proper balancing of the combustor heat absorption and the external heat exchanger heat absorption ensure the desired combustor bed temperature is achieved. The large inventory of solids within the bed with its associated large thermal inertia along with the high degree of mixing help to limit local temperature excursions within the bed. These favorable characteristics inherent within the CFB process help to make the compact oxygen fired CFB concept of Case-2 viable with respect to the mitigation of potential agglomeration.

Oxygen Fired Bench Scale FBC Testing:

Tests were performed in Task-2 of this project (Bench Scale FBC Testing) to specifically address the agglomeration concerns associated with oxygen fired CFB systems. A wide range of operating conditions was investigated in the testing as described below. These conditions covered a significant range from traditional air firing (representing a baseline condition) to 70 percent oxygen firing. Higher than 70 percent oxygen levels were not tested due to equipment limitations. Additionally, tests with various intermediate levels of oxygen were also investigated in the testing.

One of the test facilities used by ALSTOM to characterize fuels and limestones for commercial CFB design purposes is a bench-scale (four-inch inner diameter) fluidized bed combustor (FBC). This facility was used to combustion test three fuels along with a limestone sample from the Base Case CFB plant. Details of these tests are presented in Volume II of this report. A brief summary of the fuel characteristics, test conditions, results/observations, and concluding remarks are given below:

Fuel Characteristics:

Three fuels were tested in Task-2. The Base Case CFB coal has nitrogen and sulfur contents of 1.7 percent and 2.4 percent on a dry-ash-free (DAF) basis, respectively; these values are typical of those of bituminous coals. These values, combined with high ash fusibility temperatures (I.T. = 2,400 ^oF and F.T. = 2,800 ^oF) place this coal in a low slagging potential range. The second coal tested (Illinois #6) has nitrogen and sulfur contents of 1.5 percent and 4.9 percent. These values, combined with low ash fusibility temperatures (I.T. = 2,070 ^oF and F.T. = 2,260 ^oF) place this coal in a high slagging potential range. The third fuel tested, Petroleum Coke, has nitrogen and sulfur contents of 4.0 percent and 1.5 percent (DAF). Furthermore, it has a very low ash content of 0.3 percent. These conditions are indicative of a fuel with low slagging potential.

Test Conditions:

All three fuels were burned in air and 30 percent O_2/70 percent CO_2 (by volume) mediums. Additionally, the Base Case CFB and Illinois #6 coals were burned in O_2/CO_2 mediums containing from 21 to 70 percent O_2, and from 21 to 50 percent O_2, respectively. Bed temperatures ranged from about 1,635 ^oF to 1,900 ^oF. The limestone sample from the Base Case CFB plant was used selectively to determine its impact on the flue gas desulfurization potential. All tests were carried out in high excess O_2 (with stoichiometric ratios ranging from about 2 to about 4.7). Table 2.2.2 summarizes these test conditions.

Table 2.2. 2: Test Matrix for Bench-Scale Fluidized Bed Combustion

Fuel	Series	Ca/S Mole Ratio	Combustion Gas Medium	Bed Temeperature, ^oF	Superficial Gas Velocity, ft/sec.	Stoichiometry, λ	Measurements/Observations:
Base Case CFB Coal	I	---	Air	1635-1908	1.8 - 3.3	2.0 - 4.7	❑ Ash slagging/sintering potentials as a result of increased oxygen concentrations in the combustion gas, and/or use of a slagging coal
	II	---	23%O_2 -70% O_2/in CO_2 Balance				
	III	3.5	Air				
		3.5	30% O_2/70% CO_2				❑ Flue gas desulfurzation potential by the limestone sample used
Illinois #6 Coal	I	---	Air				
	II	---	21% O_2 -50% O_2/in CO_2 Balance				❑ NOx emissions reduction by use of nitrogen-free O_2/CO_2 combustion mediums
	III	3.5	Air				
		3.5	30% O_2/70% CO_2				❑ Impact of various combustion mediums on: (1) overall fuel combustion efficiency; (2) unburned carbon emissions; and (3) CO emissions
Delayed Petroleum Coke	I	---	Air				
	II	---	30% O_2/70% CO_3				
	III	3.50	Air				
		3.50	30% O_2/70% CO_3				❑ Impact of fuel nature on all the parameters specified above

Results/Observations:

Testing the Base Case CFB coal in O_2/CO_2 mediums containing up to 70 percent O_2 caused, as expected, a bed temperature rise of up to about 250 ^oF. Nevertheless, it was possible to obviate bed slagging/de-fluidization problems as long as the bed was well fluidized. As a frame of reference, the superficial gas velocity had to be maintained at = 2.5 ft/sec., whereas for combustion testing in air and O_2/CO_2 mediums containing = 40 percent O_2 a superficial gas velocity of ~1.7 ft/sec. was sufficient to keep the bed fluidized at all times. The Illinois #6 coal, which has high slagging potential, was successfully burned in the FBC in O_2/CO_2 mediums containing up to 50 percent O_2. Allthough the test in 70 percent O_2/30 percent CO_2 was not run, due principally to time and financial constraints, there was no reason to believe that it would not have been executed successfully.

Concluding Remarks:

The test conditions used here in the FBC facility are very aggressive for two reasons: (1) there is no means of recycling particles to control the bed temperature; and (2) given the prevailing stoichiometric ratios, the furnace outlet O_2 concentration ranged from 13 to 51 percent (Dry basis).

On the other hand, the operating conditions of a commercial oxy-fuel fired CFB would be much less aggressive for the following reasons: (1) the bed temperature is closely controlled through judicious recycling of cooled bed materials; (2) the superficial gas velocity is maintained at about 18 ft/sec.; and (3) the O_2 concentration in the furnace would rapidly decline from its initial value of 70 percent, for Case 2, to about 3 percent at the outlet.

Based on these findings, it was preliminarily concluded that the choice of 70 percent O_2/30 percent recycled flue gas (i.e., ~CO_2) as a combustion medium for study Case 2 (New Compact O_2-Fired CFB) was plausible. This oxygen content was also utilized in Cases 3, 4 and 6. Pilot-scale testing in Phase II will evaluate the Case-2 concept, among other things. Based on the pilot-scale testing the concept presented in Case 2 will be either affirmed for further commercial development or modified beforehand.

2.2.1. Case 2 Boiler Island Process Description and Equipment

As discussed above, the basic concept for Case-2 is to minimize the amount of gas recirculation such that the size and cost of the above mentioned components, and other components, can be reduced significantly. If oxygen firing can be used in a CFB system with minimal gas recirculation, the flue gas flow to coal flow ratio can be reduced. This ratio was reduced from about 11.9 lbm gas/lbm coal with air firing (Case-1) to about 4.2 lbm gas/lbm coal for Case-2 with oxygen firing. This gas to coal ratio equates to about 70 percent oxygen by volume in the oxidant feed streams to the combustor. Since the size of much of the Boiler Island system components is strongly dependent on gas flow, significant size and cost savings are anticipated for these components.

This section describes the Boiler Island processes for Case-2 and includes a simplified process flow diagram (PFD), material and energy balance and equipment description. The equipment description includes only the major components of the Boiler Island.

2.2.1.1. Process Description and Process Flow Diagrams

Figure 2.2.1 shows a simplified process flow diagram for the Boiler Island of the Case-2 oxygen-fired CFB concept. This process description briefly describes the function of the major equipment and systems included within the Boiler Island. Selected mass flow rates (lbm/hr) and temperatures (^oF) are shown on this figure. Complete data for all state points are shown in Table 2.2.3. In this concept coal or another high carbon content fuel (Stream 1) is reacted with a preheated mixture of substantially pure oxygen and recirculated flue gas (Stream 18) in the Combustor section of the Circulating Fluidized Bed (CFB) system. The oxygen (Streams 16, 17, 18) is provided from a cryogenic Air Separation Unit (ASU).

The products of combustion, flue gas comprised of primarily CO_2 and H_2O vapor and un-reacted hot solids, flow through a cyclone, or another type of particulate removal device, where most of the hot solids are removed and recirculated to the combustor. The temperature in the combustor is controlled to the proper level by properly splitting the flow of hot recirculated solids leaving the cyclone, between an uncooled stream which flows directly back to the Combustor and a stream flowing through the External Heat Exchanger where the solids are cooled before returning to the Combustor. Exchanging heat with the power cycle working fluid cools the hot solids in the External Heat Exchanger.

Draining hot solids through water-cooled fluidized bed ash coolers (Stream 21) controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash coolers is feedwater from the final extraction feedwater heater of the steam cycle.

The flue gas leaving the cyclone (Stream 3) is cooled in heat exchanger sections located in the convection pass of the system, also by exchanging heat with the power cycle working fluid. The flue gas leaving the convection pass heat exchanger sections (Stream 5) is further cooled in an Oxygen Heater. The oxygen stream leaving the Air Separation Unit (Stream 16) is mixed with a small stream of recirculated flue gas (Stream 15) and the mixture is preheated in the Oxygen Heater. The quantity of recirculated flue gas is only the amount necessary to provide proper fluidization for the bed.

The flue gas leaving the Oxygen Heater (Stream 6) is cleaned of fine particulate matter in the baghouse and further cooled in a Parallel Feedwater Heater (PFWH) by transferring heat to feedwater in parallel with extraction feedwater heaters. Finally, a Gas Cooler is used to cool the gas before the flue gas enters the Induced Draft (ID) Fan (Stream 10). The gas cooler is used to cool the flue gas to the lowest temperature possible before recycling to minimize the power requirements for the draft system (induced draft fan, fluidizing air blower, and gas recirculation fan) and the product gas compression system. Some H_2O vapor is condensed in the Gas Cooler. The flue gas leaving the ID Fan (Stream 11), comprised of mostly CO_2 and H_2O vapor, is split with most of the flue gas going to the product stream (Stream 12) for further processing and the remainder recirculated to the CFB system.

By using oxygen instead of air for combustion, and by minimizing the amount of recirculated flue gas, the size and cost of the Combustor, Cyclone, and Convection Pass Heat Exchanger sections and other equipment can be reduced as compared to many other concepts for CO_2 capture with CFB systems.

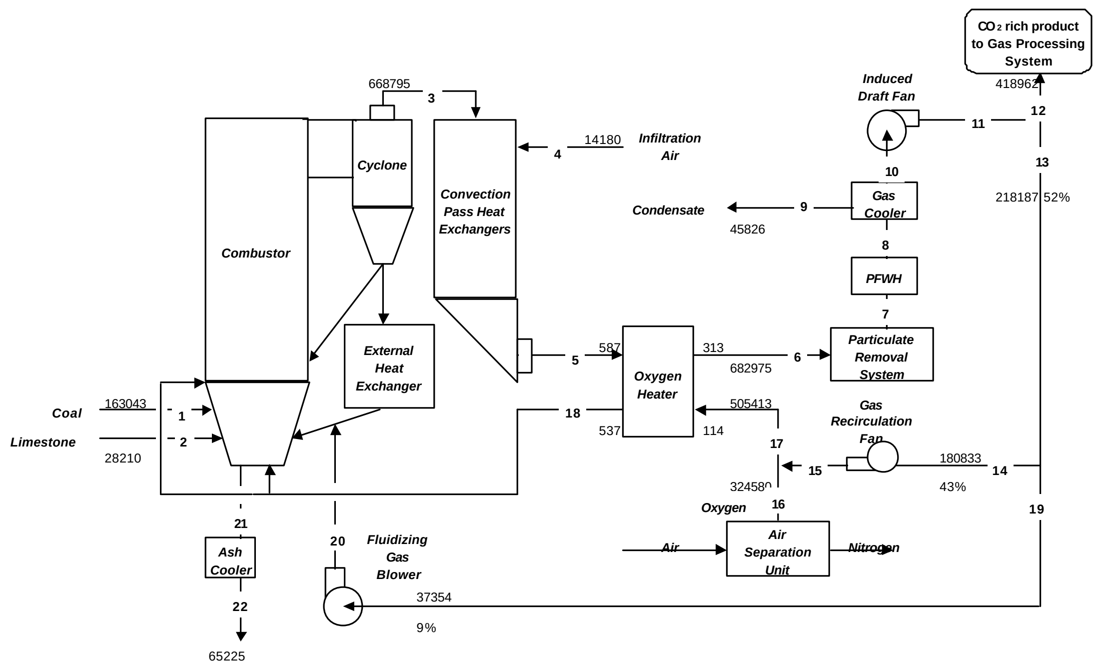


Figure 2.2. 1: Case-2: Simplified Boiler Island Gas Side Process Flow Diagram

2.2.1.2. Material and Energy Balance

Table 2.2.3 shows the Boiler Island material and energy balance for Case-2. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-2 simplified PFD for the Boiler Island (Figure 2.2.1). This performance was calculated at MCR conditions for this unit.

The MCR condition is defined as high-pressure turbine inlet conditions of 1,400,555 lbm/hr, 1,815 psia, and 1,000 ^{o}F and intermediate-pressure turbine inlet conditions of 1,305,632 lbm/hr 469 psia 1,000 ^{o}F. These conditions were also used for the Base Case. The boiler was fired with enough oxygen to leave about 3 percent by volume in Stream 3, the same as for the Base Case. This oxygen requirement results in a stoichiometry of about 1.05 for Case-2. The resulting boiler efficiency calculated for Case-2 was 94.12

percent (HHV basis) with an oxygen heater exit gas temperature of 313 °F and the PFWH exit gas temperature of 136 °F.

Table 2.2. 3: Case-2 Boiler Island Gas Side Material and Energy Balance

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11
O2	(Lbm/hr)	5152		17814	3246	21060	21060	21060	21060		21060	21060
N2	"	2380		14156	10753	24909	24909	24909	24909		24909	24909
H2O	"	6505		65228	182	65409	65409	65409	65409	45826	19583	19583
CO2	"			570438		570438	570438	570438	570438		570438	570438
SO2	"			1159		1159	1159	1159	1159		1159	1159
H2	"	5821										
Carbon	"	101168										
Sulfur	"	3815										
CaO	"											
CaSO4	"											
CaCO3	"		26800									
Ash	"	38201	1411									
		Coal	Limestone	Flue Gas	Infiltration Air	Flue Gas	Flue Gas	Flue Gas	Flue Gas	Condensate	Flue Gas	Flue Gas
Total Gas	(Lbm/hr)			668795	14180	682975	682975	682975	682975		637149	637149
Total Solids	"	163043	28210									
Total Flow	"	163043	28210	668795	14180	682975	682975	682975	682975	45826	637149	637149
Temperature	(Deg F)	80	80	1600	80	587	313	313	136	100	100	112
Pressure	(Psia)	14.7	14.7	14.7	14.7	14.4	14.2	13.8	13.7	14.7	13.6	14.7
hsensible	(Btu/lbm)			439.702	0.000	128.882	56.045	56.045	12.794	19.960	4.228	6.826
Chemical	106 Btu/hr	1805.540										
Sensible	106 Btu/hr	0.0	0.000	294.071	0.000	88.023	38.277	38.277	8.738	0.915	2.694	4.349
Latent	106 Btu/hr	0.0	0.000	68.489	0.191	68.680	68.680	68.680	68.680	0.000	20.562	20.562
Total Energy(1)	106 Btu/hr	1805.540	0.000	362.560	0.191	156.703	106.957	106.957	77.418	0.915	23.256	24.911

Constituent	(Units)	12	13	14	15	16	17	18	19	20	21	22
O2	(Lbm/hr)	13848	7212	5977	5977	321334	327311	327311	1235	1235		
N2	"	16379	8530	7069	7069	3246	10315	10315	1460	1460		
H2O	"	12877	6706	5558	5558		5558	5558	1148	1148		
CO2	"	375095	195342	161899	161899		161899	161899	33443	33443		
SO2	"	762	397	329	329		329	329	68	68		
H2	"											
Carbon	"										2023	2023
Sulfur	"											0
CaO	"										9009	9009
CaSO4	"										14581	14581
CaCO3	"											0
Ash	"										39612	39612
		Flue Gas	Recirc Gas	Recirc Gas	Recirc Gas	Oxygen	Oxy + Recirc	Oxy + Recirc	Grease Gas	Grease Gas	Hot Ash Drain	Cool Ash Drain
Total Gas	(Lbm/hr)	418962	218187	180833	180833	324580	505413	505413	37354	37354		
Total Solids	"							0			65225	65225
Total Flow	"	418962	218187	180833	180833	324580	505413	505413	37354	37354	65225	65225
Temperature	(Deg F)	112	112	112	140	100	114	537	112	195	1600	520
Pressure	(Psia)	14.7	14.7	14.7	17.4	17.4	17.4	17.19	14.7	23.7	14.7	14.7
hsensible	(Btu/lbm)	6.826	6.826	6.826	12.934	4.407	7.458	105.883	6.826	25.005	407.729	95.391
Chemical	106 Btu/hr)										28.515	28.515
Sensible	106 Btu/hr	2.860	1.489	1.234	2.339	1.431	3.770	53.515	0.255	0.934	26.594	6.222
Latent	106 Btu/hr	13.521	7.041	5.836	5.836	0.000	5.836	5.836	1.206	1.206	0.000	0.000
Total Energy(1)	106 Btu/hr	16.381	8.531	7.070	8.175	1.431	9.606	59.351	1.460	2.140	55.109	34.737

Notes:
(1) Energy Basis; Chemical based on Higher Heating Value (HHV); Sensible energy above 80F; Latent based on 1050 Btu/Lbm of water vapor

Figure 2.2.2 shows a general comparison of overall boiler performance between the air firing of Case-1 and the oxygen firing of Case-2. The boiler performance is compared in terms of the heat absorption distribution within the boiler components. Excluding the parallel feedwater heater (PFWH) heat absorption of Case-2, the total heat absorption is the same in both Case-1 and Case-2. The primary differences occur in the Convection Pass, Combustor and External Heat Exchanger. The Convection Pass heat absorption for Case-2 is about 40 percent of the Case-1 value due to the reduced gas flow. Similarly, the Combustor heat absorption for Case-2 is about 32 percent of the Case-1 value due to the reduction in size. Finally, the External Heat Exchanger heat absorption for Case-2 is about 3.5 times greater than the Case-1 value. Because of the very large heat duty of the external heat exchanger of Case-2 (about 70 percent of the total duty as compared to about 20 percent for Case-1) a moving bed design was selected based on cost considerations. The Case-1 external heat exchanger was a bubbling fluidized bed design.

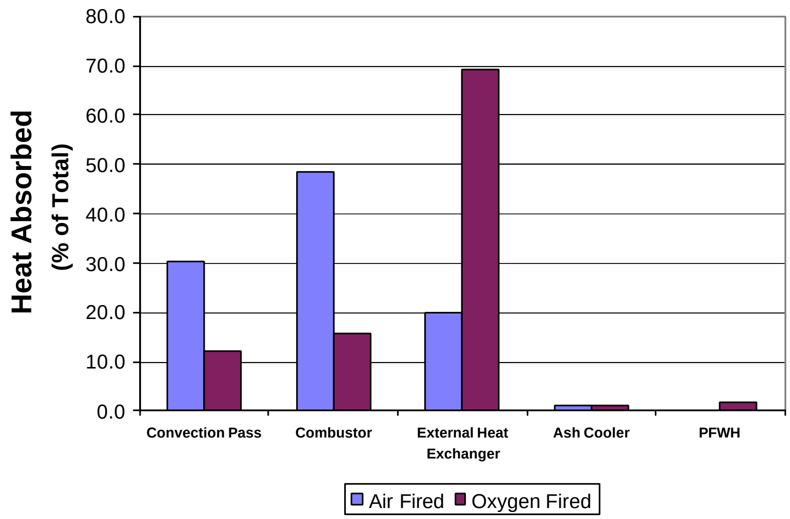


Figure 2.2. 2: CFB Boiler Heat Absorption Comparison (Air and O_2 Firing)

A moving bed external heat exchanger design that was implemented in this case provides several advantages over a bubbling fluidized bed. A significant advantage of the moving bed is that a higher temperature difference is obtained between the bed material and the steam cycle working fluid thus reducing the quantity of heat exchanger pressure part surface that is required. This occurs because the moving bed can be designed as a counterflow heat exchanger. The bubbling fluid bed on the other hand is more of a “stirred” heat exchanger where the bed material is at a “stirred temperature”. The “stirred temperature” is much lower than the inlet solids temperature in the moving bed. Additionally, the moving bed does not require any fluidizing air, fluidizing nozzles, and fluidizing air piping thus providing a simpler system. With these advantages, the moving bed allows for a much more compact and less expensive design.

2.2.1.3. Boiler Island Equipment

This section describes major equipment included in the Boiler Island for Case-2. The major components included in the Boiler Island include the combustor, ash coolers, fuel feed system, fuel silos, sorbent feed system, sorbent silo, cyclones, seal pots, external moving bed heat exchanger (MBHE), convection pass, superheater, reheater, economizer, oxygen heater, baghouse, parallel feedwater heater (PFWH), gas cooler, and draft system.

Figures 2.2.3 and 2.2.4 show a general arrangement drawing of the Case-2 CFB boiler. The plan area for the Case-2 Boiler Island is about 52 percent of that for Case-1. Similarly, the building volume for Case-2 is about 56 percent of that for Case-1. The complete Boiler Island Equipment List for Case-2 is shown in Appendix I. Appendix II shows several additional drawings of the Boiler (key plan view, boiler plan view, side elevation, and various sectional views).

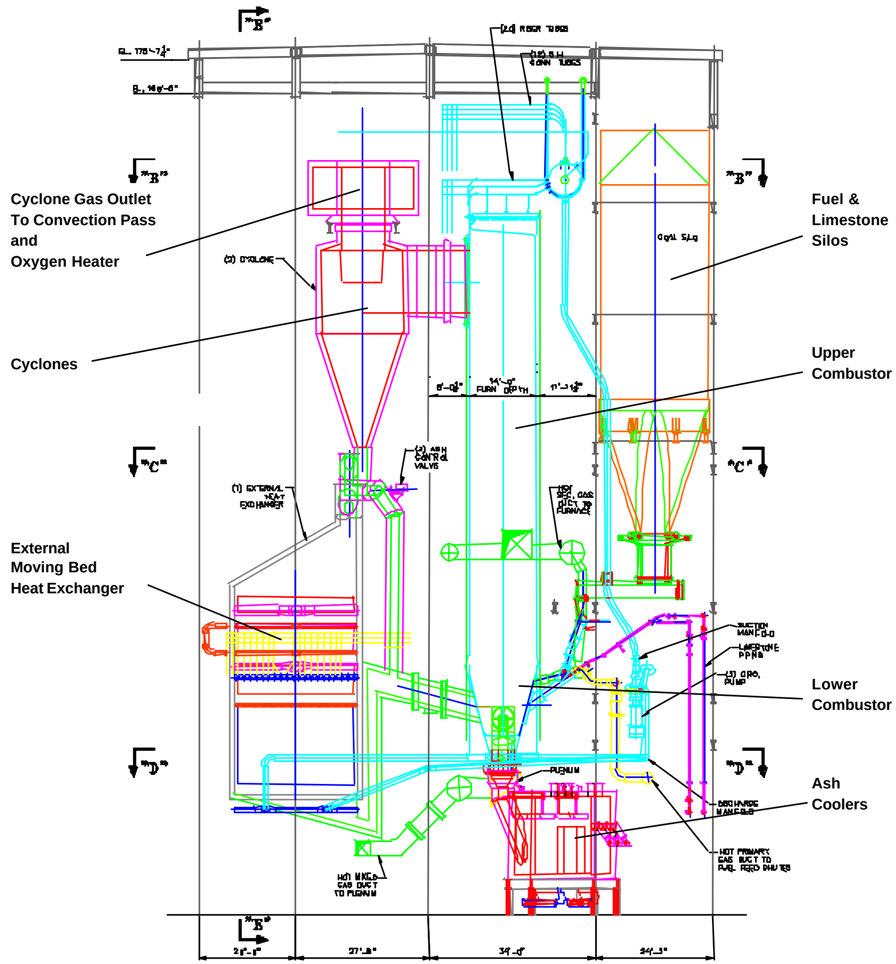


Figure 2.2. 3: Case-2 Boiler Island General Arrangement Drawing – Side Elevation

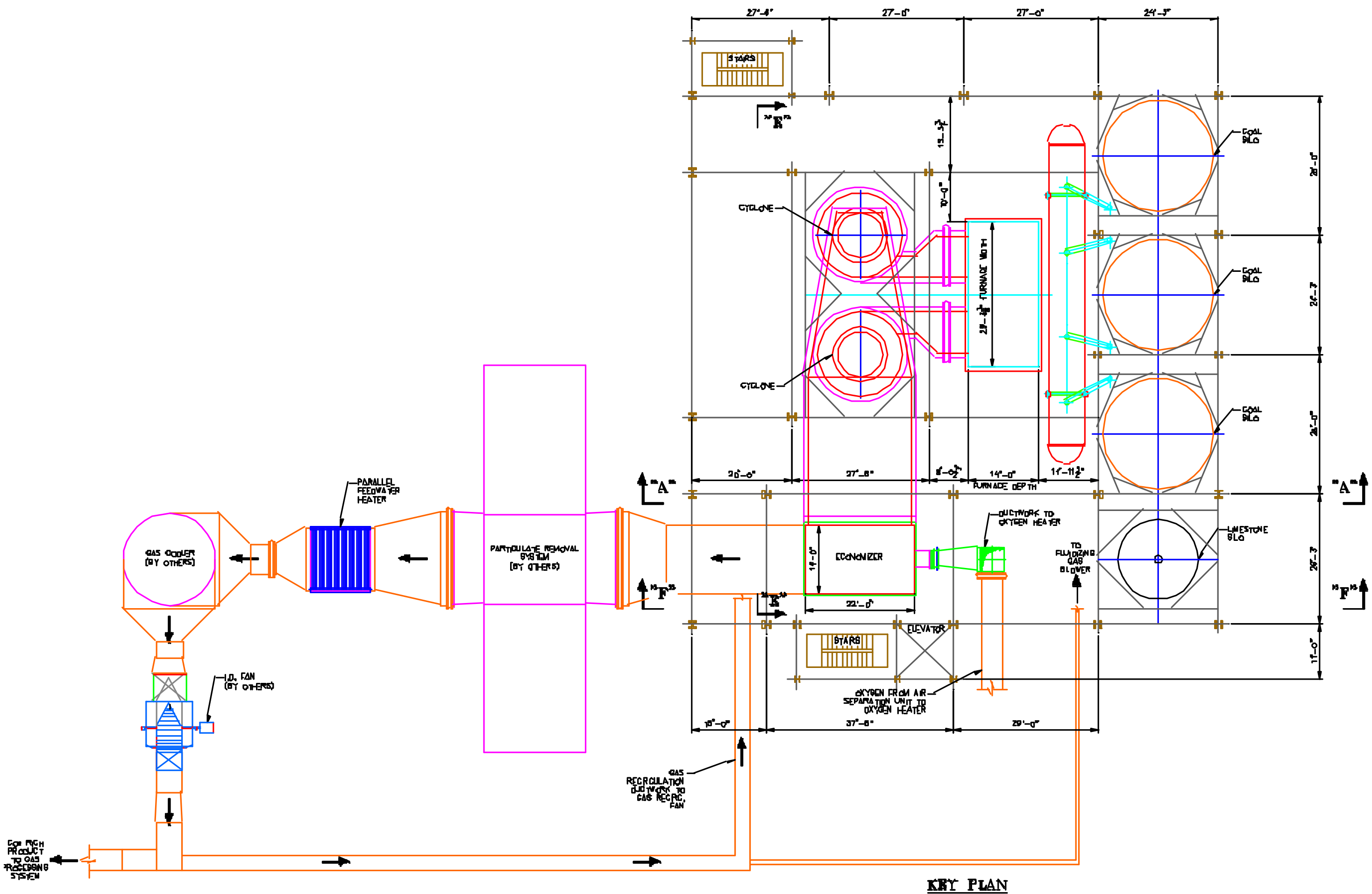


Figure 2.2. 4: Case-2 Boiler Island General Arrangement Drawing - Plan View

Combustor:

The combustor size is reduced significantly for Case-2. The combustor for Case-2 is about 29 ft wide, 14 ft deep and 100 ft high. Thus, the combustor plan area for Case-2 is about 30 percent of the Case-1 plan area for nearly equivalent fuel heat input quantities. Because of the relatively small combustor size, evaporator panels are not used in the Case-2 combustor as was done for Case-1. Crushed fuel, sorbent, and recycle solids are fed to the lower portion of the combustor. Primary “air” (actually a mixture of about 65 percent by weight oxygen and 35 percent recycled flue gas) is fed to the combustor bottom through a grid plate with secondary “air” supplied higher up in the lower combustor region. The combustor is constructed in the same fashion as the Case-1 combustor. The lower combustor region is a tapered rectangular section formed from fusion welded waterwall tubing with a refractory lining. Combustion occurs throughout both the lower and upper combustor, which are filled with bed material. The upper combustor section is a rectangular straight walled section formed from fusion welded waterwall tubing. The combustor is cooled with continuous forced circulation of water from the steam drum. Combustor bed temperature is maintained at an optimum level for sulfur capture and combustion efficiency by balancing heat absorption in the combustor walls and in the MBHE. No additional heat-absorbing surface (i.e. evaporator panels as were used in Case-1) is included in the combustor for Case-2.

Fuel Feed System:

The fuel feed system for Case-2 is very similar to the system used for Case-1. It is designed to transport prepared coal from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, fuel feeders, feeder isolation valves, and fuel piping to the furnace. In Case-2 only three feed points are used as compared to six in Case-1 because of the reduced combustor plan area.

Sorbent Feed System:

The limestone feed system for Case-2 is the same as for Case-1. The limestone feed system pneumatically transports prepared limestone from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, rotary feeders, blower, and piping from the blower to the furnace injection ports.

Ash Coolers:

The ash cooler design for Case-2 is the same as for Case-1 as the ash flow is nearly identical to Case-1. Draining hot solids through 2 water-cooled ash coolers controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is provided by feedwater from the final extraction feedwater heater of the steam cycle. The heated water leaving the ash cooler is then combined with water from the economizer located in the convection pass to feed the steam drum.

Cyclones:

Flue gas and entrained solids exit the upper combustor and enter the cyclones. Two relatively small cyclones were selected for Case-2. Although a single cyclone could have been selected for this case because of the reduced gas flow resulting from the oxygen firing, two smaller cyclones were selected to reduce overall unit height. The cost savings associated with a height reduction were greater than the cost increase for two cyclones instead of one. The cyclones are shaped like cylindrical cones constructed from steel plate with a multiple layer refractory lining. Solids are separated from the flue gas in the cyclones and fall into seal pots. Well over 99 percent of the entrained solids are captured in the cyclones. Two 17 ft diameter cyclones were selected for this case.

Seal Pots:

The seal pots for Case-2 are of the same design as in Case-1 although smaller in size since fewer solids are recirculated due to the reduced gas flow for Case-2. The seal pot is a device that provides a pressure seal between the combustor, which is at relatively high pressure (~ 40 inwg at the bottom), and the cyclone that is at near atmospheric pressure. The seal pot is a non-mechanical valve, which moves solids collected in the cyclone back to the combustor. The seal pot is constructed of steel plate with a multiple layer refractory lining with fluidizing nozzles located along the bottom to assist solids flow. Some of the solids flow directly from the seal pot back to the combustor while other solids are diverted through a plug valve, through the external Moving Bed Heat Exchangers (MBHE), and then back to the combustor.

Moving Bed Heat Exchanger:

The external heat exchanger for Case-2 is a single moving bed rather than 2 fluidized bubbling bed units as were used in Case-1. The moving bed heat exchanger is not fluidized and contains several immersed tube bundles, which cool the hot solids leaving the seal pot before the cooled solids return to the lower combustor. The tube bundles in the MBHE utilize spiral-finned tubes and include superheater, reheater and evaporator sections. Very high heat transfer rates are obtained in the MBHE due to the conduction heat transfer mechanism between the solids and tube. The MBHE is bottom supported and is constructed using steel plate refractory lined enclosure walls. It is rectangular in cross section with a hopper shaped bottom. The solids move through the bed by gravity at a design velocity of about 150 ft/hr. The cooled solids leaving the MBHE are feed to the combustor.

Convection Pass:

The flue gas leaving the cyclones is ducted into the Convection Pass. The convection pass for Case-2 is much smaller than for Case-1 and includes only an economizer section (3 banks) for heat absorption. The cross sectional area of the Case-2 convection pass is about 30 percent of the area required for Case-1. The convection pass is constructed similar to those used for pulverized coal firing with fusion welded steam cooled enclosure walls and fly-ash hoppers located at the bottom. Soot-blowers are used to keep the heat transfer surfaces clean.

Superheater:

The superheater is divided into two major sections, both of which are located in the MBHE. Saturated steam leaving the steam drum first cools the combustor roof and convection pass walls before supplying the horizontal low temperature and finishing superheater sections located in the external moving bed heat exchanger. The superheater sections are both horizontal sections comprised of spiral-finned tubing. There are no superheater banks located in the convection pass for Case-2. The steam leaving the low temperature section is piped to the de-superheater for superheater outlet temperature control and then to the finishing section. The steam leaving the finishing superheater is piped to the high-pressure turbine where it is expanded to reheat pressure and then returned to the reheat section located in the MBHE.

Reheater:

The reheater, located in the MBHE, is designed as a single section. The steam is supplied to the reheater inlet header from the de-superheating spray station, which is fed from the high-pressure turbine exhaust. The reheater is a horizontal section comprised of spiral-finned tubing and located between the superheat finishing section and the low temperature superheat section. There are no reheater banks located in the convection pass for Case-2. The steam leaving the de-superheating spray station is piped to the reheater located in the external moving bed heat exchanger. The steam leaving the reheater is returned to the intermediate pressure turbine where it continues its expansion through the intermediate and low-pressure turbines before being exhausted to the condenser.

Economizer:

The flue gas leaving the cyclones is then further cooled in an economizer section located in the convection pass. The economizer is comprised of four banks of horizontal tubes, which heats high-pressure boiler feedwater. The water exiting the economizer tube banks then cools the convection pass hanger tubes, which support the economizer sections, before it is supplied to the steam drum. The feedwater supplying the economizer is piped from the final extraction feedwater heater and the ash cooler.

Oxygen Heater:

A tubular regenerative oxygen heater is used to cool the flue gas leaving the economizer by heating both the primary and secondary "air" streams prior to combustion in the furnace. The primary and secondary "air" is actually a mixture of about 65 percent by weight oxygen with recycled flue gas.

Baghouse:

Particulate matter for Case-2 is removed from the cooled flue gas leaving the oxygen heater in a baghouse. The baghouse for Case-2 is much smaller than for Case-1 due to the reduced gas flow (about 30 percent of the Case-1 flow). The ash collected in the baghouse is supplied to the ash handling system.

Parallel Feedwater Heater:

The Parallel Feedwater Heater (PFWH) of Case-2 is used to recover additional heat in the steam cycle for this case as shown in Figure 2.2.5. The feedwater flow is in parallel with the bottom two extraction feedwater heaters included in the steam cycle. The PFWH is used because in Case-2 the gas temperature leaving the Oxygen Heater is significantly higher than the gas temperature leaving the Air Heaters of Case-1 (313 ^{o}F vs. 275 ^{o}F) and some of this energy can be economically recovered. This occurs because the ratio of air to gas in the air heater is higher in Case-1 than is the ratio of oxidant to gas in the oxygen heater of Case-2. This causes the Air heater of Case-1 to be more effective than the Oxygen Heater of Case-2. The PFWH heat exchanger is located in the flue gas stream following the baghouse and is constructed similarly to economizer heat exchanger banks used in Heat Recovery Steam Generator units. The tubes used are heavily finned since the gas is clean. The enclosure walls are constructed with insulated steel liners.

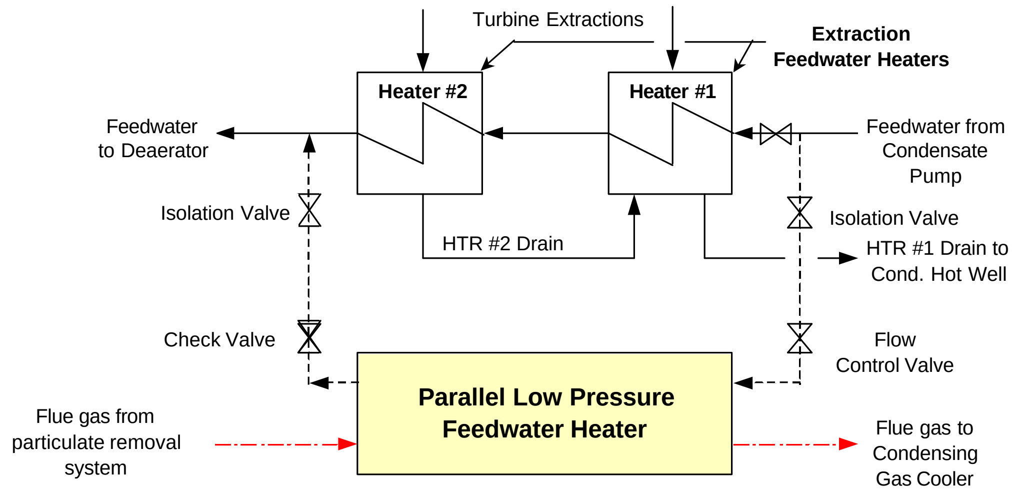


Figure 2.2. 5: Case-2 Parallel Feedwater Heater Arrangement

Gas Cooler:

The gas cooler of Case-2 is used to cool the flue gas leaving the PFWH to as low a temperature as possible in order to minimize the power requirements of the Draft System of the boiler and the Gas Processing System which produces the CO_2 product. The Gas Cooler is a direct contact, water spray type of system. Some of the water vapor contained within the flue gas also condenses out in this cooler. This cooler is designed to cool the flue gas to 100 ^{o}F. A detailed description of this process and equipment is contained in Section 2.2.2 as the Gas Cooler System is considered part of the Gas Processing System.

Draft System:

The flue gas is moved through the Boiler Island equipment with the draft system. The draft system includes the gas recirculation fan, the fluidizing gas blower, the induced draft (ID) Fan, the associated ductwork, and expansion joints. The induced draft fan, gas recirculation fan, and fluidizing gas blower are driven with electric motors and are controlled to operate the unit in a balanced draft mode with the cyclone inlets maintained at a slightly negative pressure (typically, -0.5 inwg).

2.2.2. Case-2 Gas Processing System Process Description and Equipment

The purpose of this system is to processes the flue gas stream leaving the oxygen-fired Boiler Island to provide a liquid CO_2 product stream of suitable purity for sequestration or usage.

The Case-2 CO_2 capture system is designed for about 94 percent CO_2 capture. Cost and performance estimates were developed for all the systems and equipment required to cool, purify, clean, compress and liquefy the CO_2, to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO_2 specification for EOR, given in Table 2.0.1, was used as the basis for the CO_2 capture system design.

A very low concentration of oxygen, in particular, is specified for meeting current pipeline operating practices, due to the corrosive nature of the oxygen. Hence, for Case-2, whereby the final CO_2 liquid product was found to contain about 11,400 ppmv of O_2, the design of the transport pipe to an EOR site for example would have to take this characteristic under consideration.

The nitrogen concentration specified in Table 2.0.1 is < 300 ppmv. It should be noted that according to Charles Fox of Kinder Morgan (Fox, 2002), this specification is very conservative as his company specifies a maximum nitrogen concentration of 4 percent (by volume) to control the minimum miscibility pressure. In Case 2 the nitrogen concentration in the liquid product was 11,800 ppmv. The exact reasoning behind the very low nitrogen specification listed in Table 2.0.1 is not clear.

2.2.2.1. Process Description

The following describes a CO_2 recovery system that cools and then compresses a CO_2 rich flue gas stream from an oxygen-fired CFB boiler to a pressure high enough so CO_2 can be liquefied. The resulting liquid CO_2 is passed through a CO_2 Stripper to reduce the N_2/O_2 content to a level that is optimum with respect to energy consumption. Then the liquid CO_2 is pumped to a high pressure (2,000 psig) so it can be economically transported for sequestration or usage. Pressure in the transport pipeline will be maintained above the critical pressure of CO_2 to avoid 2-phase flow. The overhead gas from the CO_2 Stripper which contains about 39 percent by volume CO_2, 36 percent N_2, and 25 percent O_2, is vented to atmosphere.

The key process parameters (pressures, temperatures, duties etc.) are shown in the material and energy balance tables shown in Section 2.2.2.3 and will not be repeated here except in selected instances.

Figure 2.2.6 (Refer to Section 2.2.2.2) shows the Flue Gas Cooling process flow diagram and Figure 2.2.7 shows the Flue Gas Compression and Liquefaction process flow diagram.

Flue Gas Cooling:

Please refer to Figure 2.2.6 (Drawing D 12173-02001-0).

The feed to the Gas Processing System is the flue gas stream that leaves the PFWH of the Boiler Island. At this point, the flue gas is near the dew point of H_2O. All of the flue gas leaving the boiler is cooled to 100 °F in Gas Cooler DA-101 which operates slightly below atmospheric pressure. A significant amount of water condenses out in this cooler. Excess condensate is blown down to the cooling water system. A single vessel has been provided for this cooler.

The Gas Cooler is configured in a packed tower arrangement where the flue gas is contacted with cold water in countercurrent fashion. Warm water from the bottom of the contactor is recycled back to the top of the contactor by Water Pump GA-101 after first cooling it in an external water cooled heat exchanger, Water Cooler EB-101 (plate and frame exchanger). The cooling water for this exchanger comes from the new cooling tower.

Because the flue gas may carry a small amount of fly ash, the circulating water is filtered in Water Filter FD-101A-E to prevent solids build-up in the circulating water. Condensate blowdown is filtered and is taken out downstream of the filter. However, the stream is not cooled and is split off before EB-101. Thus the heat load to the cooling tower is minimized.

From the Gas Cooler the gas stream then is boosted in pressure by the ID fan followed by a split of the gas into two streams. This design was developed to minimize the length of ducting operating at a slight vacuum and to minimize the temperature of the gas being recycled back to the boiler. The mass flow rate of the gas recirculation stream is about 52 percent of the flow rate of the product gas stream, which proceeds to the gas compression area. The recycle stream is sized to provide an oxygen content of about 70 percent by volume in the oxidant stream supplying the boiler. The Gas Cooler minimizes the volumetric flow rate to and the resulting power consumption of the Flue Gas Compression equipment located downstream and the gas recirculation fans in the Boiler Island.

Three-Stage Gas Compression System:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

The compression section, where CO_2 is compressed to 365 psig by a three-stage centrifugal compressor, includes Flue Gas Compressor GB-101. After the aftercoolers, the stream is then chilled in a propane chiller to a temperature of –21 °F. Note that both the trim cooling water and water for the propane condenser comes from the cooling tower. At this pressure and temperature, about 80 mole percent of the stream can be condensed. The flash vapors contain approximately 80 weight percent of the inlet oxygen and nitrogen, but also about 13.7 weight percent of the CO_2. Therefore, a rectifier tower has been provided to reduce the loss of CO_2 to an acceptable level (about 6 weight percent). Then the pressure of the liquid is boosted to 2,000 psig by CO_2 Pipeline Pump GA-103. This stream is now available for sequestration or usage.

The volumetric flow to the compressor inlet is about 69,000 ACFM and only a single frame is required. The discharge pressures of the stages have been balanced to give reasonable power distribution and discharge temperatures across the various stages. They are:

- 1st Stage 28 psig
- 2nd Stage 108 psig
- 3rd Stage 365 psig

Power consumption for this large compressor has been estimated assuming adiabatic efficiency of 75 percent.

The hot gas from each stage is first cooled in an air cooler to 120 ^{o}F (Flue Gas Compressor 1st/ 2nd / 3rd Stage Aftercooler EC-101/2/3) and then further cooled by a water-cooled heat exchanger to 95 ^{o}F (Flue Gas Compressor 1st/ 2nd Stage Trim Cooler EA-101/2). The flue gas compressor 3rd stage cooler (EA-103) cools the gas to 90 ^{o}F to reduce the size of the dryers. Due to their large size, many of these heat exchangers consist of multiple shells. Because of highly corrosive conditions, the process side of the coolers must be stainless steel.

Because the flue gas stream leaving DA-101 is saturated, some water condenses out in the three aftercoolers. The sour condensate is separated in knockout drums (FA-100/1/2/3) equipped with mist eliminator pads. Condensate from these drums is drained to the cooling tower or to waste water treatment. To prevent corrosion, these drums have a stainless steel liner.

Gas Drying:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

It is necessary to dry the CO_2 stream to meet the product specification. Flue gas leaving the 3rd stage discharge knockout drum (FA-103) is fed to Flue Gas Drier FF-101 A/B where additional moisture is removed. An alumina drier has been selected.

The performance of a fixed-bed drier improves as pressure increases. This favors locating the drier at the discharge of the compressor. However, as the operating pressure of the drier increases, so does the design pressure of the equipment. This favors low-pressure operation. But, at low pressure the diameter or number of the drier vessels grows, increasing the cost of the vessel. Having to process the recycle gas from the rectifier condenser cooling would also increase the diameter of the vessel. However, this is less than 10 percent of the forward flow. For this design the drier has been optimally located downstream of the 3rd stage compressor. The CO_2 Drier system consists of two vessels; FF-101 A/B. One vessel is on line while the other is being regenerated. Flow direction is down during operation and up during regeneration.

The drier is regenerated with the noncondensable vent gas from the rectifier after it exits the third stage discharge trim cooler in a simple once through scheme. During regeneration, the gas is heated in Regeneration Heater FH-101 before passing it through the exhausted drier. After regeneration, heating is stopped while the gas flow continues. This cools the bed down to the normal operating range. The regeneration gas and the impurities contained in it are vented to the atmosphere.

Regeneration of an alumina bed requires relatively high temperature and, because HP steam pressure may fluctuate, a gas-fired heater has been specified for this service.

Flue Gas Filter FD-102 has been provided at the drier outlet to remove any fines that the gas stream may pick up from the desiccant bed.

CO_2 Condensation and Stripping:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

From the CO_2 Drier, the gas stream is cooled down further to -21 ^oF with propane refrigeration in CO_2 Condenser EA-104 A-F. From EA-104 the partially condensed flue gas stream continuesonto CO_2 Rectifier DA-102.

At this pressure and temperature, 80 mole percent of the stream can be condensed. The flash vapors contain approximately 80 weight percent of the inlet oxygen and nitrogen, but also 12 weight percent of the CO_2. Therefore, as mentioned, a rectifier tower has been provided to reduce the loss of CO_2 to an acceptable level. The pressure of the liquid is boosted to 2,000 psig by CO_2 Pipeline Pump GA-103 for delivery to a sequestration or usage location.

The vapors in the feed to the rectifier contain the nitrogen and the oxygen that flashed from the liquid CO_2. To keep the CO_2 loss to the minimum, the rectifier also has an overhead condenser, CO_2 Rectifier Condenser EA-107. This is a floodback type condenser installed on top of the Rectifier. It cools the overhead vapor from the tower down to –48 ^oF. The condensed CO_2 acts as cold reflux in the CO_2 Rectifier.

Taking a slipstream from the inert-free liquid CO_2 from the Rectifier bottoms and letting it down to the Flue Gas Compressor 3rd stage suction pressure cools EA-107. At this pressure, CO_2 liquid boils at –55 ^oF thus providing the refrigeration necessary to condense some of the CO_2 from the Stripper overhead gas. The process has been designed to achieve at least 94 percent CO_2 recovery. The vaporized CO_2 from the cold side of EA-107 is fed to the suction of the Flue Gas Compressor 3rd stage.

Any system containing liquefied gas such as CO_2 is potentially subject to very low temperatures if the system is depressurized to atmospheric pressure while the system contains cryogenic liquid. If the CO_2 Rectifier (and all other associated equipment that may contain liquid CO_2) were to be designed for such a contingency, it would have to be made of stainless steel. However, through proper operating procedures and instrumentation such a scenario can be avoided and low temperature carbon steel (LTCS) can be used instead. Our choice here is LTCS. However, the condenser section will be made from stainless steel.

CO_2 Pumping and CO_2 Pipeline:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

The CO_2 product must be increased in pressure to 2,000 psig. A multistage heavy-duty pump (GA-103) is required for this service. This is a highly reliable derivative of an API-class boiler feed-water pump.

It is important that the pipeline pressure be always maintained above the critical pressure of CO_2 such that single-phase (dense-phase) flow is guaranteed. Therefore, pressure in the line should be controlled with a pressure controller and the associated control valve located at the destination end of the line.

Offgas:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

The vent gas from the CO_2 Rectifier overhead is at high pressure and there is an opportunity for power recovery using turbo-expanders. Because the gas cools down in the expansion process, there is also an opportunity for cold recovery. Heat recovery from the stream after let down via an expander was examined and it was determined that the amount of duty that could be recovered without the carbon dioxide in the stream freezing was small. Thus heat recovery could not be justified. The offgas leaves the Rectifier at –48 ^oF approximately. The refrigeration recovery to condense CO_2 was the best use for this cold since it also produces a reasonable temperature regeneration gas for the dryers.

2.2.2.2. Process Flow Diagrams

Two process flow diagrams are shown below for these systems:

- Figure 2.2.6 (Drawing D 12173-02001-0) Flue Gas Cooling PFD
- Figure 2.2.7 (Drawing D 12173-02002-0) CO_2 Compression and Liquefaction PFD

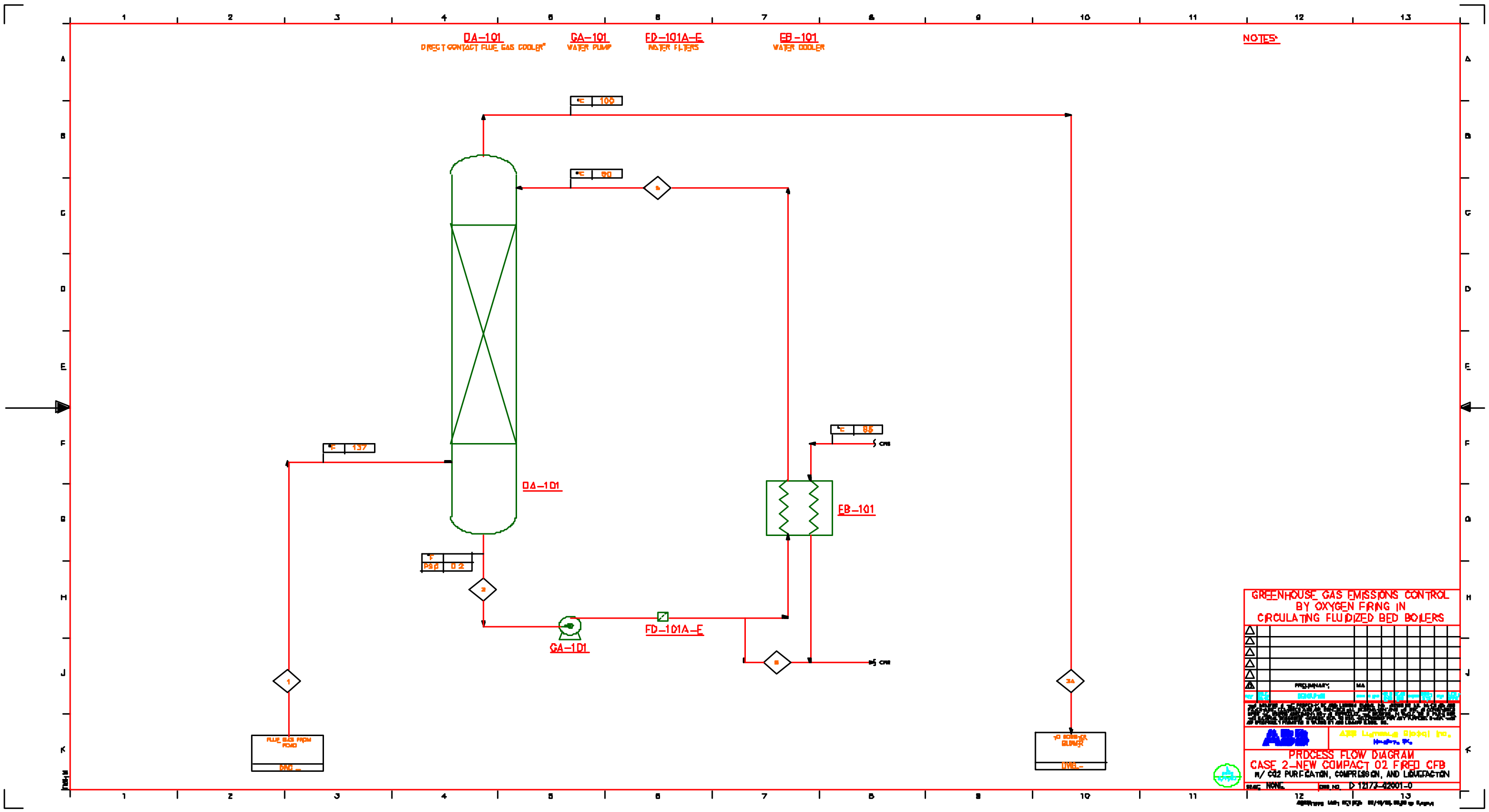


Figure 2.2. 6: Case-2 Flue Gas Cooling Process Flow Diagram

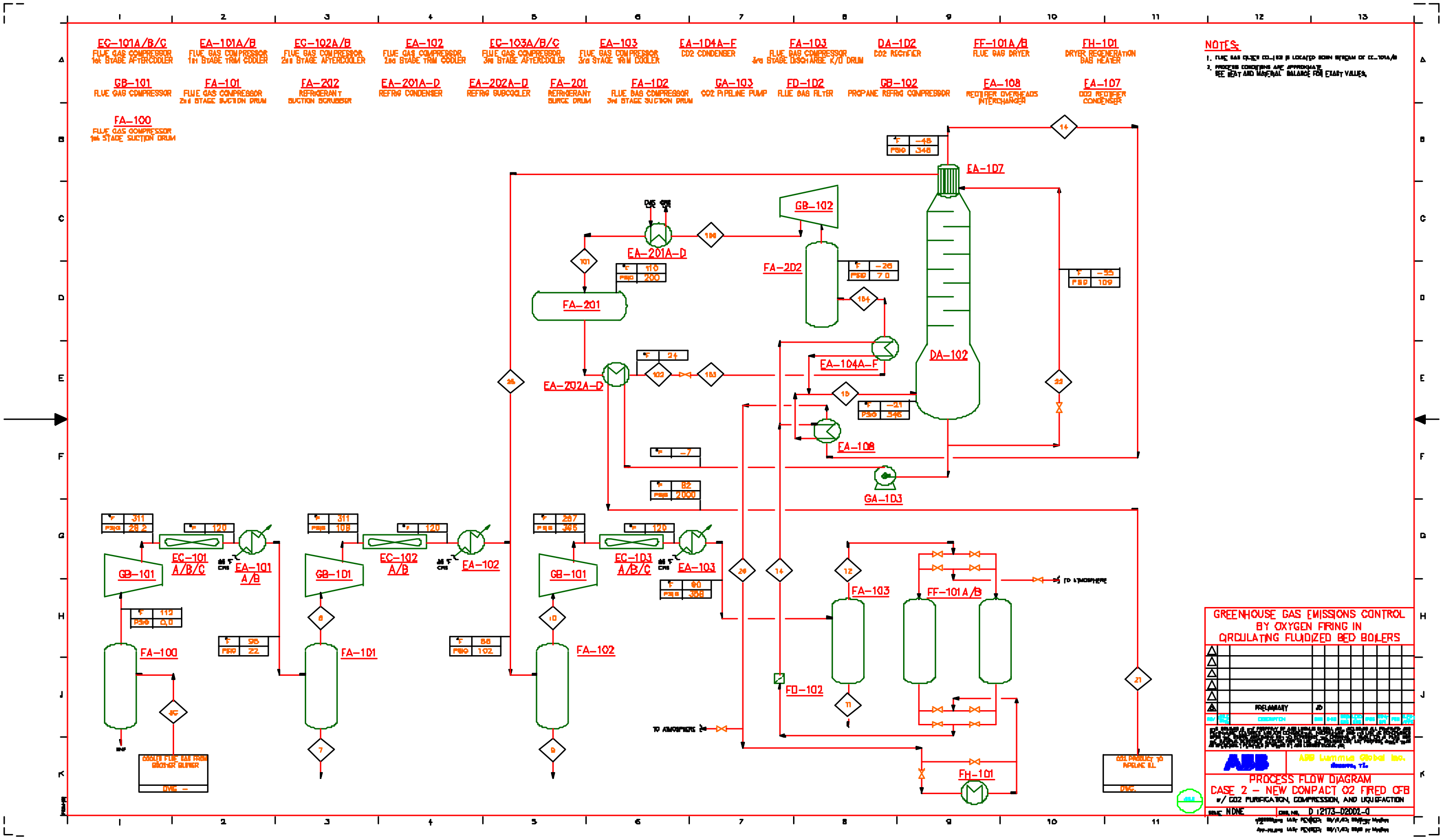


Figure 2.2. 7: Case-2 CO_2 Compression and Liquefaction Process Flow Diagram

2.2.2.3. Material and Energy Balance

Table 2.2.4 contains the overall material and energy balance for the Flue Gas Cooling System and the CO_2 Compression and Liquefaction System for Case-2 described above. It is based on 94 percent recovery of CO_2. Please refer to the Process Flow Diagrams shown in the previous section for stream numbers shown in this table.

Table 2.2. 4: Case-2 Gas Processing System Material & Energy Balance

STREAM NAME		To quench columns	From Quench columns	Excess water	From Large blowers	Quench water out	Quench water in	To liquefaction train	To boiler	First water KO	To 2nd stage	2nd water KO	To 3rd stage	Recycle from condenser
PFD STREAM NO.		**1**	**3a**	**6**	**3b**	**2**	**5**	**3c**	**3d**	**7**	**8**	**9**	**10**	**25**
VAPOR FRACTION	Molar	0.989	1.000	0.000		0.000	0.000	0.997	0.997	0.000	1.000	0.000	1.000	1.000
TEMPERATURE	°F	136.0	100	118		118	90	100	100	95	95	86	86	-48
PRESSURE	PSIA	13.7	13.5	55		14	45	13.3	13.3	34	34	117	117	117
MOLAR FLOW RATE	lbmol/hr	18,158	15,628	2,552		96,583	94,053	10,241	5,388	451	9,789	13	10,400	790
MASS FLOW RATE	lb/hr	682,970	637,380	45,992		1,740,600	1,695,000	418,390	218,990	8,141	410,250	231	441,530	34,533
ENERGY	Btu/hr	8.14E+07	6.84E+07	-3.59E+07		-1.36E+09	-1.37E+09	4.42E+07	2.42E+07	-6.54E+06	4.21E+07	-1.87E+05	4.30E+07	2.28E+06
COMPOSITON	Mol %													
CO2		71.38%	82.93%	0.02%		0.02%	0.02%	83.20%	83.20%	0.08%	87.03%	0.30%	89.32%	97.54%
H2O		4.90%	5.69%	0.00%		0.00%	0.00%	6.75%	6.75%	99.92%	2.46%	99.68%	0.59%	0.00%
Nitrogen		3.62%	4.21%	0.00%		0.00%	0.00%	5.71%	5.71%	0.00%	5.97%	0.00%	5.71%	1.18%
Ammonia		20.00%	7.05%	99.98%		99.98%	99.98%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		0.10%	0.12%	0.00%		0.00%	0.00%	4.22%	4.22%	0.00%	4.42%	0.00%	4.25%	1.14%
SO2		0.00%	0.00%	0.00%		0.00%	0.00%	0.12%	0.12%	0.01%	0.12%	0.02%	0.12%	0.14%
VAPOR														
MOLAR FLOW RATE	lbmol/hr	17,949.8	15,628.1	-		-	-	10,208.8	5,419.4	-	9,789.2	-	10,399.8	790.0
MASS FLOW RATE	lb/hr	679,220	637,380	-		-	-	417,820	219,560.0	-	410,250	-	441,530	34,533
STD VOL. FLOW	MMSCFD	163.48	142.33	-		-	-	92.98	49.4	-	89.16	-	94.72	7.20
ACTUAL VOL. FLOW	ACFM	138,840.00	114,250.00	-		-	-	69,181.03	45,069.0	-	28,427	-	8,390.08	447.36
MOLECULAR WEIGHT	MW	37.84	40.78	-		-	-	40.93	40.93	-	41.91	-	42.46	43.71
DENSITY	lb/ft³	0.08	0.09	-		-	-	0.10	0.10	-	0.24	-	0.88	1.29
VISCOSITY	cP	0.0145	0.0150	-		-	-	0.0150	0.0150	-	0.0155	-	0.0156	0.0114
LIGHT LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	-	-		-	-	-		-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-		-	-	-		-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-		-	-	-		-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-		-	-	-		-	-	-	-	-
DENSITY	lb/ft³	-	-	-		-	-	-		-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-		-	-	-		-	-	-	-	-
VISCOSITY	cP	-	-	-		-	-	-		-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-		-	-	-		-	-	-	-	-
HEAVY LIQUID														
MOLAR FLOW RATE	lbmol/hr	208.09	-	2,552.00		96,583	94,053	31.72		451.34	-	12.77	-	-
MASS FLOW RATE	lb/hr	3,750	-	45,991.83		1,740,600	1,695,000	571.64		8,141.38	-	231.26	-	-
STD VOL. FLOW	BPD	257	-	3,156		119,430	116,300	39		559	-	16	-	-
ACTUAL VOL. FLOW	GPM	7.62	-	92.76		3,510.88	3,378.10	1.14		16.26	-	0.46	-	-
DENSITY	lb/ft³	61.32	-	61.82		61.81	62.56	62.29		62.44	-	62.74	62.74	-
VISCOSITY	cP	0.4793	-	0.5654		0.5657	0.7606	0.6799		0.7185	-	0.8176	0.8176	-
SURFACE TENSION	Dyne/Cm	66.34	-	68.11		68.12	70.83	69.86		70.31	-	70.97	70.97	-

STREAM NAME		To drier	3rd water ko	From drier/ Condenser inlet	Condenser outlet	Non-condensable vent	Rectifier bottoms to condenser	CO2 to pipeline	Refrig compressor discharge	Refrig condenser out	Refrig subcooler out	Refrig to CO2 condenser	Refrig from CO2 condenser	Warm non condensable
PFD STREAM NO.		**12**	**11**	**14**	**15**	**24**	**22**	**21**	**100**	**101**	**102**	**103**	**104**	**26**
VAPOR FRACTION	Molar	1.000	0.000	1.000	0.206	1.000	0.130	0.000	1.000	0.000	0.000	0.225	1.000	1.000
TEMPERATURE	°F	90	90	90	-23	-48	-56	82	167	110	44	-26	-23	69
PRESSURE	PSIA	372	372	369	364	361	120	2,015	222	215	212	23	23	356
MOLAR FLOW RATE	lbmol/hr	10,366.29	33.47	10,337.65	10,087.65	1,358.89	790.00	8,209.08	9,100.00	9,100.00	9,100.00	9,100.00	9,100.00	1,358.89
MASS FLOW RATE	lb/hr	440,920	612	440,410	429,760	47,924	34,533	358,830	401,280	401,280	401,280	401,280	401,280	47,924
ENERGY	Btu/hr	4.01E+07	-4.83E+05	4.00E+07	-1.67E+07	3.58E+06	-2.22E+06	-2.76E+06	7.00E+07	6.77E+06	-1.13E+07	-1.13E+07	4.45E+07	4.97E+06
COMPOSITON	Mol %													
CO2		89.61%	0.87%	89.86%	89.86%	39.12%	97.54%	97.54%	0.00%	0.00%	0.00%	0.00%	0.00%	39.12%
H2O		0.28%	99.06%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		5.73%	0.00%	5.74%	5.74%	35.89%	1.18%	1.18%	0.00%	0.00%	0.00%	0.00%	0.00%	35.89%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%	0.00%
Oxygen		4.26%	0.00%	4.27%	4.27%	24.99%	1.14%	1.14%	0.00%	0.00%	0.00%	0.00%	0.00%	24.99%
SO2		0.12%	0.06%	0.12%	0.12%	0.00%	0.14%	0.14%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR														
MOLAR FLOW RATE	lbmol/hr	10,366.3	-	10,337.7	2,075.3	1,358.9	102.5	-	9,100.0	-	-	2,044.7	9,100.0	1,358.9
MASS FLOW RATE	lb/hr	440,920	-	440,410	79,499	47,924	4,282	-	401,280	-	-	90,163	401,280	47,924
STD VOL. FLOW	MMSCFD	94.41	-	94.15	18.90	12.38	0.93	-	82.88	-	-	18.62	82.88	12.38
ACTUAL VOL. FLOW	ACFM	2,410.72	-	2,425.14	373.67	241.80	56.04	-	3,758.16	-	-	6,717.44	30,084.4	342.06
MOLECULAR WEIGHT	MW	42.53	-	42.60	38.31	35.27	41.77	-	44.10	-	-	44.10	44.10	35.27
DENSITY	lb/ft³	3.05	-	3.03	3.55	3.30	1.27	-	1.78	-	-	0.22	0.22	2.34
VISCOSITY	cP	0.0165	-	0.0165	0.0145	0.0147	0.0117	-	0.0103	-	-	0.0066	0.0066	0.0183
LIGHT LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	-	-	8,012.37	-	687.48	8,209.08	-	9,100.00	9,100.00	7,055.35	-	-
MASS FLOW RATE	lb/hr	-	-	-	350,260	-	30,251	358,830	-	401,280	401,280	311,110	-	-
STD VOL. FLOW	BPD	-	-	-	28,971	-	2,506	29,682	-	54,230	54,230	42,045	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	659.06	-	52.70	895.70	-	1,735.91	1,533.76	1,091.12	-	-
DENSITY	lb/ft³	-	-	-	66.26	-	71.57	49.95	-	28.82	32.62	35.55	-	-
MOLECULAR WEIGHT	MW	-	-	-	43.72	-	44.00	43.71	-	44.10	44.10	44.10	-	-
VISCOSITY	cP	-	-	-	0.1641	-	0.2239	0.0559	-	0.0835	0.1198	0.1765	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	15.36	-	20.19	0.88	-	4.81	9.14	14.09	-	-
HEAVY LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	33.47	-	(0.00)	-	(0.00)	-	-	-	-	(0.01)	-	-
MASS FLOW RATE	lb/hr	-	611.58	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	42	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	1.21	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	62.84	62.84	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	0.7751	0.7751	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	70.20	70.20	-	-	-	-	-	-	-	-	-	-	-

2.2.2.4. Gas Processing System Utilities

The following tables define the cooling water, natural gas, and electrical requirements for the Gas Processing System.

Table 2.2. 5: Case-2 Gas Processing System Cooling Water and Fuel Gas Requirements

COOLING WATER

REV	Equipment TAG NO	SERVICE	No. Installed	DUTY MMBTU/HR	INLET TEMP, F	OUTLET TEMP, F	FLOWRATE LB/HR
	EA-101	FG Comp 1 stg trim cooler	1	7.09	85	103	393,939
	EA-102	FG Comp 1 stg trim cooler	1	3.73	85	103	207,071
	EA-103	FG Comp 1 stg trim cooler	1	3.82	85	103	212,121
	EA-201	Refrig Condenser	1	63.18	85	100	4,212,121
	EB-101	Water Cooler	1	50.00	85	105	2,500,000
		TOTAL COOLING WATER		127.82			7,525,253

FUEL GAS FUEL GAS VALUE BASIS: 930 BTU/SCF (LHV)

REV	Equipment TAG NO	SERVICE	ONLINE FACTOR	COMPR HP	HEAT RATE BTU/HP-HR	DUTY MMBTU/HR	EFFICIENCY %	FLOWRATE (Peak) MMSCFD	FLOWRATE (Peak) SCFH	FLOW (Avg) MMSCFD
	FH-101	Mole sieve regeneration	61%			4.60	80%	0.148	6,183	0.091
		TOTAL FUEL GAS				4.60		0.148	6,183	0.091

Table 2.2. 6: Case-2 Gas Processing System Electrical Requirements

Number of Trains	Item Number	Service	Number Operating per train	Power (ea) including 0.95 motor eff (kW)	Total all trains (kW)
1	EC-101	Flue Gas Compressor 1st Stage Aftercooler	1	78	78
1	EC-102	Flue Gas Compressor 2nd Stage Aftercooler	1	76	76
1	EC-103	Flue Gas Compressor 3rd Stage Aftercooler	1	67	67
1	GB-101	Stage 1	1	5434	5434
1		Stage 2	1	6531	6531
1		Stage 3	1	5997	5997
1	GB-102	Stage 1	1	4947	4947
1		Stage 2	1	2961	2961
1	GA-101	Water pump	1	130	130
1	GA-103	CO2 Pipeline pump	1	685	685
		Total			**26905**

2.2.2.5. Gas Processing System Equipment

The equipment list for the Gas Processing System is provided in Appendix I, Section 9.1.2.2.

2.2.3. Case-2 Air Separation Unit Process Description and Equipment

This section presents the process requirements for the warm end and cold box for the air separation plant. It will be designed to produce nominally 4000 tons per day (TPD) of Oxygen. This same design was also used for Case 3 and Case 4 Air Separation Units. Case-3 is Identical to Case-2 and Case-4 operates at a slightly higher output level.

Assumptions:

The following basic assumptions have been used in this design:

- No special low voltage starting equipment.
- Design based on customer inquiry dated 10/11/02 from Alstom Power.

Design Basis:

The ambient conditions presented in Table 2.2.7 below were used to evaluate performance and to generate the utility summary.

Table 2.2. 7: Ambient Conditions

Item	Value	Units
Barometric Pressure	14.7	psia
Dry Bulb Temperature	80	°F
Hot Day Temperature	95	°F
Cold Day Temperature	20	°F
Wet Bulb Temperature	52	°F
Cooling Water Temperature	90	°F

Production Rates and Purities:

The production rate indicated in Table 2.2.8 below is the net flow-rate from the Air Separation Unit's Cold Box.

Table 2.2. 8: Production and Purity

Plant Site	Oxygen Flow	Pressure @ B/L	Purity
	(Metric T/D Contained O_2)	(psia)	(%O_2)
Southeast Texas	3,568	18.0	99.0

2.2.3.1. Process Description and Process Flow Diagrams

The process description below refers to the Process Flow Diagram shown below in Figure 2.2.8 below.

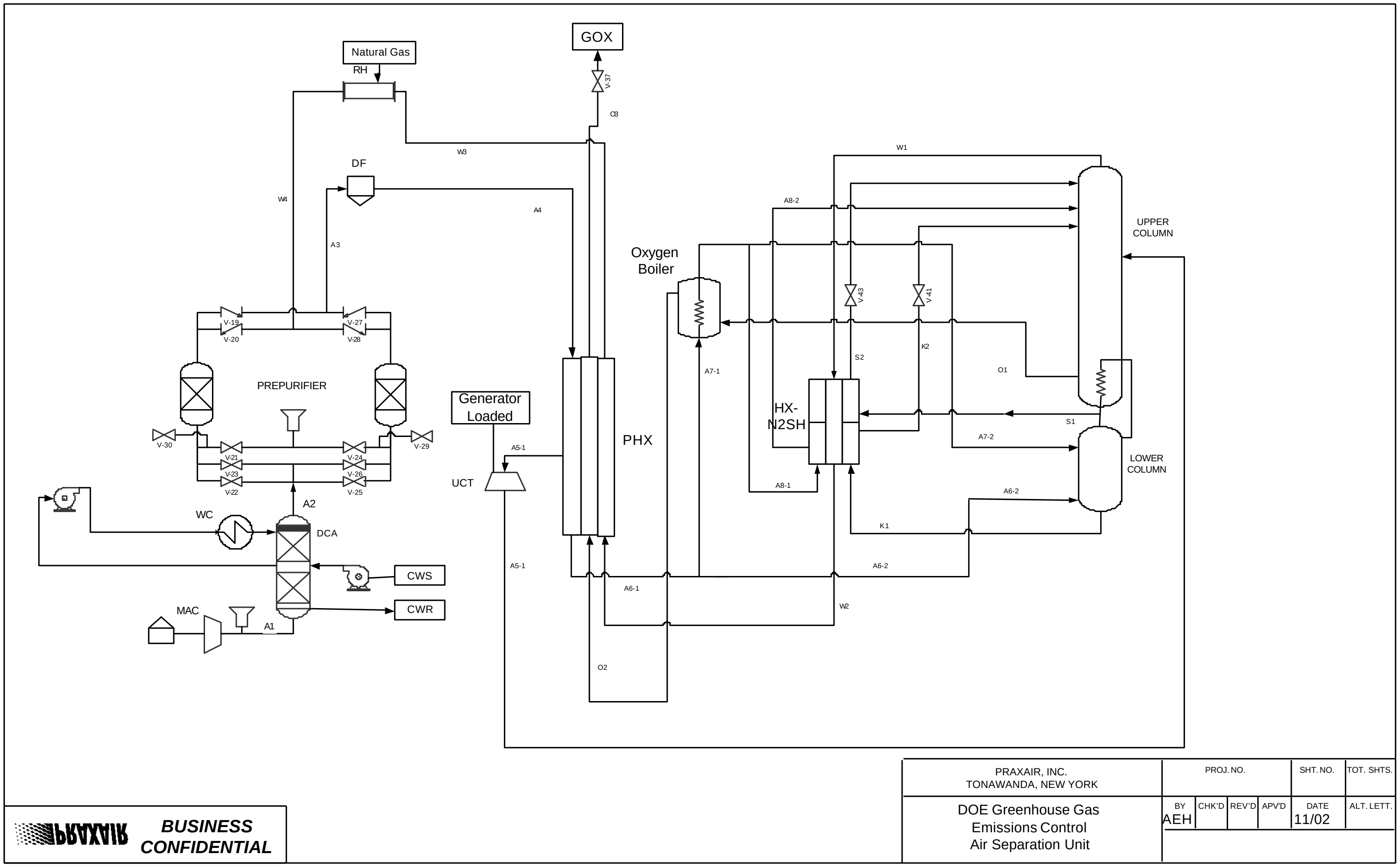


Figure 2.2. 8: Case-2 Air Separation Unit Process Flow Diagram

Air Compression:

Ambient air is drawn through the air suction filter house (ASFH) for the removal of large airborne particles prior to entering the three stage main air compressor (MAC). The filtered air is compressed in the MAC and then flows through the two-stage direct contact aftercooler (DCA). Air is cooled by exchanging heat with cooling water in the first stage and with chilled water provided by a mechanical chiller in the second stage.

Pre-purification:

The after-cooled air is then passed through the pre-purification system. The pre-purification system uses a two bed temperature-swing adsorption (TSA) process that allows continuous operation. One bed purifies the feed air while the other bed is being regenerated with first hot then cool waste nitrogen. A natural gas regeneration heater provides regeneration energy. The pre-purifier beds utilize a split adsorbent design (molecular sieve and alumina) to remove water, carbon dioxide, and most of the hydrocarbons from the air stream. After pre-purification, the air stream is passed through a dust filter to remove any solid particles.

Air Feed Streams:

The cold box requires one air feed stream. This stream is sent through the Primary Heat Exchanger (PHX) and then split into three streams. One stream is fed to the bottom of the lower column. The second air stream is fed to the oxygen boiler. The third air stream (turbine air) is cooled partially in the PHX and fed to the turbine. Adjusting the turbine airflow can modulate the total amount of refrigeration generated by the cold box.

Cold Box:

The air stream to the oxygen boiler is cooled and condensed against product oxygen and sent to both the upper and lower column.

The turbine air stream is cooled against warming nitrogen and oxygen streams. It is drawn from an intermediate location between the warm leg and the cold leg of the PHX. It is then expanded and cooled in the upper column turbine. The UCT stream enters two thirds of the way down the upper (low-pressure) distillation column.

The air entering the lower column is separated into nitrogen at the top and oxygen-enriched air (kettle liquid) at the bottom. The nitrogen at the top of the column is condensed in the main condenser against boiling oxygen from the upper column. A portion of the condensed nitrogen from the main condenser is used as reflux for the lower column. The remainder is subcooled in the cross flow passages in the nitrogen superheater section of the PHX against warming gaseous nitrogen streams from the upper column. This subcooled liquid nitrogen stream then enters the top of the upper column as reflux. The kettle liquid is subcooled in the cross flow passes of the nitrogen superheater section of the PHX and then enters the upper about 2/3 of the way down the column.

The upper column produces high purity liquid oxygen (>99.0 percent O_2) in the bottom. The upper column also produces waste nitrogen from the top. The gaseous nitrogen stream is warmed in all sections of the PHX to near-ambient temperatures. The product oxygen is boiled in the oxygen boiler against the condensing air stream and exits as product.

Products:

Gaseous oxygen is available at pressure directly from the cold box and delivered to the battery limit at 3 psig.

2.2.3.2. Material and Energy Balance

Table 2.2.9 shows the Air Separation Unit material and energy balance for Case-2. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-2 PFD for the Air Separation Unit (Figure 2.2.8). Specific energy consumption for this ASU is calculated to be about 231 kWh/ton of O_2.

Table 2.2.9: Case-2 Air Separation Unit Energy and Material Balance

Stream	**A4**	**A5-1**	**A5-2**	**A6-1**	**A6-2**
Name	Air to PHX	Air to Turbine	Turbine Air to UC	Air to O2 Boiler and LC	Air to LC
Vapour Fraction	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature (K)	300.0	185.0	128.8	104.9	104.9
Pressure (psia)	89.00	86.60	21.00	86.00	86.00
Molar Flow (kCFH-NTP)	19,050	857.0	857	18,194	12,860
Mass Flow (lb/h)	142,600	64,180	64,180	1,362,000	962,700
Liquid Volume Flow (USGPM)	3,275.0	147.4	147.4	3,128.1	2,211.0
Heat Flow (Btu/hr)	183,500,000	5,010,000	3,510,000	54,990,000	38,900,000
Comp Mole Frac (Nitrogen)	0.7811	0.7811	0.7811	0.7811	0.7811
Comp Mole Frac (Argon)	0.0093	0.0093	0.0093	0.0093	0.0093
Comp Mole Frac (Oxygen)	0.2096	0.2095	0.2096	0.2096	0.2096
Stream	**A7-1**	**A7-2**	**A8-1**	**A8-2**	**K1**
Name	Air to O2 Boiler	Boiler Air to LC	Boiler Air to N2SH	N2SH Air to UC	Ket Liq to N2SH
Vapour Fraction	1.0000	0.1279	0.0000	0.0000	0.0000
Temperature (K)	104.9	98.5	98.5	94.6	100.4
Pressure (psia)	86.00	86.20	86.20	86.20	84.50
Molar Flow (kCFH-NTP)	5,335	1,834	3,500	3,500	7,746
Mass Flow (lb/h)	399,300	137,100	262,300	262,300	595,600
Liquid Volume Flow (USGPM)	917.1	315.9	601.2	601.2	1,283.0
Heat Flow (Btu/hr)	16,090,000	-3,836,000	-9,187,000	-10,730,000	-23,270,000
Comp Mole Frac (Nitrogen)	0.7811	0.7911	0.7756	0.7756	0.5946
Comp Mole Frac (Argon)	0.0093	0.0090	0.0095	0.0095	0.0145
Comp Mole Frac (Oxygen)	0.2096	0.1999	0.2147	0.2147	0.3908
Stream	**K2**	**S1**	**S2**	**W1**	**W2**
Name	Ket Liq to UC	Shelf to N2SH	Shelf to UC	Waste N2 to N2SH	Waste N2 to PHX
Vapour Fraction	0.0000	0.0000	0.0000	13.2500	2.2780
Temperature (K)	94.6	95.6	81.3	80.3	99.9
Pressure (psia)	84.50	82.03	82.03	20.00	20.00
Molar Flow (kCFH-NTP)	7,746	6,947	6,947	15,050	15,050
Mass Flow (lb/h)	595,600	504,200	504,200	1,095,000	1,095,000
Liquid Volume Flow (USGPM)	1,283.0	1,244.0	1,244.0	2,695.0	2,695.0
Heat Flow (Btu/hr)	-26,190,000	-24,640,000	-24,640,000	36,910,000	47,140,000
Comp Mole Frac (Nitrogen)	0.5946	0.9916	0.9916	0.9886	0.9886
Comp Mole Frac (Argon)	0.0145	0.0034	0.0034	0.0094	0.0094
Comp Mole Frac (Oxygen)	0.3908	0.0050	0.0050	0.0020	0.0020
Stream	**W3**	**O1**	**O2**	**O3**	
Name	WN2 to Regen and Vent	O2 to Oxygen Boiler	O2 to PHX	Oxygen Product	
Vapour Fraction	1.0000	0.0000	1.0000	1.0000	
Temperature (K)	291.3	93.7	94.4	291.3	
Pressure (psia)	17.00	21.00	22.50	18.00	
Molar Flow (kCFH-NTP)	15,050	3,998	3,998	3,998	
Mass Flow (lb/h)	1,095,000	331,500	331,500	331,500	
Liquid Volume Flow (USGPM)	2,695.0	580.8	580.8	580.8	
Heat Flow (Btu/hr)	141,200,000	-18,080,000	11,660,000	37,530,000	
Comp Mole Frac (Nitrogen)	0.9886	0.0000	0.0000	0.0000	
Comp Mole Frac (Argon)	0.0094	0.0090	0.0090	0.0090	
Comp Mole Frac (Oxygen)	0.0020	0.9910	0.9910	0.9910	

2.2.3.3. Air Separation Unit Utility Summary

The following tables show the expected Electricity, Water and Natural Gas use for the ASU. The utilities presented here are for the quantity of oxygen specified in Table 2.2.9.

- Electric Power:

Components	kW
BLAC	36,600
Turbine	(445)
Water Chiller	1,400
DCA Pumps	170
Misc. (Incl. Lube Oil)	75
Total Average Power (+/-5%)	37,800

- Estimated Cooling Water Flow:

Estimated Cooling Water Rise: 15 F	
Total Average CW Flow, gpm	23,150

- Estimated Natural Gas Flow:

Natural Gas used for 1/3 of time	
Natural Gas Use (peak), lbm/hr	1,600

2.2.3.4. Air Separation Unit Equipment

Equipment for the Air Separation Unit is described in Appendix I, Section 9.1.2.3.

2.2.3.5. Air Separation Unit Miscellaneous Items

This section covers items such as availability, chemical requirements and operating manpower for the Air Separation Unit.

Plant Availability:

In the broad sense, “availability” means the fraction of time a unit is able to supply product at various capacity levels. Availability has two components, those related to planned outages and those related to unplanned outages.

The historic scheduled outage rate for Praxair designed large plants is less than 0.01 planned outage hours/period hours. Planned outages are normally planned to coincide with customer outages and thereby avoiding any customer impact. This activity can be planned in advance and is typically every five years.

Unplanned outages can occur because of any number of causes such as equipment failure, control system component failures, or an unnoticed degradation of a process value, which has reached a shutdown set point. In most cases, the cause of a shutdown can be rectified quickly, and the plant can be restarted in 1 to 2 hours.

With a long history of operating air separation plants, Praxair has experienced an average unplanned (forced) outage rate of less than 0.01 unplanned outage hours/period hours - availability greater than 99 percent (without backup system and excluding planned maintenance and utility outages). Molecular sieve pre-purifier plant cycles are inherently more reliable than the older reversing heat exchanger plants. The molecular sieve pre-purifier plant cycle proposed herein will demonstrate availability at the higher end of this experience.

A recent snapshot of product availability data from large plants over past 2.5 years showed all product availability of 99.16 percent. This number is not just based upon primary product like oxygen, but includes secondary products like argon and/or merchant liquid product as well. The historical distribution of unplanned outage events from the same group of plants is given below. The distribution provides an indication of how soon the product production may be restarted following an unexpected shutdown.

Duration (hours)	percent of Unplanned Outages
< 4	50 to 70
4 - 8	15 to 25
8 - 12	5 to 10
12 - 16	3 to 8
16 - 24	0 to 5
> 24	0 to 5

The high reliability of Praxair's plants are achieved through the following:

- Selection of equipment/fabrication services from qualified suppliers.
- Application of controls strategies to eliminate unnecessary outages.
- Plant design, layout, and cold box packaging to allow easy maintenance access.
- Application of best practices for plant operation and maintenance to ensure that the facility meets high expectations of safety, availability, efficiency, and regulatory requirements.

Based upon our history and practices availability of this plant is expected to be 99 percent (considering unplanned outages).

Chemical Requirements:

There are no major on-going chemical requirements, as follows:

- Cooling Water is supplied by others, thus major treatment chemicals are part of this supply.
- With a small closed loop cooling system, some minor treatment chemicals will be required.
- Minor consumable items such as analyzer zero span and fuel gas cylinders, as well as, lube oil top-off will be required.
- Pre-purifier adsorbent is included in plant pricing and is typically not replaced.
- To cover minor consumables, approximately $20,000/year is estimated.

Operating Manpower:

- Operating staff :

Supervisor	1
Plant Engineering/Assistant Manager	1
Operators	6
Maintenance (Mechanical & Instrumentation)	3

- Major maintenance would be staffed externally – either from power plant staff or contractors.

2.2.4. Case-2 Balance of Plant Equipment and Performance

The balance of plant equipment described in this section includes the steam cycle performance and equipment, the draft system equipment, the cooling system equipment, and the material handling equipment (coal, limestone, and ash). Refer to Appendix I for equipment lists and Appendix II for drawings.

2.2.4.1. Steam Cycle Performance

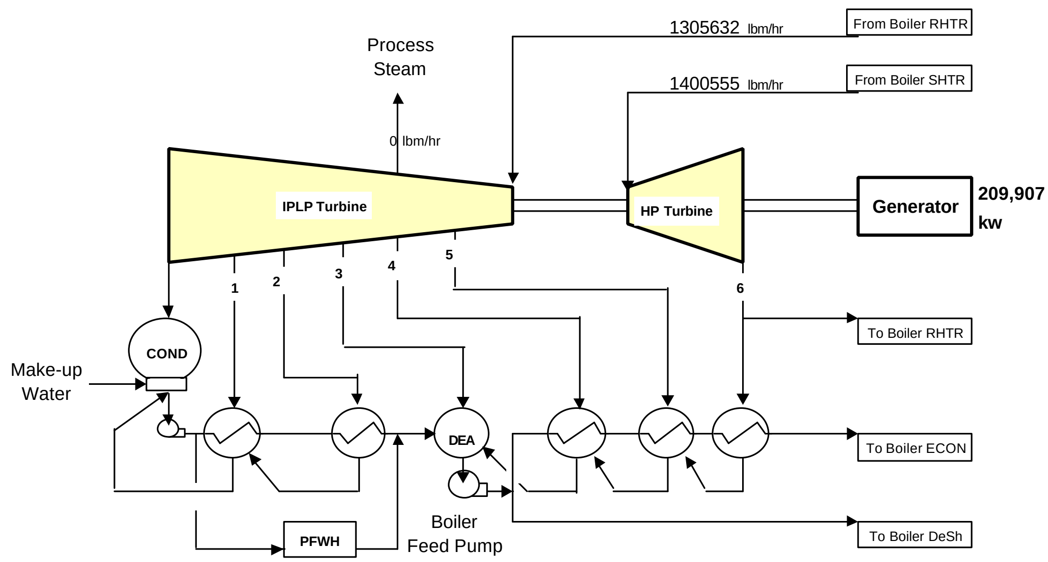
The steam cycle for the Case-2 is shown schematically in Figure 2.2.9. The Mollier diagram which illustrates the process on enthalpy - entropy coordinates is the same as for Case-1 and is not repeated here. The high-pressure turbine expands 1,400,555 lbm/hr of steam at 1,800 psia and 1,005 ^{o}F. Reheat steam (1,305,632 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,005 ^{o}F. The condenser pressure used for Case-2 and all other cases in this study was 3.0 in. Hga.

The steam cycle starts at the condenser hot well, which is a receptacle for the condensed steam from the exhaust of the steam turbine. The condensate flows to the suction of the condensate pumps, which increase the pressure of the fluid by a nominal 250 psi to transport it through the piping system and enable it to enter the open contact heater, or deaerator. The condensate passes through a gland steam condenser (SPE) first, followed in series by two low-pressure extraction feedwater heaters. The heaters successively increase the condensate temperature to a nominal 221°F by condensing and partially sub-cooling steam extracted from the LP steam turbine section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (now referred to as heater drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the condenser. The Case-2 condensate and feedwater system differs from Case 1 in that there is a parallel low-pressure Feedwater Heater (PFWH - heated by flue gas) in a parallel feedwater stream with the two low-pressure extraction feedwater heaters as shown in Figure 2.2.9

The condensate entering the deaerator is heated and stripped of noncondensable gases by contact with the steam entering the unit. The steam is condensed and, along with the heated condensate, flows by gravity to a deaerator storage tank. The boiler feedwater pumps take suction from the storage tank and increase the fluid pressure to a nominal 2200 psig. Both the condensate pump and boiler feed pump are electric motor driven. The boosted condensate flows through three more high-pressure feedwater heaters, increasing in temperature to 470°F at the entrance to the boiler economizer section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the deaerator.

Within the CFB boiler the feedwater is evaporated and finally superheated. The high-pressure superheated steam leaving the finishing superheater (1,400,555 lbm/hr of steam at 1,815 psia and 1,000 ^{o}F) is expanded through the high-pressure turbine. Reheat steam (1,305,632 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,000 ^{o}F. These conditions (temperatures, pressures) represent common steam cycle operating conditions for existing utility scale CFB power generation systems in use today. The reheated steam expands through the intermediate and low-pressure turbines before exhausting to the condenser. The condenser pressure used for Case-1 and all other cases in this study was 3.0 in. Hga.

The steam turbine performance analysis results show the generator produces 209,907 kW output and the steam turbine heat rate is about 8,256 Btu/kWh. The generator output, turbine heat rate and condenser losses are slightly higher for Case-2 than for Case-1. This is a result of the PFWH, which reduces extraction flows to the low-pressure extraction feedwater heaters and increases LP turbine power output.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Steam Turbine Power Outpu	728	Boiler Heat Input	1702
Process Steam Heat Loss	0	BFP & CP Input	12
Condenser Loss	986	**Total Energy Input**	1715
Total Energy Output	1715	**In - Out**	0

Turbine Heat Rate 8256 (Btu/kwhr)

Figure 2.2. 9: Case-2 Steam Cycle Schematic and Performance

2.2.4.2. Steam Cycle Equipment

This section provides a brief description of the major equipment (steam turbine, condensate and feedwater systems) utilized for the steam cycle of this case.

Steam Turbine:

The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the boiler passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheated steam flows through the reheat stop valves and intercept valves and enters the IP section at 465 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser.

The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for economic unit cycling.

Condensate and Feedwater Systems:

The function of the condensate system is to pump condensate from the condenser hot well to the deaerator, through the gland steam condenser and the LP feedwater heaters. The Case-2 condensate and feedwater system differs from Case 1 in that there is a parallel low-pressure Feedwater Heater (PFWH - heated by flue gas) in a parallel feedwater stream with the traditional extraction feedwater heaters. This PFWH is part of the Boiler scope of supply.

The system consists of one main condenser; two 50 percent capacity, motor-driven vertical condensate pumps; one gland steam condenser; two LP heaters, and one deaerator with a storage tank. Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump feedwater from the deaerator storage tank to the boiler economizer. Two motor-driven boiler feed pumps are provided to pump feedwater through the three stages of HP feedwater heaters. Pneumatic flow control valves control the recirculation flow. In addition, the suctions of the boiler feed pumps are equipped with startup strainers, which are utilized during initial startup and following major outages or system maintenance.

2.2.4.3. Other Balance of Plant Equipment

The systems for draft, solids handling (coal, limestone, and ash), cooling, electrical, and other BOP systems are described in this section for Case-2.

Draft System:

The flue gas is moved through the boiler, baghouse and other Boiler Island equipment with the draft system. The draft system includes the Gas Recirculation (GR) fans, the fluidizing gas blowers, the induced draft (ID) Fan, and the associated ductwork and expansion joints. This case has no traditional stack as the flue gas generated is supplied to the gas processing system where the CO_2 is purified and liquefied for sequestration or usage. The fans, and blowers are driven with electric motors and controlled to operate the unit in a balanced draft mode with the cyclone inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

Recirculated flue gas from the GR fan is mixed with oxygen from the ASU to provide a combustion oxidant stream, which is split into several flow paths.

Combustion gases exit the furnace and flow through two cyclones, which separate out ash and partially burned fuel particles. These solids are recycled back to the furnace, passing through J-valves, or seal pots, located below the cyclones. The solids leaving the seal pots are then split into two streams. The first stream is uncooled and flows directly to the combustor. The second stream flows through the moving bed heat exchanger where it is cooled before re-entering the furnace at the back wall.

The gas exiting the cyclones passes to the convection pass of the CFB, flowing through only an economizer section. The gases leaving the convection pass flow through the tubular oxygen preheater and then exit the CFB steam generator to the baghouse for particulate capture. The flue gas leaving the baghouse are further cooled in a PFWH which is a low temperature economizer section and finally in a spray water cooler to about 100 ^{o}F. The gases are drawn through the CFB, baghouse, PFWH, and spray cooler with the Induced Draft Fan and then are recirculated to the CFB or discharged to the Gas Processing System.

The following fans and blowers are provided with the scope of supply of the Oxygen-fired CFB steam generator:

- Gas Recirculation fan, which provides recirculated flue gas to be mixed with oxygen from the ASU such that the mixed oxidant stream contains about 70 percent oxygen. This fan is a centrifugal type unit, supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.2.10). The electric power required for the electric motor drive is 341 kW.

Table 2.2. 10: Gas Recirculation Fan Specification

Gas Analysis			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	180833	217000
Gas Inlet Temperature	(Deg F)	112.2	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	17.41	
Pressure Rise	(in wg)	75.0	97.5

- Induced draft fan, a centrifugal unit supplied with electric motor drive and inlet damper (see Table 2.2.11). The electric power required for the electric motor drive is 511 kW.

Table 2.2. 11: Induced Draft Fan Specification

Gas Analysis			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	637149	764579
Gas Inlet Temperature	(Deg F)	100.0	
Inlet Pressure	(psia)	13.64	
Outlet Pressure	(psia)	14.70	
Pressure Rise	(in wg)	29.5	38.4

- Fluidizing gas blowers, centrifugal units that provide recirculated flue gas for cooling and sealing the seal pots, and for assisting in the conveyance of cyclone bottoms (see Table 2.2.12). The electric power required for the electric motor drive is 209 kW.

Table 2.2. 12 Fluidizing Gas Blower Specification

Gas Analysis			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	37354	44825
Gas Inlet Temperature	(Deg F)	112.2	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	23.70	
Pressure Rise	(psia)	9.0	11.7

Ducting and Stack:

There is no stack included in Case-2. The flue gas product leaving the Boiler Island, which is rich in CO_2, is delivered to the Gas Processing System (GPS) where the CO_2 stream is further purified for sequestration or usage. The impurities removed in the GPS, primarily nitrogen and oxygen are vented to atmosphere.

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1/4" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the three silos.

Technical Requirements and Design Basis:

- Coal burn rate:
  - Maximum coal burn rate = 163,043 lbm/h = 81.6 tph plus 10 percent margin = 90 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
  - Average coal burn rate = 139,000 lbm/h = 70 tph (based on MCR rate multiplied by an 85 percent capacity factor)
  - Coal delivered to the plant by unit trains:
  - One and one-half unit trains per week at maximum burn rate
  - One unit train per week at average burn rate
  - Each unit train shall have 10,000 tons (100-ton cars) capacity
  - Unloading rate = 9 cars/hour (maximum)
  - Total unloading time per unit train = 11 hours (minimum)
  - Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
  - Reclaim rate = 300 tph
  - Storage piles with liners, run-off collection, and treatment systems:
  - Active storage = 6,600 tons (72 hours at maximum burn rate)
  - Dead storage = 50,000 tons (30 days at average burn rate)

Table 2.2. 13: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	90
Active Storage, tons	6,600
Dead Storage, tons	50,000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,000-ton silo to accommodate 3 days operation.

Bottom Ash Removal:

Bottom ash, or bed drain material, constitutes approximately two-thirds of the solid waste material discharged by the CFB steam generator. This bottom ash is discharged through a complement of two bed coolers (any one of which must be able to operate at 100 percent load on the design coal). The stripper/coolers cool the bed material to a temperature in the range of 300 ^oF (design coal) to a maximum of 500 ^oF (worst fuel) prior to discharge via rotary valves to the bed material conveying system. The steam generator scope terminates at the outlets of the rotary valves.

Fly Ash Removal:

Fly ash comprises approximately one-third of the solid waste discharged from the CFB steam generator. Approximately 8 percent of the total solids (fly ash plus bed material) is separated out in the economizer and oxygen heater hoppers; 25 percent of the total solids is carried in the gases leaving the steam generator en route to the baghouse. Fly ash is removed from the stack gas through a baghouse filter. Particulate conditions are as follows:

Design Specification for Particulate Removal System:

- Total solids to particulate removal system (stream 6, Figure 2.2.1) = 11,998 lbm/h
- Particle size distribution of particulate matter leaving cyclone (streams 5, 6, Figure 2.2.1), see Table 2.2.14.

Table 2.2. 14: Particle Size Distribution

% Wt. Less	Diameter (Micron, μ)
100	192
99	160
90	74
80	50
70	37
60	30
50	24
40	16
30	12
20	8
10	4
1	< 4

- Solids leaving particulate removal system (stream 7, Figure 2.2.1) meet applicable environmental regulations, see Table 2.2.15.

Table 2.2. 15: Fly Ash Removal Design Summary

Design Parameter	Value
Flue Gas Temperature, ^oF	313
Flue Gas Flow Rate, lbm/h	682,975
Flue Gas Flow Rate, acfm	170,561
Particulate Removal, lbm/h	11,998
Particulate Loading, grains/acf	8207

Ash Handling:

The function of the ash handling system is to provide the equipment required for conveying, preparing, storing, and disposing the bottom ash and fly ash that is produced on a daily basis by the boiler. The scope of the system is from the bag filter hoppers, oxygen heater hopper collectors, and bottom ash hoppers to the truck filling stations.

The fly ash collected in the bag filter and the oxygen heater is conveyed to the fly ash storage silo. A pneumatic transport system using low-pressure air from a blower provides the transport mechanism for the fly ash. Fly ash is discharged through a wet unloader, which conditions the fly ash and conveys it through a telescopic unloading chute into a truck for disposal.

The bottom ash from the boiler is drained from the bed, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. Ash from the fluidized bed ash coolers is drained to a complement of screw coolers, which discharge the cooled ash to a drag chain conveyor for transport to a surge bin. The latter is within the boiler scope of supply.

The cooled ash is pneumatically conveyed to the bottom ash silo from the surge bin. The silo is sized for a nominal holdup capacity of 36 hours of full-load operation (1,200 tons capacity). At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately 30 truck loads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.2.16: Ash Handling System Design Summary

Design Parameter	Value
Flyash from Baghouse, lbm/h	11,998
Ash from Boiler, lbm/h	65,225
Ash temperature, ^{o}F	520

Circulating Water System:

The function of the circulating water system is to supply cooling water to condense the main turbine exhaust steam. The system consists of two 50 percent capacity vertical circulating water pumps, a multi-cell mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Condenser Analysis:

The condenser system analysis is detailed in Table 2.2.17.

Table 2.2.17: Condenser Analysis

Item	Value	Units
Pressure	3.0	in. Hga
M stm-in	1,016,890	lbm/h
T stm-in	115.1	°F
P stm-in	1.474	psia
H stm-in	1051.7	Btu/lbm
M drain-in	81,274	lbm/h
H drain-in	89.7	Btu/lbm
H condensate	83.0	Btu/lbm
M condensate	1,098,164	lbm/h
Q condenser	986.3	10^6 Btu/h

Waste Treatment System:

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment is housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge de-watering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water are provided. A 200,000-gallon storage tank provides a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide distributed control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes

from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Plant Layout and Plot Plan:

The Case-2 plant is arranged functionally to address the flow of material and utilities through the plant site. A plan view of the boiler, power-generating components, and overall site plan for the entire plant is shown in Appendix II.

2.2.5. Case-2 Overall Plant Performance and CO_2 Emissions

Overall plant performance and emissions for Case-2 are summarized in Table 2.2.18. The Case-1 (Base Case) values are also listed along side for comparison purposes.

Boiler efficiency for Case-2 is calculated to be 94.12 percent (HHV basis) as compared to 89.46 percent for the Base Case. There is a significant improvement in boiler efficiency with oxygen firing as compared to air firing. The improvement is primarily due to the reduced dry gas loss resulting from the oxygen firing. With air firing about 3.3 lbm of nitrogen are caried into the system at ambient temperature with each 1.0 lbm of oxygen. The nitrogen and other combustion products leave the boiler at the exit flue gas temperature. The sensible energy in this exit flue gas stream represents an energy loss to the boiler. For Case-1 the sensible energy leaving the boiler with the nitrogen is nearly 4 percent of the coal heat input. With oxygen firing this loss associated with inert nitrogen is eliminated. Other attributes associated with oxygen firing also contribute to the improved boiler efficiency but the elimination of nitrogen provides the large majority of the improvement.

The steam cycle thermal efficiency including the boiler feed pump debit is about 41.3 percent as compared to 41.9 percent for Case-1. The slight reduction is due to some low level heat recovery, which is required with the oxygen-fired system. The generator output is 209,907 kW.

The net plant heat rate and thermal efficiency for Case-2 are calculated to be 13,546 Btu/kWh and 25.2 percent respectively (HHV basis).

Auxiliary power for Case-2 is 75,393 kW (about 35.9 percent of generator output). The large auxiliary power increase, as compared to the Base Case, is due primarily to the large power requirement of the cryogenic based ASU and the gas compression requirement in the Gas Processing System of Case-2.

The resulting net plant output for Case-2 is 134,514 kW or about 70 percent of the Base Case output.

Carbon dioxide emissions for Case-2 are 24,618 lbm/hr or about 0.18 lbm/kWh on a normalized basis. This represents about 9 percent of the Case-1 normalized CO_2 emissions and a CO_2 avoided value of 1.81 lbm/kWh.

Table 2.2.18: Case-2 Overall Plant Performance and Emissions

		CFB Air Fired (Case 1)	CFB Cryogenic O_2 Fired (Case 2)
Auxiliary Power Listing	**(Units)**		
Induced Draft Fan	(kW)	2285	511
Primary Air Fan	(kW)	2427	n/a
Secondary Air Fan	(kW)	1142	n/a
Fluidizing Air Blower	(kW)	920	209
Transport Air Fan	(kW)	n/a	n/a
Gas Recirculation Fan	(kW)	n/a	341
Coal Handling, Preparation, and Feed	(kW)	300	292
Limestone Handling and Feed	(kW)	200	195
Limestone Blower	(kW)	150	146
Ash Handling	(kW)	200	195
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	151
Boiler Feed Pump	(kW)	3715	3715
Condensate Pump	(kW)	79	79
Circulating Water Pump	(kW)	1400	1877
Cooling Tower Fans	(kW)	1400	1877
Steam Turbine Auxiliaries	(kW)	200	206
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719
Transformer Loss	(kW)	470	472
Subtotal	(kW)	16007	10983
	(frac. of Gen. Output)	0.077	0.052
Auxiliary Power Summary			
Traditional Power Plant Auxiliary Power	(kW)	16007	10983
Air Separation Unit or Fuel Compressor	(kW)	n/a	37505
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a
CO_2 Removal System Auxiliary Power	(kW)	n/a	26905
Total Auxiliary Power	(kW)	16007	75393
	(frac. of Gen. Output)	0.077	0.359
Output and Efficiency			
Main Steam Flow	(lbm/hr)	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8256
OTM System Expander Generator Output	(kW)	n/a	n/a
Gas Turbine Generator Output		n/a	n/a
Steam Turbine Generator Output	(kW)	209041	209907
Net Plant Output	(kW)	193034	134514
	(frac. of Case-1 Net Output)	1.00	0.70
Boiler Efficiency (HHV)[1]	(fraction)	0.8946	0.9412
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855	1806
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	16.5
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855	1822
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	13546
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.2520
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.71
CO_2 Emissions			
CO_2 Produced	(lbm/hr)	385427	376995
CO_2 Captured	(lbm/hr)	0	352377
Fraction of CO2 Captured	(fraction)	0.00	0.93
CO_2 Emitted	(lbm/hr)	385427	24618
Specific CO_2 Emissions	(lbm/kwhr)	2.00	0.18
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.09
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.81

1 Boiler Heat Output / (Qcoal-HHV + Qcredits)

2 Required for GPS Desiccant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4

2.3. Case-3: Oxygen Fired CFB with CO_2 Capture (sequestration only option)

This section describes a power plant comprised of oxygen-fired Circulating Fluidized Bed (CFB) boiler, a cryogenic type Air Separation Unit (ASU), and a subcritical steam plant with reheat (1,800 psia / 1,000 ^oF / 1,000 ^oF). The plant is designed to produce a flue gas having a high concentration of CO_2. This stream is then further processed to be acceptable for sequestration. The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

The basic CO_2 capture concept for Case-3 is similar to Case-2 in that combustion air is replaced with oxygen thereby creating a high CO_2 content flue gas stream. The oxygen is again produced from a cryogenic based ASU. The difference between Case-3 and Case-2 is in the requirements for and the design of the Gas Processing System. In Case-3, the flue gas product stream leaving the Boiler Island system is processed in such a manner (drying and compression only) as to be suitable for sequestration only whereas for Case-2 it was processed to meet a specification, which could be used for EOR. Because of this difference, the gas quality requirements for Case-3 are significantly less stringent than they were for Case-2; and therefore, the Gas Processing System design is somewhat simplified and the associated Gas Processing System equipment cost is reduced. Additionally, it should be emphasized that this plant has zero gaseous emissions from the boiler.

A brief performance summary for this plant reveals the following information. The Case-3 plant produces a net plant output of about 135 MW, about 2 percent more than Case-2. The net plant heat rate and thermal efficiency are calculated to be 13,492 Btu/kWh and 25.3 percent respectively (HHV basis) for this case, about 2 percent better than Case-2. Carbon dioxide emissions are about 0.04 lbm/kWh on a normalized basis, which is about 9 percent of the Case 2 emissions. A more detailed presentation of plant performance is shown in Section 2.3.5.

2.3.1. Case-3 Boiler Island Process Description and Equipment

The Case-3 Boiler Island process description is the same as for Case-2 and is not repeated here. Refer to section 2.2.1 for the relevant Boiler Island process description.

2.3.1.1. Process Description and Process Flow Diagrams

The Case-3 Boiler Island process flow diagram is the same as for Case-2 and is not repeated here. Refer to section 2.2.1.1 and Figure 2.2.1 for the simplified Boiler Island process flow diagram.

2.3.1.2. Material and Energy Balance

The Case-3 Boiler Island material and energy balance is the same as for Case-2 and is not repeated here. Refer to section 2.2.1.2 and Table 2.2.1 for the Boiler Island material and energy balance.

2.3.1.3. Boiler Island Equipment

The Case-3 Boiler Island equipment is the same as for Case-2 and is not repeated here. Refer to section 2.2.1.3 for the Boiler Island equipment description.

2.3.2. Case-3 Gas Processing System Process Description and Equipment

This system processes the entire flue gas stream leaving the oxygen-fired Boiler Island to provide a CO_2 product stream of suitable purity for sequestration only. This represents the primary difference between Case-3 and Case-2. In Case-2 the gas specification was for EOR whereas for Case-3 it is for sequestration only and therefore is somewhat less stringent.

Cost and performance estimates were developed for all the systems and equipment required to cool, compress, and dry the CO_2, to a product quality acceptable for sequestration only.

The CO_2 / flue gas analysis specification for sequestration, given in Table 2.3.1, was used as the basis for the CO_2 capture system design. The pressure specification for the fluid is 2,000 psig and the temperature is 82 ^oF. With the fluid at these conditions it has a density of about 40 lbm/ft^3. By comparison, pure CO_2 at these conditions would exhibit a density of about 53 lbm/ft^3 and typical power plant flue gas (air fired, assuming compressibility = 1.0) would have a density of about 10 lbm/ft^3. Therefore, gas using these specifications represents a reasonable usage (about 75 percent of maximum density as compared to pure CO_2 density) of the underground volume.

Table 2.3. 1: Gas Analysis Specification for Sequestration Only

Component	Volume Percent
CO_2	89.3
N_2	6.1
O_2	4.5
SO_2	0.1

2.3.2.1. Process Description

The following describes a CO_2 recovery system that cools, compresses, and dries a CO_2 rich flue gas stream from an oxygen-fired CFB boiler to a pressure of 2,000 psig without any purification other than removal of water.

The key process parameters (pressures, temperatures, duties etc.) are shown in the material and energy balance tables and will not be repeated here except in selected instances.

Figure 2.3.1 (Refer to Section 2.3.2.2) shows the Flue Gas Cooling process flow diagram and Figure 2.3.2 shows the Flue Gas Compression and Drying process flow diagram.

Flue Gas Cooling:

Please refer to Figure 2.3.1 (Drawing D 12173-03001-0).

The feed to the Gas Processing System is the flue gas stream that leaves the PFWH of the Boiler Island. At this point, the flue gas is near the dew point of H_2O. All of the flue gas leaving the boiler is cooled to 100 °F in Gas Cooler DA-101 which operates slightly below atmospheric pressure. A significant amount of water condenses out in this cooler. Excess condensate is blown down to the cooling water system. A single vessel has been provided for this cooler.

The Gas Cooler is configured in a packed tower arrangement where the flue gas is contacted with cold water in countercurrent fashion. Warm water from the bottom of the contactor is recycled back to the top of the contactor by Water Pump GA-101 after first cooling it in an external water cooled heat exchanger, Water Cooler EB-101 (plate and frame exchanger). The cooling water for this exchanger comes from the cooling tower.

Because the flue gas may carry a small amount of fly ash, the circulating water is filtered in Water Filter FD-101 A-E to prevent solids build-up in the circulating water. Condensate blowdown is filtered and is taken out downstream of the filter. However, the stream is not cooled and is split off before EB-101. Thus the heat load to the cooling tower is minimized.

From the Gas Cooler the gas stream is boosted in pressure by the ID fan followed by a split of the gas into two streams. This design was developed to minimize the length of ducting operating at a slight vacuum and to minimize the temperature of the gas being recycled back to the boiler. The mass flow rate of the gas recirculation stream is about 52 percent of the flow rate of the product gas stream, which proceeds to the gas compression area. The recycle stream is sized to provide an oxygen content of about 70 percent by volume in the oxidant stream supplying the boiler. The Gas Cooler minimizes the volumetric flow rate to, and the resulting power consumption of, the Flue Gas Compression equipment located downstream.

Five-Stage Gas Compression System:

Please refer to Figure 2.3.2 (Drawing D 12173-03002-0).

The initial compression section is used to compress the CO_2 rich stream to 323 psia by a three-stage centrifugal compressor (Flue Gas Compressor GB-101). After the third stage aftercoolers and knockout drum, the CO_2 rich stream is dried and followed by two more compression stages with aftercoolers. This stream is now available for sequestration or usage.

The volumetric flow to the 1st stage compressor inlet is about 71,000 ACFM and therefore, only a single frame is required. The discharge pressures of the various compressor stages have been balanced to give reasonable power distribution and discharge temperatures across the various stages. The following list shows these discharge pressures:

- 1st Stage 25 psig
- 2nd Stage 86 psig
- 3rd Stage 323 psig
- 4th Stage 821 psig
- 5th Stage 2007 psig

Power consumption for this large compressor has been estimated assuming adiabatic efficiency of 75 percent.

The hot gas leaving each compressor stage is cooled first by air coolers to 120 ^{o}F (EC-101 A-C, EC-102 A/B, EC-103 A/B, EC-105, and EC-106 A/B). The air coolers are followed by water-cooled heat exchangers (trim coolers) which further cools the gas to 95 ^{o}F (Flue Gas Compressor Trim Coolers EA-101 A/B, EA-102, EA-105, and EA-106). The one exception is flue gas compressor 3rd stage trim cooler (EA-103) which cools the gas to 90 ^{o}F to reduce the size of the dryers. Due to their large size, many of these heat exchangers consist of multiple shells. Because of highly corrosive conditions, the process side of the coolers must be stainless steel.

Because the flue gas stream leaving Flue Gas Cooler DA-101 is saturated, some water condenses out in the aftercoolers. The sour condensate is separated in knockout drums (FA-100/1/2/3/4) equipped with mist eliminator pads. Condensate from these drums is drained to the cooling tower or to waste water treatment. To prevent corrosion, these drums have a stainless steel liner.

Gas Drying:

Please refer to Figure 2.3.2 (Drawing D 12173-03002-0).

It is necessary to dry the CO_2 stream to meet the product specification. Flue gas leaving the 3rd stage compressor discharge knockout drum (FA-103) is fed to Flue Gas Drier PA-101 where additional moisture is removed. A mole sieve drier has been selected.

The performance of a fixed-bed drier improves as pressure increases. This favors locating the drier at the discharge of the compressor. However, as the operating pressure of the drier increases, so does the design pressure of the equipment. This favors low-pressure operation. But, at low pressure the diameter or number of the drier vessels grows, increasing the cost of the vessels. For this design the drier has been optimally located downstream of the 3rd stage compressor. The CO_2 Drier system consists of four vessels. One vessel is on line while the others are being regenerated. Flow direction is down during operation and up during regeneration.

Regeneration of a mole sieve bed requires relatively high temperature and, because HP steam pressure may fluctuate, a gas-fired heater has been specified for this service.

Flue Gas Filter FD-102 has been provided at the drier outlet to remove any fines that the gas stream may pick up from the desiccant bed.

2.3.2.2. Process Flow Diagrams

Two process flow diagrams are shown below for these systems:

- Figure 2.3.1 (Drawing D 12173-03001-0) Flue Gas Cooling PFD
- Figure 2.3.2 (Drawing D 12173-03002-0) CO_2 Compression and Liquefaction PFD

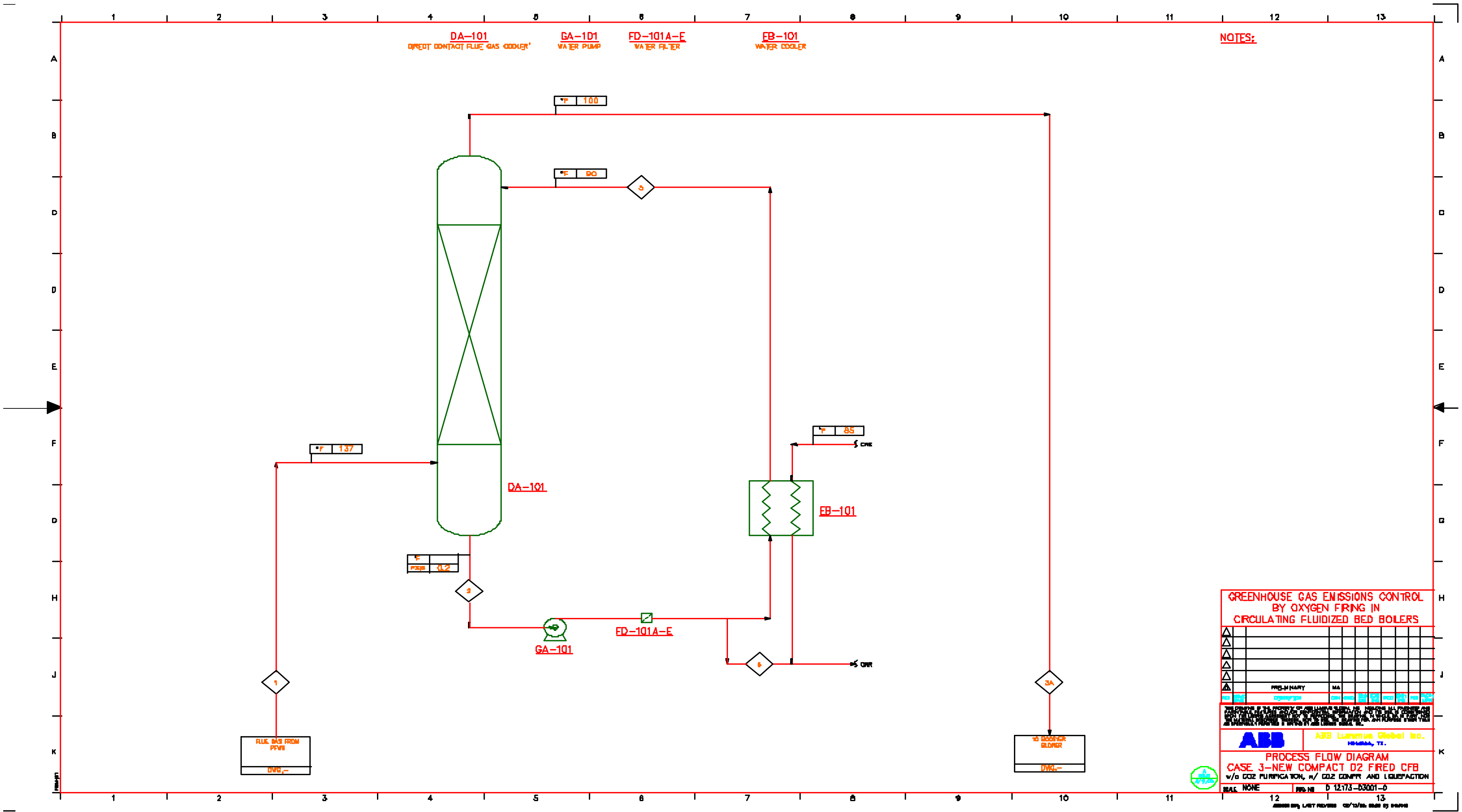


Figure 2.3. 1: Case-3 Flue Gas Cooling Process Flow Diagram

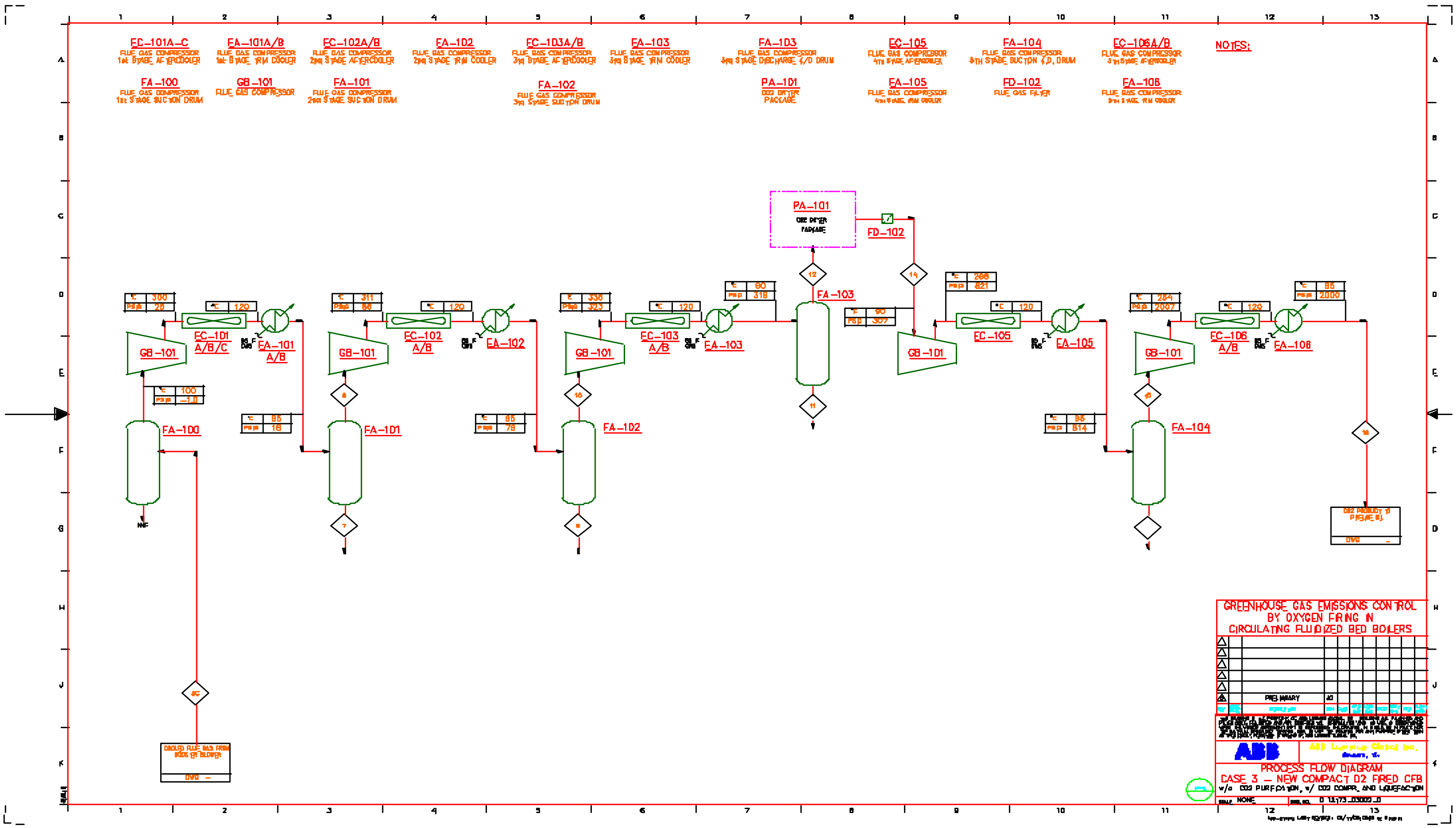


Figure 2.3. 2: Case-3 CO_2 Compression and Liquefaction Process Flow Diagram

2.3.2.3. Material and Energy Balance

Table 2.3.2 shows the material and energy balance for the Case-3 Gas Processing System.

Table 2.3. 2: Case-3 Gas Processing System Material & Energy Balance

STREAM NAME		To quench columns	From Quench columns	Excess water	From Large blowers	Quench water out	Quench water in	To liquefaction train	To boiler	To second stage
PFD STREAM NO.		**1**	**3a**	**6**	**3b**	**2**	**5**	**3c**	**3d**	**8**
VAPOR FRACTION	Molar	0.989	1.000	0.000		0.000	0.000	1.000	1.000	1.000
TEMPERATURE	°F	136.0	100	118		118	90	100	100	95
PRESSURE	PSIA	13.7	13.5	55		14	45	13.3	13.3	33
MOLAR FLOW RATE	lbmol/hr	18,158	15,628.14	2,552.00		96,582.7	94,053.0	10,240.50	5,387.64	9,798.74
MASS FLOW RATE	lb/hr	682,970	637,380	45,992		1,740,600	1,695,000	418,390	218,990	410,420
ENERGY	Btu/hr	8.14E+07	6.84E+07	-3.59E+07		-1.36E+09	-1.37E+09	4.60E+07	2.24E+07	4.22E+07
COMPOSITON	Mol %									
CO2		71.38%	82.93%	0.02%		0.02%	0.02%	83.20%	83.20%	86.95%
Nitrogen		4.90%	5.69%	0.00%		0.00%	0.00%	5.71%	5.71%	5.97%
Oxygen		3.62%	4.21%	0.00%		0.00%	0.00%	4.22%	4.22%	4.42%
H2O		20.00%	7.05%	99.98%		99.98%	99.98%	6.75%	6.75%	2.55%
Ammonia		0.00%	0.00%	0.00%		0.00%	0.00%	0.00%	0.00%	0.00%
SO2		0.10%	0.12%	0.00%		0.00%	0.00%	0.12%	0.12%	0.12%
VAPOR										
MOLAR FLOW RATE	lbmol/hr	17,949.8	15,628.1	-		-	-	10,240.5	5,387.6	9,798.7
MASS FLOW RATE	lb/hr	679,220	637,380	-		-	-	418,390	218,990	410,420
STD VOL. FLOW	MMSCFD	163.48	142.33	-		-	-	93.27	49.1	89.24
ACTUAL VOL. FLOW	ACFM	138,840	114,250	-		-	-	71,093.38	43,156.6	29,604.9
MOLECULAR WEIGHT	MW	37.84	40.78	-		-	-	40.86	40.86	41.89
DENSITY	lb/ft³	0.08	0.09	-		-	-	0.10	0.10	0.23
VISCOSITY	cP	0.0145	0.0150	-		-	-	0.0154	0.0154	0.0154
LIGHT LIQUID										
MOLAR FLOW RATE	lbmol/hr	-	-	-		-	-	-		-
MASS FLOW RATE	lb/hr	-	-	-		-	-	-		-
STD VOL. FLOW	BPD	-	-	-		-	-	-		-
ACTUAL VOL. FLOW	GPM	-	-	-		-	-	-		-
DENSITY	lb/ft³	-	-	-		-	-	-		-
MOLECULAR WEIGHT	MW	-	-	-		-	-	-		-
VISCOSITY	cP	-	-	-		-	-	-		-
SURFACE TENSION	Dyne/Cm	-	-	-		-	-	-		-
HEAVY LIQUID										
MOLAR FLOW RATE	lbmol/hr	208.09	-	2,552.00		96,582.7	94,053.0	-		-
MASS FLOW RATE	lb/hr	3,750	-	45,992		1,740,600	1,695,000	-		-
STD VOL. FLOW	BPD	257	-	3,156		119,430	116,300	-		-
ACTUAL VOL. FLOW	GPM	7.62	-	92.76		3,510.88	3,378.10	-		-
DENSITY	lb/ft³	61.32	-	61.82		61.81	62.56	-		-
VISCOSITY	cP	0.4793	-	0.5654		0.5657	0.7606	-		-
SURFACE TENSION	Dyne/Cm	66.34	-	68.11		68.12	70.83	-		-

STREAM NAME		First water KO	Second water KO	To third stage	To drier	3rd water KO	Fr om drier	To fifth stage	From 5th stage	To pipeline
PFD STREAM NO.		**7**	**9**	**10**	**12**	**11**	**14**			
VAPOR FRACTION	Molar	0.000	0.000	1.000	1.000	0.000	1.000	1.000	1.000	1.000
TEMPERATURE	°F	95	95	95	90	90	90	95	254	95
PRESSURE	PSIA	33	94	94	331	331	321	814	2,021	2,015
MOLAR FLOW RATE	lbmol/hr	441.77	160.15	9,638.58	9,575.92	62.67	9,547.61	9,547.61	9,547.61	9,547.61
MASS FLOW RATE	lb/hr	7,968	2,896	407,530	406,380	1,143	405,870	405,870	405,870	405,870
ENERGY	Btu/hr	-6.40E+06	-2.32E+06	4.08E+07	3.75E+07	-9.05E+05	3.75E+07	3.16E+07	4.22E+07	7.81E+06
COMPOSITON	Mol %									
CO2		0.07%	0.22%	98.08%	88.96%	0.78%	89.22%	89.22%	89.22%	89.22%
Nitrogen		0.00%	0.00%	0.78%	6.10%	0.00%	6.12%	6.12%	6.12%	6.12%
Oxygen		0.00%	0.00%	0.97%	4.52%	0.00%	4.53%	4.53%	4.53%	4.53%
H2O		99.92%	99.76%	0.00%	0.30%	99.16%	0.00%	0.00%	0.00%	0.00%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
SO2		0.01%	0.02%	0.17%	0.12%	0.06%	0.12%	0.12%	0.12%	0.12%
VAPOR										
MOLAR FLOW RATE	lbmol/hr	-	-	9,638.6	9,575.9	-	9,547.6	9,547.6	9,547.6	9,547.6
MASS FLOW RATE	lb/hr	-	-	407,530	406,380	-	405,870	405,870	405,870	405,870
STD VOL. FLOW	MMSCFD	-	-	87.78	87.21	-	86.96	86.96	86.96	86.96
ACTUAL VOL. FLOW	ACFM	-	-	9,879.38	2,541.77	-	2,625.22	848.01	482.16	180.65
MOLECULAR WEIGHT	MW	-	-	42.28	42.44	-	42.51	42.51	42.51	42.51
DENSITY	lb/ft³	-	-	0.69	2.66	-	2.58	7.98	14.03	37.45
VISCOSITY	cP	-	-	0.0158	0.0164	-	0.0164	0.0187	0.0272	0.0473
LIGHT LIQUID										
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-
HEAVY LIQUID										
MOLAR FLOW RATE	lbmol/hr	441.77	160.15	-	-	62.67	-	-	-	-
MASS FLOW RATE	lb/hr	7,968.30	2,895.56	-	-	1,143.23	-	-	-	-
STD VOL. FLOW	BPD	547	199	-	-	79	-	-	-	-
ACTUAL VOL. FLOW	GPM	15.91	5.78	-	-	2.27	-	-	-	-
DENSITY	lb/ft³	62.44	62.49	62.49	62.81	62.81	-	-	-	-
VISCOSITY	cP	0.7185	0.7511	0.7511	0.7772	0.7772	-	-	-	-
SURFACE TENSION	Dyne/Cm	70.31	70.20	70.20	70.27	70.27	-	-	-	-

2.3.2.4. Gas Processing System Utilities

The following tables define the cooling water, natural gas, and electrical requirements for the Case-3 Gas Processing System.

Table 2.3. 3: Case-3 Gas Processing System Cooling Water and Fuel Gas Requirements

COOLING WATER

REV	Equipment TAG NO	SERVICE	No. Installed	DUTY MMBTU/HR	INLET TEMP, F	OUTLET TEMP, F	FLOWRATE LB/HR
	EA-101	FG Comp 1 stg trim cooler	1	7.09	85	103	393,939
	EA-102	FG Comp 2 stg trim cooler	1	3.73	85	103	207,071
	EA-103	FG Comp 3 stg trim cooler	1	3.82	85	103	212,121
	EA-104	FG Comp 4 stg trim cooler	1	3.27	85	103	181,818
	EA-105	FG Comp 5 stg trim cooler	1	8.36	85	103	464,646
		TOTAL COOLING WATER		26.27			1,459,596

FUEL GAS FUEL GAS VALUE BASIS: 930 BTU/SCF (LHV)

REV	Equipment TAG NO	SERVICE	ONLINE FACTOR	COMPR HP	HEAT RATE BTU/HP-HR	DUTY MMBTU/HR	EFFICIENCY %	FLOWRATE (Peak) MMSCFD	FLOWRATE (Peak) SCFH	FLOW (Avg) MMSCFD
	PA-101	Mole sieve regeneration	73%			7.90	80%	0.255	10,618	0.186
		TOTAL FUEL GAS				7.90		0.255	10,618	0.186

Table 2.3. 4: Case-3 Gas Processing System Electrical Requirements

Number of Trains	Item Number	Service	Power (ea) including 0.95 motor eff (kW)	Total all trains (kW)
1	EC-101	Flue Gas Compressor 1st Stage Aftercooler	71	71
1	EC-102	Flue Gas Compressor 2nd Stage Aftercooler	62	62
1	EC-103	Flue Gas Compressor 3rd Stage Aftercooler	71	71
1	EC-104	Flue Gas Compressor 4th Stage Aftercooler	51	51
1	EC-105	Flue Gas Compressor 5th Stage Aftercooler	84	84
1	PA-101	Drier regen cooler	11	11
1	GB-101	1st Stage	5488	5488
1		2nd Stage	5379	5379
1		3rd Stage	7265	7265
1		4th Stage	4328	4328
1		5th Stage	3276	3276
1	GA-101	Water pump	105	105
1	PA-101	Drier regen compressor	173	173
		Total		**26364**

2.3.2.5. Gas Processing System Equipment

The equipment list for the Case-3 Gas Processing System is provided in Appendix I, Section 9.1.3.2.

2.3.3. Case-3 Air Separation Unit Process Description and Equipment

The Case-3 Air Separation Unit process description is identical to Case-2 and is not repeated here. Refer to section 2.2.3 for the relevant ASU process description.

2.3.3.1. Process Description and Process Flow Diagrams

The Case-3 ASU process flow diagram is identical to Case-2 and is not repeated here. Refer to section 2.2.3.1 and Figure 2.2.7 for the ASU process flow diagram.

2.3.3.2. Material and Energy Balance

The Case-3 ASU material and energy balance is identical to Case-2 and is not repeated here. Refer to section 2.2.3.2 and Table 2.2.10 for the ASU material and energy balance.

2.3.3.3. Air Separation Unit Utility Summary

The Case-3 ASU utilities are identical to Case-2 and are not repeated here. Refer to section 2.2.3.3

2.3.3.4. Air Separation Unit Equipment

The Case-3 ASU equipment description is identical to Case-2 and is not repeated here. Refer to section 2.2.3.4 for the relevant ASU equipment description.

2.3.4. Case-3 Balance of Plant Equipment and Performance

The Case-3 Balance of Plant equipment is nearly identical to Case-2 and the description of this equipment is not repeated here. Refer to section 2.2.4 for the relevant BOP equipment description. Refer to Appendix I for equipment lists and Appendix II for drawings.

2.3.4.1. Steam Cycle Performance

The Case-3 steam cycle performance and equipment are identical to Case-2 and the description of this equipment and performance is not repeated here. Refer to section 2.2.4.1 for the relevant steam cycle performance and equipment description.

2.3.4.2. Other Balance of Plant Equipment

The Case-3 Other Balance of Plant equipment is nearly identical to Case-2 and the description of this equipment is not repeated here. Refer to section 2.2.4.2 for the relevant BOP equipment description.

The only difference is that there is a small reduction in the heat removal by the cooling water system due to the difference in the Gas Processing System heat rejection quantity for Case-3 as compared to Case-2 as shown in Appendix I.

2.3.5. Case-3 Overall Plant Performance and CO_2 Emissions

Overall plant performance and emissions for Case-3 are summarized in Table 2.3.5. The Case-1 (Base Case) and Case-2 values are also listed along side for comparison purposes. The Base Case is shown because it is the primary comparison case for all the combustion bases CO_2 removal cases. Case-2 is shown because Case-3 and Case-2 differ only in the design and performance of the Gas Processing Systems.

Boiler efficiency for Case-3 is calculated to be 94.12 percent (HHV basis) as compared to 89.46 percent for the Base Case. The improvement is primarily due to the reduced dry gas loss resulting from the oxygen firing. Refer to Section 2.2.5 for a discussion of why the dry gas loss is reduced with oxygen firing. The boiler efficiency for Case-3 is identical to Case-2.

The steam cycle thermal efficiency including the boiler feed pump debit is about 41.3 percent. This is the same as for Case-2 and compares to 41.9 percent for Case-1. The slight reduction is due to low level heat recovery for Cases 2 and 3.

Auxiliary power for Case-3 is 74,556 kW or about 35.5 percent of generator output. The large auxiliary power increase as compared to the Base Case is due primarily to the large power requirement of the cryogenic based ASU and the gas compression requirement for the Gas Processing System of Case-3. The power requirements for the ASU and the power plant systems are identical for Case-3 and Case-2. The only difference in auxiliary power for these cases is for the Gas Processing Systems (GPS). The power requirement for the GPS of Case-3 is about 98 percent of that for Case-2. The total plant auxiliary power for Case-3 is about 99 percent of the Case-2 requirement.

The resulting net plant output for Case-3 is about 70 percent of the Base Case output and about 1 percent greater than the Case-2 output.

The net plant heat rate and thermal efficiency for Case-3 are calculated to be 13,491 Btu/kWh and 25.30 percent respectively (HHV basis). This thermal efficiency is about 67 percent of the Base Case efficiency and about 0.4 percent better than Case-2.

Carbon dioxide emissions for Case-3 are the lowest of all the plants studied since the entire flue gas stream is sequestered without any purification, which was done for the other plants. The only CO_2 emitted from the plant is from natural gas fired desiccant drier systems associated with the Air Separation Unit and the Gas Processing System. Emissions were 2,371 lbm/hr of CO_2 or about 0.02 lbm/kWh on a normalized basis. This represents about 1 percent of the Case-1 normalized CO_2 emissions and a CO_2 avoided quantity of about 1.98 lbm/kWh. Normalized CO_2 emissions (lbm/kWh) are about 10 percent of the Case-2 emissions.

Table 2.3. 5: Case-3 Overall Plant Performance and Emissions

	(Units)	CFB Air Fired (Case 1)	CFB Cryogenic O_2 Fired (Case 2)	CFB Cryogenic O_2 Fired (Case 3)
Auxiliary Power Listing				
Induced Draft Fan	(kW)	2285	511	511
Primary Air Fan	(kW)	2427	n/a	n/a
Secondary Air Fan	(kW)	1142	n/a	n/a
Fluidizing Air Blower	(kW)	920	209	209
Transport Air Fan	(kW)	n/a	n/a	n/a
Gas Recirculation Fan	(kW)	n/a	341	341
Coal Handling, Preparation, and Feed	(kW)	300	292	292
Limestone Handling and Feed	(kW)	200	195	195
Limestone Blower	(kW)	150	146	146
Ash Handling	(kW)	200	195	195
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	151	151
Boiler Feed Pump	(kW)	3715	3715	3715
Condensate Pump	(kW)	79	79	79
Circulating Water Pump	(kW)	1400	1877	1729
Cooling Tower Fans	(kW)	1400	1877	1729
Steam Turbine Auxiliaries	(kW)	200	206	206
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719	719
Transformer Loss	(kW)	470	472	472
Subtotal	(kW)	16007	10983	10687
	(frac. of Gen. Output)	0.077	0.052	0.051
Auxiliary Power Summary				
Traditional Power Plant Auxiliary Power	(kW)	16007	10983	10687
Air Separation Unit or Fuel Compressor	(kW)	n/a	37505	37505
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a	n/a
CO_2 Removal System Auxiliary Power	(kW)	n/a	26905	26364
Total Auxiliary Power	(kW)	16007	75393	74556
	(frac. of Gen. Output)	0.077	0.359	0.355
Output and Efficiency				
Main Steam Flow	(lbm/hr)	1400555	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8256	8256
OTM System Expander Generator Output	(kW)	n/a	n/a	n/a
Gas Turbine Generator Output		n/a	n/a	n/a
Steam Turbine Generator Output	(kW)	209041	209907	209907
Net Plant Output	(kW)	193034	134514	135351
	(frac. of Case-1 Net Output)	1.00	0.70	0.70
Boiler Efficiency (HHV)[1]	(fraction)	0.8946	0.9412	0.9412
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855	1806	1806
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	16.5	20.6
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855	1822	1826

[1] Boiler Heat Output / (Qcoal-HHV + Qcredits)
[2] Required for GPS Desiccant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4

	(Units)	CFB Air Fired (Case 1)	CFB Cryogenic O_2 Fired (Case 2)	CFB Cryogenic O_2 Fired (Case 3)
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	13546	13492
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.2520	0.2530
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.71	0.71
CO_2 Emissions				
CO_2 Produced	(lbm/hr)	385427	376995	377466
CO_2 Captured	(lbm/hr)	0	352377	375095
Fraction of CO2 Captured	(fraction)	0.00	0.93	0.99
CO_2 Emitted	(lbm/hr)	385427	24618	2371
Specific CO_2 Emissions	(lbm/kwhr)	2.00	0.18	0.02
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.09	0.01
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.81	1.98

2.4. Case-4: Oxygen Fired Circulating Moving Bed with CO_2 Capture

This section describes a power plant comprised of an oxygen-fired Circulating Moving Bed (CMB) boiler, a Moving Bed Heat Exchanger (MBHE), a cryogenic type Air Separation Unit (ASU), and a subcritical steam plant with reheat (1,800 psia / 1,000 ^{o}F / 1,000 ^{o}F). The plant is designed to produce a flue gas having a high concentration of CO_2. This stream is further processed in a Gas Processing System to produce a CO_2 product for usage or sequestration. The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

The basic CO_2 capture concept for Case-4 is again similar to Case-2, 3 and 6 in that combustion air is replaced with oxygen thereby creating a high CO_2 content flue gas stream. The oxygen is produced from a cryogenic based ASU. The difference for Case-4 is in the design of the boiler system. In this concept the Circulating Fluidized Bed (CFB) of Case-2 is replaced with a novel advanced boiler concept (Circulating Moving Bed - CMB) which, in air fired applications, has been shown to be less costly than the comparable CFB system (Jukkola, 2003). The CMB system for this application consists of the following major components: falling solids oxygen-fired combustor, low temperature cyclone, seal pot, moving bed heat exchanger (MBHE), oxygen heater, ash cooler, and solids return system.

The CMB system is designed to locate all the pressure parts (which are the components that contain the high-pressure steam/water working fluid of the steam cycle) in one location, the MBHE, which contains spiral-finned heat exchanger surface. The heat transfer rates are quite high in the MBHE since conduction represents a primary heat transfer mechanism. Additionally, the Cyclone is operated at a much lower temperature than Case-2 and the convection pass is eliminated. These factors and others contribute to the cost reduction for the CMB concept as compared to the traditional CFB. The complete process is explained in Section 2.4.1.

A brief performance summary for this plant reveals the following information. The Case-4 plant produces a net plant output of about 132 MW. The net plant heat rate and thermal efficiency are calculated to be 13,894 Btu/kWh and 24.6 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 0.21 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.4.5.

2.4.1. Case-4 Boiler Island Process Description and Equipment

2.4.1.1. Process Description and Process Flow Diagrams

Figure 2.4.1 shows a simplified process flow diagram for the Boiler Island of the Case-4 oxygen-fired CMB concept. Selected mass flow rates (lbm/hr) and temperatures (^{o}F) are shown on this figure. Complete data for all state points are shown in Table 2.4.1. In this concept coal or another high carbon content fuel (Stream 1) is reacted with a mixture of substantially pure oxygen and recycled flue gas (Stream 18) in the Combustor section of the Circulating Moving Bed (CMB) system. The oxygen contained in Streams 16, 17, and 18 is provided from a cryogenic Air Separation Unit (ASU).

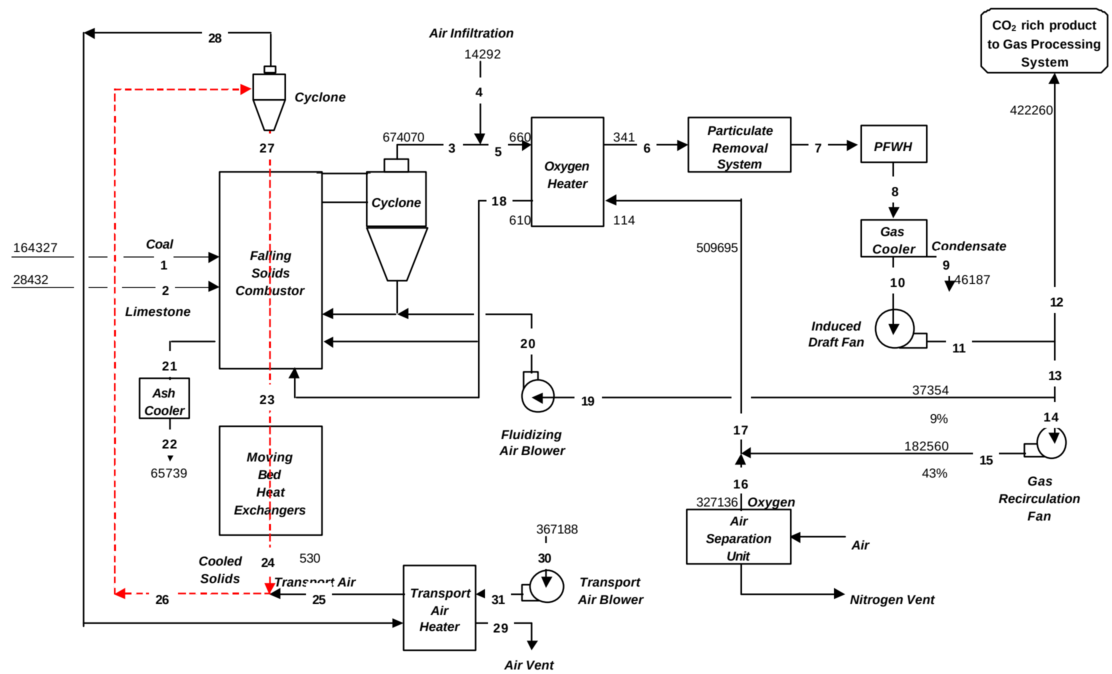


Figure 2.4. 1: Case-4 Simplified Boiler Island Gas Side Process Flow Diagram

The products of combustion leaving the combustor, flue gas comprised of primarily CO_2, H_2O vapor and un-reacted solids with smaller amounts of N_2 and O_2, flow through a cyclone, or another type of particulate removal device, where most of the solids are removed and recirculated to the Combustor.

The temperature of the flue gas stream leaving the combustor and cyclone (Stream 3) is relatively cool (about 660 ^oF). The cooling of the combustor flue gas stream is accomplished by transferring heat from the flue gases in the combustor to a relatively cool stream of bauxite solids (Stream 27). The combustion products flow vertically up the combustor exchanging heat in a counter current fashion with the stream of bauxite solids flowing vertically down the combustor.

The bauxite is referred to as a “designer solid” in that it is optimally sized to accomplish this gas to solids and then solids to working fluid (steam/water) heat transfer in a highly effective manner. The bauxite particles are relatively dense, show very little attrition and exhibit a high specific heat. Because of their size and density they are not entrained with the gas and fine bed material but continue to fall to the bottom of the falling solids combustor. The bauxite particles are heated to about 2,000 ^oF by the time they reach the combustor bottom (Stream 23).

At this point they drop by gravity through refractory lined connecting tubes into the Moving Bed Heat Exchanger located directly below the Combustor. An advantage of this system is that the heat transfer process to the steam cycle working fluid is completely separate from the combustion process. It also allows location of all the steam cycle pressure parts in the same location (near ground level) thus minimizing interconnecting piping length and cost. The bauxite stream is cooled in the counter flow Moving Bed Heat Exchanger by exchanging heat with the power cycle working fluid (superheater, reheater, evaporator, and economizer) which is contained in spiral finned tubes. The bauxite particles leaving the MBHE (Stream 24) are cooled to a temperature of about 530 ^oF.

The bauxite particles are designed to be very free flowing as they move through the compact array of spiral finned tubing comprising the MBHE. In-house test results have confirmed this gravity-induced flowability of the particles against the spiral finned tube surfaces. The solids velocity in the MBHE is slow (typically, 60-150 ft/hr) such that erosion and attrition is minimal.

The bauxite particles leaving the MBHE are then pneumatically transported in several parallel vertical pipes using hot air (Stream 26) as the transport medium back to the top of the combustor. At this location the hot air is separated from the bauxite in an array of small cyclones. The low temperature bauxite (Stream 27) then starts another cycle through the system.

The hot air leaving the small cyclones (Stream 28) is ducted to a tubular transport air heater provided to exchange heat with the incoming cool supply air (Stream 31) thus recovering most of this energy. The cool air stream leaving the transport air heater (Stream 29) is then vented to atmosphere. The transport supply air (Stream 30) is boosted to the required pressure with the transport air blower. The pressure required for Stream 31 includes all system pressure drops including the static pressure required to lift the bauxite stream to the top of the combustor.

Draining hot bed solids through water-cooled fluidized bed ash coolers (Stream 21) controls solids inventory in the system while recovering heat from the hot ash in an efficient manner. The cooling water used for the ash coolers is feedwater provided from the final extraction feedwater heater of the steam cycle. The bauxite particles are kept out of this stream with a classifier system (not shown on this diagram).

The flue gas cooling system of Case-4, which is described below, is very similar to that of Case-2 except a traditional convection pass following the cyclone is not necessary due to the low gas temperature leaving the cyclone in this case. The flue gas stream leaving the cyclone (Stream 3) is first cooled in an Oxygen Heater. The oxygen stream leaving the Air Separation Unit (Stream 16) is mixed with a small stream of recirculated flue gas (Stream 15) and the mixture is preheated in the Oxygen Heater. The quantity of recirculated flue gas is only the amount necessary to provide oxygen content of about 70 percent by volume in Streams 17 and 18.

The flue gas leaving the Oxygen Heater (Stream 6) is cleaned of fine particulate matter in the baghouse and further cooled in a Parallel Feedwater Heater (PFWH) by transferring heat to a feedwater stream in parallel with extraction feedwater heaters.

Finally, a direct contact water spray Gas Cooler is used to cool the gas before the flue gas enters the Induced Draft (ID) Fan (Stream 10). The gas cooler is used to cool the flue gas to the lowest temperature possible before recycling to minimize the power requirements for the draft system (induced draft fan, fluidizing air blower, and gas recirculation fan) and the product gas compression system. Some H_2O vapor is condensed in the Gas Cooler. This Gas Cooler system is described in detail in Section 2.4.2 as it is considered a part of the Gas Processing System.

The flue gas leaving the ID Fan (Stream 11), comprised of mostly CO_2 and H_2O vapor with smaller amounts of O_2 and N_2, is split with about 48 percent of the flue gas going to the product stream (Stream 12) for further processing and the remainder recirculated to the CMB system. The quantity of recirculated gas (Stream 13) is about 52 percent of the product gas stream (Stream 12). This quantity provides an oxygen content in Streams 17 and 18 of about 70 percent by volume.

By using oxygen instead of air for combustion, and by minimizing the amount of recirculated flue gas, the size and cost of many components (Combustor, Cyclone, Oxygen Heater ductwork, fans and other equipment) can be reduced as compared to many other concepts for CO_2 capture with CFB systems as was shown previously in Case-2.

2.4.1.2. Material and Energy Balance

Table 2.4.1 shows the Boiler Island material and energy balance for Case-4. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-4 simplified Boiler Island Process Flow Diagram (Figure 2.4.1). This performance was calculated at MCR conditions for this unit.

The MCR condition is defined as high-pressure turbine inlet conditions of 1,400,555 lbm/hr, 1,815 psia, and 1,000 ^{o}F and intermediate-pressure turbine inlet conditions of 1,305,632 lbm/hr 469 psia 1,000 ^{o}F. These conditions were used for the Base Case and all other combustion cases in this study although reheat flow was slightly higher in some cases due to differences in low-level heat recovery arrangements. The boiler was fired with enough oxygen to leave about 3 percent by volume in Stream 3, the same as for the Base Case and Cases 2 and 3. This oxygen requirement results in a stoichiometry of about 1.05 for Case-4. The resulting boiler efficiency calculated for Case-4 was 94.00 percent (HHV basis) with an oxygen heater exit gas temperature of 341 ^{o}F and the PFWH exit gas temperature of 135 ^{o}F. This boiler efficiency value takes credit for the PFWH heat recovery.

Table 2.4. 1: Case-4 Boiler Island Gas Side Material and Energy Balance

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11	12
O_2	(Lbm/hr)	5193		17956	3271	21228	21228	21228	21228		21228	21228	13958
N_2	"	2399		14268	10837	25105	25105	25105	25105		25105	25105	16508
H_2O	"	6557		65741	183	65924	65924	65924	65924	46187	19737	19737	12978
CO_2	"			574936		574936	574936	574936	574936		574936	574936	378048
SO_2	"			1168		1168	1168	1168	1168		1168	1168	768
H_2	"	5866											
Carbon	"	101965											
Sulfur	"	3845											
CaO	"												
$CaSO_4$	"												
$CaCO_3$	"		27011										
Ash	"	38502	1422										
		Coal	Limestone	Flue Gas	Infiltration Air	Flue Gas	Flue Gas	Flue Gas	Flue Gas	Condensate	Flue Gas	Flue Gas	Flue Gas
Total Gas	(Lbm/hr)			674070	14292	688361	688361	688361	688361		642174	642174	422260
Total Solids	"	164327	28432										
Total Flow	"	164327	28432	674070	14292	688361	688361	688361	688361	46187	642174	642174	422260
												0.355489	
Temperature	(Deg F)	**80**	80	670	80	660	341	341	135	100	100	112	112
Pressure	(Psia)	**14.7**	14.7	14.7	14.7	14.4	14.2	13.8	13.7	14.7	13.6	14.7	14.7
$h_{sensible}$	(Btu/lbm)			152.426	0.000	149.261	63.098	63.098	12.649		4.228	6.826	6.826
										19.960			
Chemical	(10^6 Btu/hr)	1819.753											
Sensible	(10^6 Btu/hr)	**0.0**	0.000	102.746	0.000	102.746	43.434	43.434	8.707	0.922	2.715	4.383	2.882
Latent	(10^6 Btu/hr)	**0.0**	0.000	69.028	0.192	69.220	69.220	69.220	69.220	0.000	20.724	20.724	13.627
Total Energy[(1)]	(10^6 Btu/hr)	1819.753	0.000	171.774	0.192	171.966	112.655	112.655	77.927	0.922	23.439	25.107	16.509

Constituent	(Units)	13	14	15	16	17	18	19	20	21	22
O_2	(Lbm/hr)	7269	6035	6035	323864	329899	329899	1235	1235		
N_2	"	8597	7137	7137	3271	10408	10408	1460	1460		
H_2O	"	6759	5611	5611		5611	5611	1148	1148		
CO_2	"	196888	163445	163445		163445	163445	33443	33443		
SO_2	"	400	332	332		332	332	68	68		
H_2	"										
Carbon	"									2039	2039
Sulfur	"										0
CaO	"									9080	9080
$CaSO_4$	"									14696	14696
$CaCO_3$	"										0
Ash	"									39923	39923
		Recirc Gas	Recirc Gas	Recirc Gas	Oxygen	Oxy + Recirc	Oxy + Recirc	Grease Gas	Grease Gas	Hot Ash Drain	Cool Ash Drain
Total Gas	(Lbm/hr)	219914	182560	182560	327136	509695	509695	37354	37354		
Total Solids	"						0			65739	65739
Total Flow	"	219914	182560	182560	327136	509695	509695	37354	37354	65739	65739
				0.323422					0.30		
Temperature	(Deg F)	112	112	140	**100**	114	610	112	195	2000	520
Pressure	(Psia)	14.7	14.7	17.4	17.4	17.4	17.19	14.7	23.7	14.7	14.7
$h_{sensible}$	(Btu/lbm)	6.826	6.826	12.934	4.407	7.462	123.827	6.826	25.005	545.335	95.391
Chemical	(10^6 Btu/hr)									28.740	28.740
Sensible	(10^6 Btu/hr)	1.501	1.246	2.361	1.442	3.803	63.114	0.255	0.934	35.850	6.271
Latent	(10^6 Btu/hr)	7.097	5.891	5.891	0.000	5.891	5.891	1.205	1.205	0.000	0.000
Total Energy[(1)]	(10^6 Btu/hr)	8.598	7.138	8.253	1.442	9.695	69.005	1.460	2.140	64.589	35.011

Constituent	(Units)	23	24	25	26	27	28	29	30	31
Air	(Lbm/hr)	0	0	367188	367188	0	367188	367188	367188	367188
Bauxite	"	3671881	3671881	0	3671881	3671881	0	0	0	0
Total Gas	(Lbm/hr)	0	0	367188	367188	0	367188	367188	367188	367188
Total Solids	"	3671881	3671881	0	3671881	3671881	0	0	0	0
Total Flow	"	3671881	3671881	367188	4039069	3671881	367188	367188	367188	367188
Temperature	(Deg F)	**2000**	530	430	**521**	521	521	**240**	**80**	148
Pressure	(Psia)			19.7	19.7	14.7	15.02	14.7	**14.70**	20.53
$h_{sensible\ solids}$	(Btu/lbm)	545.3	97.8		95.6	95.6				
$h_{sensible\ gas}$				85.997	108.647		108.647	39.114	0.000	16.464
Chemical	(10^6 Btu/hr)	**0**	0	0	0	0	0	0	0	0
Sensible	(10^6 Btu/hr)	2002.404	359.250	31.577	390.827	350.933	39.894	14.362	0.000	6.045
Latent	(10^6 Btu/hr)	**0**	0	0	0	0	0	0	0	0
Total Energy[(1)]	(10^6 Btu/hr)	2002.404	359.250	31.577	390.827	350.933	39.894	14.362	0.000	6.045

Notes:
(1) Energy Basis; Chemical based on Higher Heating Value (HHV); Sensible energy above 80F; Latent based on 1050 Btu/Lbm of water vapor

2.4.1.3. Boiler Island Equipment

This section describes major equipment included in the Boiler Island for Case-4. Figures 2.4.2 and 2.4.3 show general arrangement drawings of the Case-4 CMB boiler. The complete Equipment List for Case-4 is shown in Appendix I. Appendix II shows several drawings of the Boiler (key plan, boiler plan view, side elevation, and various section views). The major components include the falling solids combustor, ash coolers, fuel feed system, sorbent feed system, bauxite recycle system, cyclone, seal pots, external moving bed heat exchanger (MBHE), superheater, reheater, economizer, oxygen heater, baghouse, parallel feedwater heater (PFWH), gas cooler, and draft system.

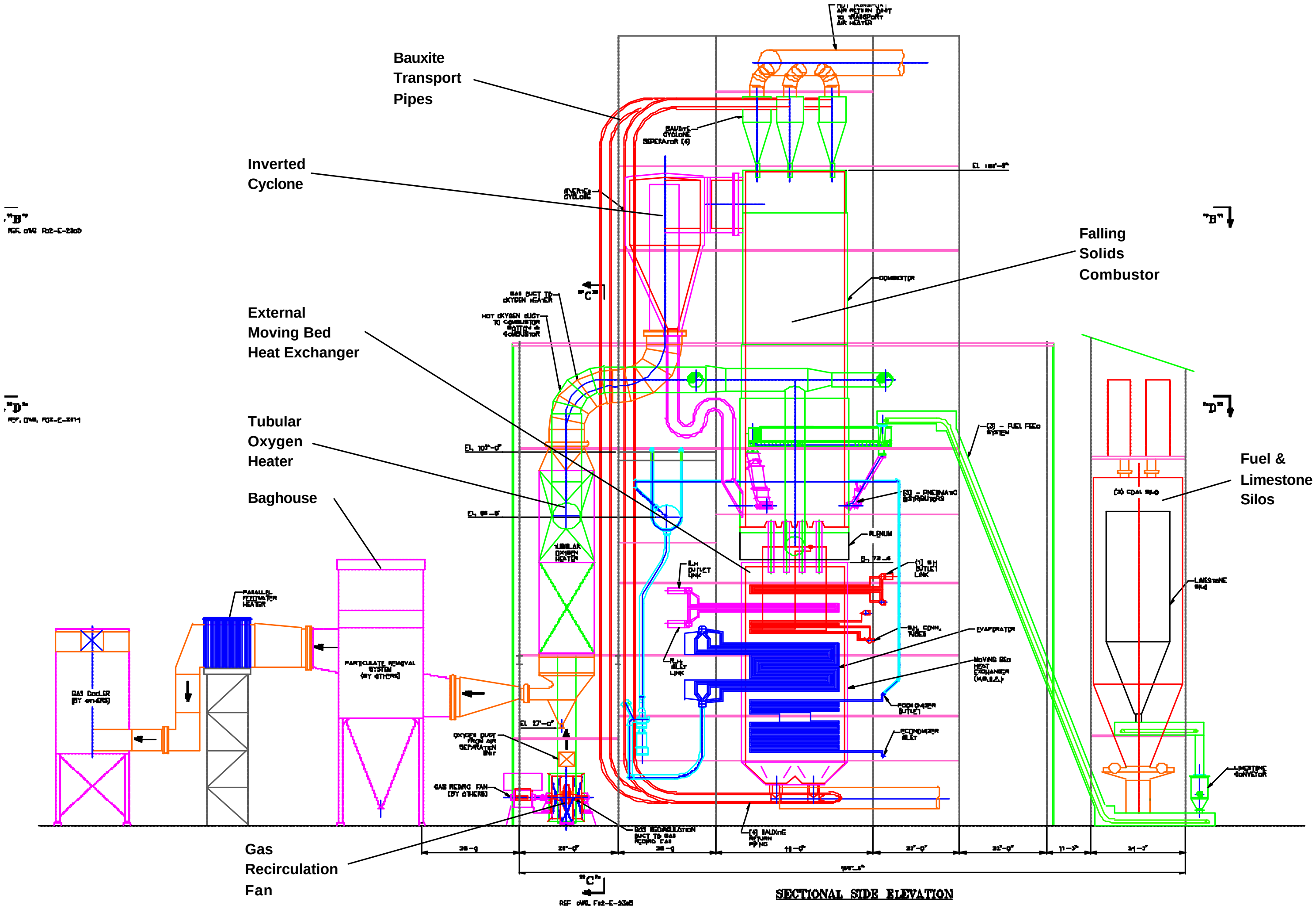


Figure 2.4. 2: Case-4 Boiler Island General Arrangement Drawing – Side Elevation

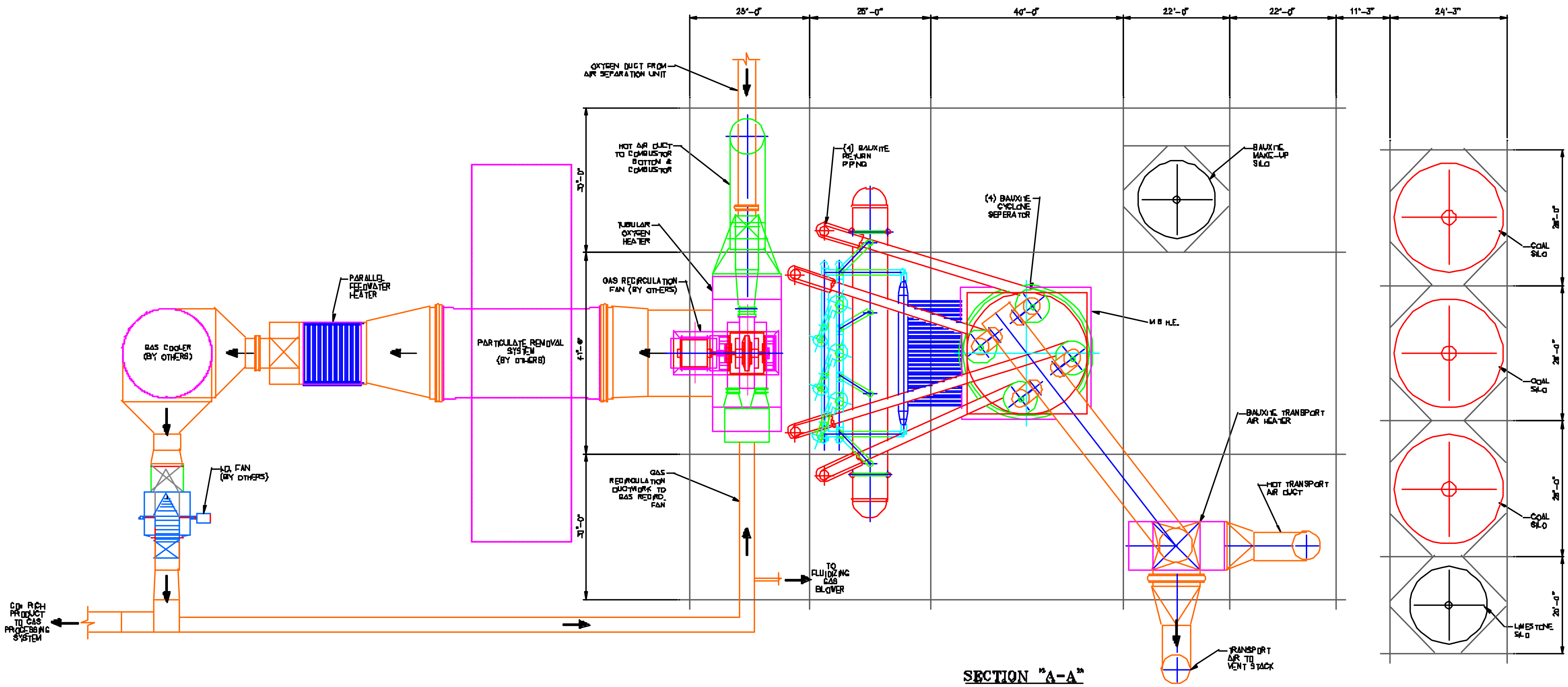


Figure 2.4. 3: Case-4 Boiler Island General Arrangement Drawing – Plan View

Falling Solids Combustor:

The combustor size is reduced significantly for Case-4 as compared to Case-1. The cylindrical combustor for Case-4 is about 25 ft in diameter and 100 ft high. Thus, the plan area for Case-4 is about 35 percent of the Case-1 plan area for nearly equivalent fuel heat input quantities. Crushed fuel, sorbent, and recycle solids are fed to the lower portion of the combustor. Primary "air" (actually a mixture of oxygen and recycled flue gas) is fed to the combustor bottom through a grid plate with secondary "air" supplied higher up in the lower combustor region. The air stream is split to provide combustion staging for NO_x reduction. Cooled bauxite (530 ^oF) leaving the moving bed heat exchanger is transported to the top of the combustor where it is fed and distributed. The bauxite provides an intermediate heat transfer material. As the cool bauxite particles fall downward in the combustor and the hot flue gas and entrained bed material moves upward, in counter-current fashion, the flue gas and bed material transfer some of their heat to the bauxite. The flue gas is cooled to about 670 ^oF at the combustor outlet (Cyclone inlet). Similarly, the bauxite particles, which exit the combustor at the bottom, are heated to about 2,000 ^oF. The bauxite particles are sized large enough to avoid entrainment by the flue gas but small enough to provide the proper gas to solids heat transfer. The hot bauxite particles leaving the lower combustor region then enter the moving bed heat exchanger, described below, and transfer heat to the steam cycle.

The combustor is constructed significantly differently than the Case-1, Case-2, and Case-3 combustors. It can be described as a cylindrical refractory lined vessel with vertical walls. The lower and upper combustor regions are formed with a multilayer refractory liner without any waterwall panels. The lower combustor has penetrations for the admission of fuel, sorbent, recycle bed material, and oxidant. These penetrations are similar to those used for Case-1, 2, and 3. Additionally, the hot bauxite must be removed from the lower combustor. This is done with a series of bauxite drain tubes connecting the lower combustor to the MBHE. In order to avoid bed material draining into the MBHE along with the bauxite particles, a portion of the oxidant is introduced in the upper MBHE. The oxidant flows up through the bauxite drain tubes at a velocity high enough to entrain bed material but not high enough to entrain the bauxite particles. Combustion occurs throughout the lower combustor, which is filled with bauxite particles and normal bed material. The upper combustor section is a cylindrical straight walled section formed with a multilayer refractory lining. The combustor flue gas is cooled exclusively by transferring heat to the bauxite particles, as mentioned above. Modulating the flow of cooled bauxite into the upper combustor controls combustor bed temperature. The bauxite stream entering the upper combustor is distributed evenly across the plan area to ensure proper heat transfer.

Fuel Feed System:

The fuel feed system for Case-4 is very similar to that for Case-1. It is designed to transport prepared coal from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, fuel conveyors, fuel feeders, feeder isolation valves, and fuel piping to the furnace.

Sorbent Feed System:

The limestone feed system for Case-4 is the same as for Case-1. The limestone feed system pneumatically transports prepared limestone from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, rotary feeders, blower, and piping from the blower to the furnace injection ports.

Bauxite Recycle System:

The bauxite recycle system is designed to transport the cooled bauxite particles leaving the moving bed heat exchanger to the top of the falling solids combustor in an energy efficient manner. The particles are then fed into the combustor and provide an indirect heat transfer medium. The dense phase pneumatic system uses ambient air as a transport medium. The air is pressurized as required in the transport air blower and then preheated in the Transport Air Heater. The preheated air then transports the cooled bauxite particles leaving the MBHE to the top of the falling solids combustor. The bauxite particles are separated from the air in a simple cyclone and then fed to the combustor. The air stream leaving the cyclone is cooled by exchanging heat with the cool incoming air stream, thus recovering its energy, in the Transport Air Heater and is then exhausted to atmosphere.

Ash Cooler:

The ash cooler design for Case-4 is the same as for Case-1. Draining hot solids through a water-cooled ash cooler controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is feedwater from the final extraction feedwater heater of the steam cycle.

Cyclone:

Flue gas and entrained solids exit the upper combustor at about 670 ^{o}F and enter the cyclone. Only one cyclone (inverted design) is required for Case-4 because of the reduced gas flow resulting from the oxygen firing. The gas temperature is also significantly reduced. The cyclone is shaped like a cylindrical cone constructed from steel plate. Solids are separated from the flue gas in the cyclone and fall into a seal pot. Well over 99 percent of the entrained solids are captured in the cyclone. Flue gas leaving the cyclone is then ducted directly to the Oxygen Heater, as there is no convection pass required for Case-4.

Seal Pot:

The seal pot for Case-4 is of the same design as in Case-1, although smaller, since less solids are recirculated due to the reduced gas flow in Case-4. The seal pot is a device that provides a pressure seal between the combustor, which is at a relativly high pressure (~ 40 inwg at the bottom), and the cyclone that is near atmospheric pressure. It is designed to move solids collected in the cyclone back to the combustor. The seal pot for Case-4 is constructed of steel plate with fluidizing nozzles located along the bottom. All of the solids in this case flow directly from the seal pot back to the combustor.

Moving Bed Heat Exchanger:

The external heat exchanger for Case-4 is a moving bed design as was used in Case-2, rather than a fluidized bed as was used in Case-1. The moving bed heat exchanger is not fluidized and contains several immersed tube bundles, which cool the hot bauxite particles leaving the lower combustor. The tube bundles in the MBHE are spiral finned and include superheater, reheater, evaporator, and economizer sections. Very high heat transfer rates are obtained in the MBHE due to the conduction heat transfer between the solids and tube. The solids moving through the heat exchanger in this case are bauxite particles as opposed to the typical bed material used in Case-2.

The MBHE is constructed of steel plate with stiffeners and refractory lined enclosure walls. It is rectangular in cross section with a hopper shaped bottom. The solids move through the bed by gravity at about 150 ft/hr.

Convection Pass:

Because the temperature of the flue gas leaving the cyclone is so low in this case (670 ^oF), there is no convection pass and the flue gas leaving the cyclone is simply ducted directly to the Oxygen Heater.

Superheater:

The superheater is divided into two major sections. Saturated steam leaving the steam drum supplies the horizontal low temperature section and finishing superheater sections located in the external moving bed heat exchanger. These sections are comprised of spiral finned tubes. There are no superheater banks located in the convective pass for Case-4. The steam leaving the finishing superheater is piped to the high-pressure turbine, where it is expanded to reheat pressure and then returned to the low temperature reheat section of the MBHE.

Reheater:

The reheater is designed as a single section located between the finishing and low temperature superheater sections. The steam is supplied to the reheater from the de-superheating spray station, which follows the high-pressure turbine exhaust. The reheater is designed as a horizontal spiral finned section located in the upper area of the MBHE. There are no reheater banks located in the convective pass for Case-4. The steam leaving the de-superheating spray station supplies the reheat section. Steam leaving the reheater is returned to the intermediate pressure turbine where it continues its expansion through the intermediate and low-pressure turbines before being exhausted to the condenser.

Evaporator:

The evaporator section for Case-4 is located in the middle portion of the MBHE. The evaporator is comprised of three banks of horizontal tubes, which evaporate high-pressure boiler feedwater. The water/steam mixture exiting the evaporator tube banks is supplied to the steam drum where the steam and water phases are separated. The separated steam supplies the low temperature superheater section. The feedwater supplying the evaporator is piped from the steam drum through circulating water pumps and is comprised of a combination of separated saturated water and subcooled water from the economizer.

Economizer:

The economizer section for Case-4 is located in the lower MBHE. The economizer is comprised of two banks of horizontal spiral finned tubes, which heat high-pressure boiler feedwater. The water exiting the economizer tube banks is supplied to the steam drum. The feedwater supplying the economizer is piped from the final extraction feedwater heater and the ash cooler.

Oxygen Heater:

A tubular regenerative oxygen heater is used to cool the flue gas leaving the cyclone by heating both the primary and secondary "air" streams prior to combustion in the furnace. The primary and secondary "air" is actually a mixture of about 65 percent by weight oxygen with recycled flue gas.

Baghouse:

The fine particulate matter for Case-4 is removed from the cooled flue gas stream leaving the oxygen heater in a baghouse. The baghouse for Case-4 is much smaller than for Case-1 due to the reduced gas flow (about 30 percent of the Case-1 flow). The ash collected in the baghouse is supplied to the ash handling system. This system is described further in Section 2.4.4.2 under Balance of Plant Equipment.

Parallel Feedwater Heater:

The Parallel Feedwater Heater (PFWH) of Case-4 is used to recover additional heat in the steam cycle for this case as shown in Figure 2.4.4. The feedwater flow is in parallel with the bottom two extraction type feedwater heaters included in the steam cycle.

The PFWH is used because in Case-4 the gas temperature leaving the Oxygen Heater is significantly higher than the gas temperature leaving the Air Heaters of Case-1 (341 ^{o}F vs. 275 ^{o}F). This occurs because the ratio of air to gas is higher in Case-1 than in Case-4 making the Air heater of Case-1 more effective than the Oxygen Heater of Case-4. Additionally, the gas must be cooled before compression to minimize the power required.

The PFWH heat exchanger is constructed similarly to the economizer heat exchanger banks used in Heat Recovery Steam Generator units. The tubes used are heavily finned, since the gas is clean and the enclosure walls are insulated steel liners. The PFWH cools the flue gas to about 135 ^{o}F. About 30 percent of the feedwater is bypassed around the bottom two extraction heaters to provide this cooling duty.

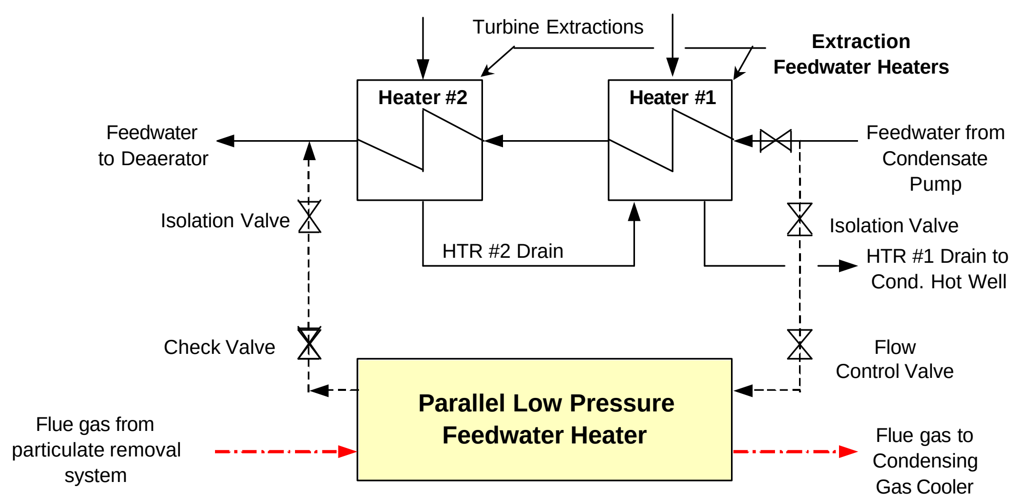


Figure 2.4. 4: Case-4 Parallel Feedwater Heater Arrangement

Gas Cooler:

The gas cooler of Case-4 is used to cool the flue gas leaving the PFWH to as low a temperature as possible in order to minimize the power requirements of the Draft System and the Gas Processing System. The Gas Cooler is a direct contact, water spray type of system. Some of the water vapor contained within the flue gas also condenses out in this cooler. This cooler is designed to cool the flue gas to 100 ^{o}F. This equipment is described further in Section 2.4.2 as it is considered a part of the Gas Processing System.

Draft System:

The flue gas is moved through the Boiler Island equipment with the draft system. The draft system includes the gas recirculation fan, the fluidizing gas blower, the induced draft (ID) Fan, the associated ductwork, and expansion joints. The induced draft fan, gas recirculation fan, and fluidizing gas blower are driven with electric motors and controlled to operate the unit in a balanced draft mode with the cyclone inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

2.4.2. Case-4 Gas Processing System Process Description and Equipment

The purpose of this system is to processes the flue gas stream leaving the oxygen-fired Boiler Island to provide a liquid CO_2 product stream of suitable purity for an EOR application.

Although the Case-4 CO_2 capture system is essentially identical to that for Case-2 except for slight gas flow changes, it is described here for completeness. This system is designed for about 94 percent CO_2 capture. Cost and performance estimates were developed for all the systems and equipment required to cool, purify, clean, compress and liquefy the CO_2, to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO_2 specification for EOR, given in Table 2.0.1, was used as the basis for the CO_2 capture system design.

A very low concentration of oxygen, in particular, is specified for meeting current pipeline operating practices, due to the corrosive nature of the oxygen. Hence, for Cases-2 and 4, whereby the final CO_2 liquid product was found to contain about 11,400 ppmv of O_2, the design of the transport pipe to an EOR site for example would have to take this characteristic under consideration.

The nitrogen concentration specified in Table 2.0.1 is < 300 ppmv. It should be noted that according to Charles Fox of Kinder Morgan (Fox, 2002), this specification is very conservative as his company specifies a maximum nitrogen concentration of 4 percent (by volume) to control the minimum miscibility pressure. In Case 4 the nitrogen concentration in the liquid product was 11,800 ppmv. The exact reasoning behind the very low nitrogen specification listed in Table 2.0.1 is not clear.

2.4.2.1. Process Description

The Case-4 Gas Processing System process description is nearly identical to Case-2 and is repeated here for completeness.

The following describes a CO_2 recovery system that cools and then compresses a CO_2 rich flue gas stream from an oxygen-fired CFB boiler to a pressure high enough so CO_2 can be liquefied. The resulting liquid CO_2 is passed through a CO_2 Stripper to reduce the N_2 and O_2 content to a level, which is optimized from an energy consumption standpoint. Then the liquid CO_2 is pumped to a high pressure so it can be economically transported for sequestration or usage. Pressure in the transport pipeline will be maintained above the critical pressure of CO_2 to avoid 2-phase flow. The overhead gas from the CO_2 Stripper is vented to atmosphere.

The key process parameters (pressures, temperatures, duties etc.) are shown in the material and energy balance tables shown in Section 2.4.2.3 and will not be repeated here except in selected instances.

Figure 2.4.5 (Refer to Section 2.4.2.2) shows the Flue Gas Cooling process flow diagram and Figure 2.4.6 shows the Flue Gas Compression and Liquefaction process flow diagram.

Flue Gas Cooling:

Please refer to Figure 2.4.5 (Drawing D 12173-04001-0).

The feed to the Gas Processing System is the flue gas stream that leaves the PFWH of the Boiler Island. At this point, the flue gas is near the dew point of H_2O. All of the flue gas leaving the boiler is cooled to 100 °F in Gas Cooler DA-101 which operates slightly below atmospheric pressure. A significant amount of water condenses out in this cooler. Excess condensate is blown down to the cooling water system. A single vessel has been provided for this cooler.

The Gas Cooler is configured in a packed tower arrangement where the flue gas is contacted with cold water in countercurrent fashion. Warm water from the bottom of the contactor is recycled back to the top of the contactor by Water Pump GA-101 after first cooling it in an external water cooled heat exchanger, Water Cooler EB-101 (plate and frame exchanger). The cooling water for this exchanger comes from the new cooling tower.

Because the flue gas may carry a small amount of fly ash, the circulating water is filtered in Water Filter FD-101A-E to prevent solids build-up in the circulating water. Condensate blowdown is filtered and is taken out downstream of the filter. However, the stream is not cooled and is split off before EB-101. Thus the heat load to the cooling tower is minimized.

From the Gas Cooler the gas stream then is boosted in pressure by the ID fan followed by a split of the gas into two streams. This design was developed to minimize the length of ducting operating at a slight vacuum and to minimize the temperature of the gas being recycled back to the boiler. The mass flow rate of the gas recirculation stream is about 52 percent of the flow rate of the product gas stream, which proceeds to the gas compression area. The recycle stream is sized to provide oxygen content of about 70 percent by volume in the oxidant stream supplying the boiler. The Gas Cooler minimizes the volumetric flow rate to, and the resulting power consumption of, the Flue Gas Compression equipment located downstream.

Three-Stage Gas Compression System:

Please refer to Figure 2.4.6 (Drawing D 12173-04002-0).

The compression section, where CO_2 is compressed to 365 psig by a three-stage centrifugal compressor, includes Flue Gas Compressor GB-101. After the aftercoolers, the stream is then chilled in a propane chiller to a temperature of –21 °F. Note that both the trim cooling water and water for the propane condenser comes from the cooling tower. At this pressure and temperature, about 80 mole percent of the stream can be condensed. The flash vapors contain approximately 80 weight percent of the inlet oxygen and nitrogen, but also about 13.7 weight percent of the CO_2. Therefore, a rectifier tower has been provided to reduce the loss of CO_2 to an acceptable level (about 6 weight percent). Then the pressure of the liquid is boosted to 2,000 psig by CO_2 Pipeline Pump GA-103. This stream is now available for sequestration or usage.

The volumetric flow to the compressor inlet is about 71,000 ACFM and only a single frame is required. The discharge pressures of the stages have been balanced to give reasonable power distribution and discharge temperatures across the various stages. They are:

- 1st Stage 28 psig
- 2nd Stage 108 psig
- 3rd Stage 365 psig

Power consumption for this large compressor has been estimated assuming adiabatic efficiency of 75 percent.

The hot gas from each stage is first cooled in an air cooler to 120 ^{o}F (Flue Gas Compressor 1st/ 2nd / 3rd Stage Aftercooler EC-101/2/3) and then further cooled by a water-cooled heat exchanger to 95 ^{o}F (Flue Gas Compressor 1st/ 2nd Stage Trim Cooler EA-101/2). The flue gas compressor 3rd stage cooler (EA-103) cools the gas to 90 ^{o}F to reduce the size of the dryers. Due to their large size, many of these heat exchangers consist of multiple shells. Because of highly corrosive conditions, the process side of the coolers must be stainless steel.

Because the flue gas stream leaving DA-101 is wet, some water condenses out in the three aftercoolers. The sour condensate is separated in knockout drums (FA-100/1/2/3) equipped with mist eliminator pads. Condensate from these drums is drained to the cooling tower or to waste water treatment. To prevent corrosion, these drums have a stainless steel liner.

Gas Drying:

Please refer to Figure 2.4.6 (Drawing D 12173-04002-0).

It is necessary to dry the CO_2 stream to meet the product specification. Flue gas leaving the 3rd stage discharge knockout drum (FA-103) is fed to Flue Gas Drier FF-101 A/B where additional moisture is removed. An alumina bed drier has been selected.

The performance of a fixed-bed drier improves as pressure increases. This favors locating the drier at the discharge of the compressor. However, as the operating pressure of the drier increases, so does the design pressure of the equipment. This favors low-pressure operation. But, at low pressure the diameter or number of the drier vessels grows, increasing the cost of the vessel. Having to process the recycle gas from the rectifier condenser cooling would also increase the diameter of the vessel. However, this is less than 10 percent of the forward flow. For this design the drier has been optimally located downstream of the 3rd stage compressor. The CO_2 Drier system consists of two vessels; FF-101 A/B. One vessel is on line while the other is being regenerated. Flow direction is down during operation and up during regeneration.

The drier is regenerated with the noncondensable vent gas from the rectifier after it exits the third stage discharge trim cooler in a simple once through scheme. During regeneration, the gas is heated in Regeneration Heater FH-101 before passing it through the exhausted drier. After regeneration, heating is stopped while the gas flow continues. This cools the bed down to the normal operating range. The regeneration gas and the impurities contained in it are vented to the atmosphere.

Regeneration of an alumina bed requires relatively high temperature and, because HP steam pressure may fluctuate, a gas-fired heater has been specified for this service.

Flue Gas Filter FD-102 has been provided at the drier outlet to remove any fines that the gas stream may pick up from the desiccant bed.

CO_2 Condensation and Stripping:

Please refer to Figure 2.4.6 (Drawing D 12173-04002-0).

From the CO_2 Drier, the gas stream is cooled down further to -21 °F with propane refrigeration in CO_2 Condenser EA-104 A-F. From EA-104 the partially condensed flue gas stream continuesontoCO$_2$ Rectifier DA-102.

At this pressure and temperature, 80 mole percent of the stream can be condensed. The flash vapors contain approximately 80 weight percent of the inlet oxygen and nitrogen, but also 12 weight percent of the CO_2. Therefore, as mentioned, a rectifier tower has been provided to reduce the loss of CO_2 to an acceptable level. The pressure of the liquid is boosted to 2,000 psig by CO_2 Pipeline Pump GA-103 for delivery to a sequestration or usage location.

The vapors in the feed to the rectifier contain the nitrogen and the oxygen that flashed from the liquid CO_2. To keep the CO_2 loss to the minimum, the rectifier also has an overhead condenser, CO_2 Rectifier Condenser EA-107. This is a floodback type condenser installed on top of the Rectifier. It cools the overhead vapor from the tower down to –48 °F. The condensed CO_2 acts as cold reflux in the CO_2 Rectifier.

Taking a slipstream from the inert-free liquid CO_2 from the Rectifier bottoms and letting it down to the Flue Gas Compressor 3rd stage suction pressure cools EA-107. At this pressure, CO_2 liquid boils at –55 °F thus providing the refrigeration necessary to condense some of the CO_2 from the Stripper overhead gas. The process has been designed to achieve at least 94 percent CO_2 recovery. The vaporized CO_2 from the cold side of EA-107 is fed to the suction of the Flue Gas Compressor 3rd stage.

Any system containing liquefied gas such as CO_2 is potentially subject to very low temperatures if the system is depressurized to atmospheric pressure while the system contains cryogenic liquid. If the CO_2 Rectifier (and all other associated equipment that may contain liquid CO_2) were to be designed for such a contingency, it would have to be made of stainless steel. However, through proper operating procedures and instrumentation such a scenario can be avoided and low temperature carbon steel (LTCS) can be used instead. Our choice here is LTCS. However, the condenser section will be made from stainless steel.

CO_2 Pumping and CO_2 Pipeline:

Please refer to Figure 2.4.6 (Drawing D 12173-04002-0).

The CO_2 product must be increased in pressure to 2,000 psig. A multistage heavy-duty pump (GA-103) is required for this service. This is a highly reliable derivative of an API-class boiler feed-water pump.

It is important that the pipeline pressure be always maintained above the critical pressure of CO_2 such that single-phase (dense-phase) flow is guaranteed. Therefore, pressure in the line should be controlled with a pressure controller and the associated control valve located at the destination end of the line.

Offgas:

Please refer to Figure 2.4.6 (Drawing D 12173-04002-0).

The vent gas from the CO_2 Rectifier overhead is at high pressure and there is an opportunity for power recovery using turbo-expanders. Because the gas cools down in the expansion process, there is also an opportunity for cold recovery. Heat recovery from the stream after let down via an expander was examined and it was determined that the amount of duty that could be recovered without the carbon dioxide in the stream freezing was small. Thus heat recovery could not be justified. The offgas leaves the Rectifier at –48 °F approximately. The refrigeration recovery to condense CO_2 was the best use for this cold since it also produces a reasonable temperature regeneration gas for the dryers.

2.4.2.2. Process Flow Diagrams

Two process flow diagrams are shown below for these systems:

- Figure 2.4.5 (Drawing D 12173-04001-0) Flue Gas Cooling PFD
- Figure 2.4.6 (Drawing D 12173-04002-0) CO_2 Compression and Liquefaction PFD

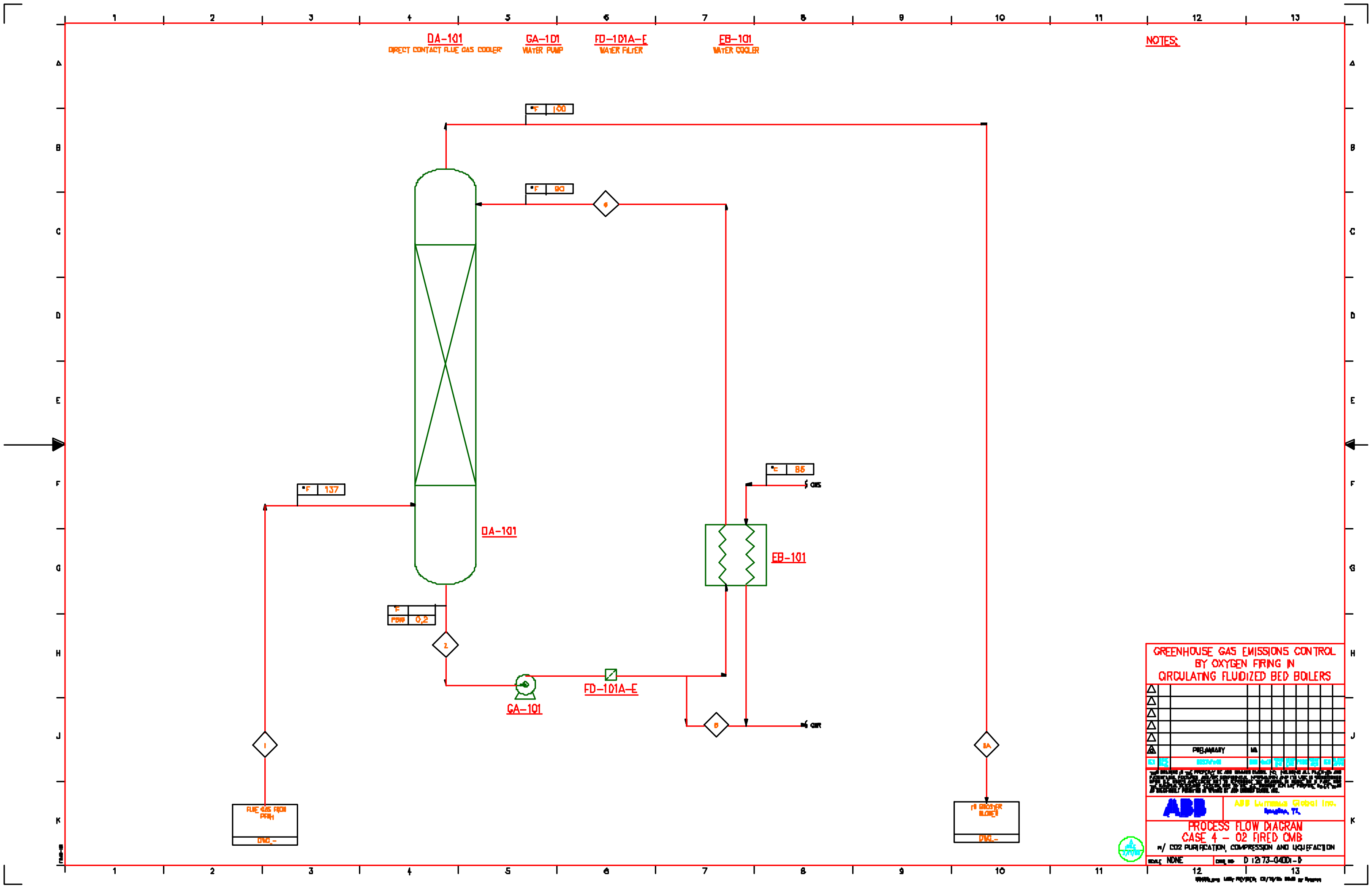


Figure 2.4. 5: Case-4 Flue Gas Cooling Process Flow Diagram

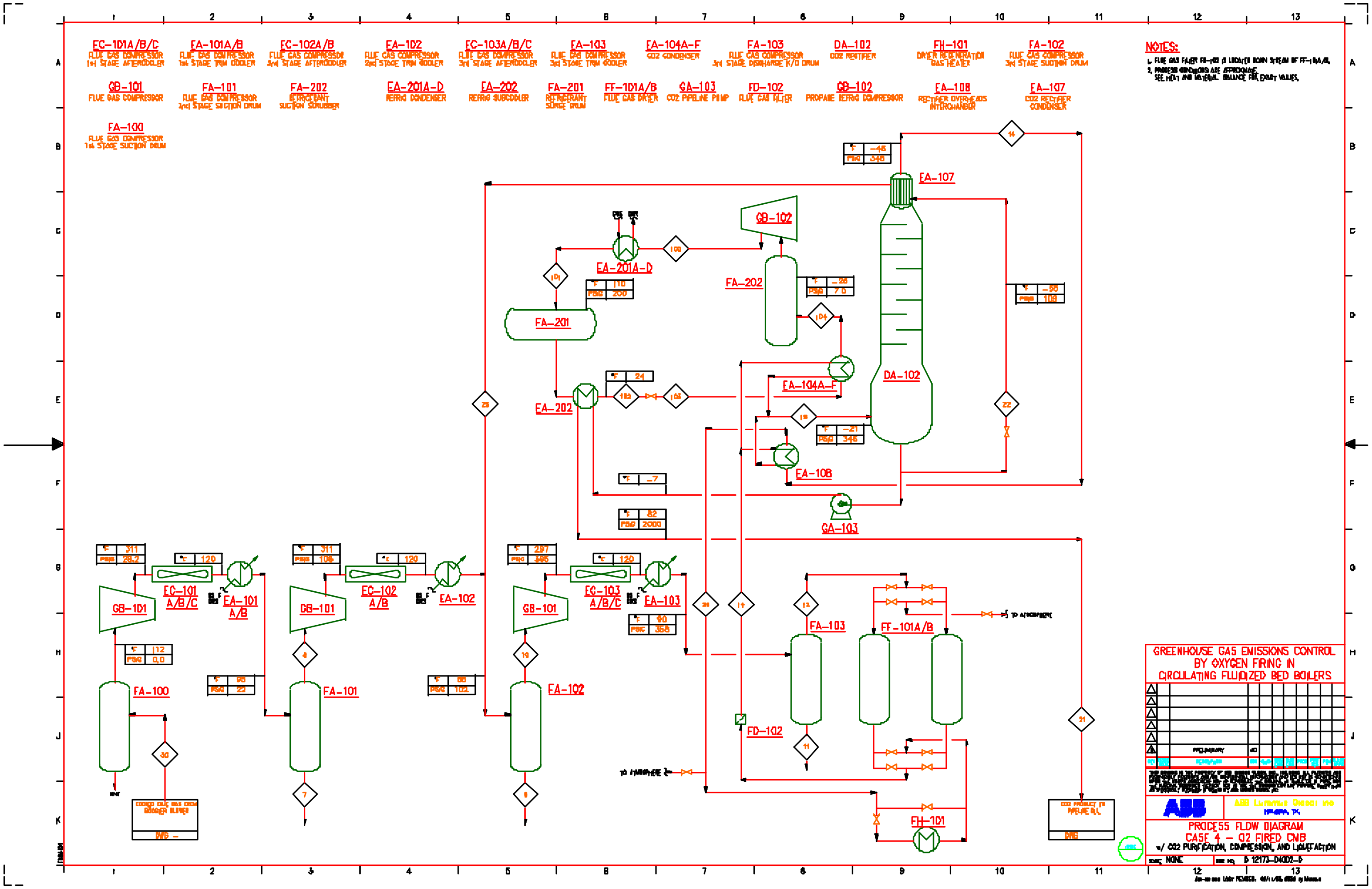


Figure 2.4. 6: Case-4 CO_2 Compression and Liquefaction Process Flow Diagram

2.4.2.3. Material and Energy Balance

The Case-4 Gas Processing System material and energy balance is very similar to Case-2 except the gas flow entering the system is slightly higher as reflected in the following Material and Energy Balance table.

Table 2.4. 2: Case-4 Gas Processing System Material & Energy Balance

STREAM NAME		To quench column	From quench column	Excess water	From blower	Quench water out	Quench water in	To liquefaction train	To boiler	First water KO	To second stage	2nd water KO	To third stage	Recycle from condenswer
PFD STREAM NO.		**1**	**3a**	**6**	**3b**	**2**	**5**	**3c**	**3d**	**7**	**8**	**9**	**10**	**25**
VAPOR FRACTION	Molar	0.989	1.000	0.000	1.000	0.000	0.000	1.000	1.000	0.000	1.000	0.000	1.000	1.000
TEMPERATURE	°F	136.0	99	99	113	117	90	113	113	95	95	86	86	-48
PRESSURE	PSIA	13.7	14	14	15	14	45	15	15	37	37	117	117	117
MOLAR FLOW RATE	lbmol/hr	18,234	15,643.96	2,590.29	15,643.96	96,555.95	94,000	10,297.55	5,346.41	475.16	9,822.39	12.78	10,453.36	790.00
MASS FLOW RATE	lb/hr	685,840	639,160	46,687	639,160	1,740,100	1,694,000	420,720	218,430	8,572	412,150	231	443,800	34,533
ENERGY	Btu/hr	8.18E+07	6.82E+07	-3.74E+07	7.03E+07	-1.36E+09	-1.37E+09	4.62E+07	2.40E+07	-6.88E+06	4.22E+07	-1.87E+05	4.32E+07	2.28E+06
COMPOSITON	Mol %													
CO2		71.38%	83.20%	0.03%	83.20%	0.02%	0.02%	83.20%	83.20%	0.09%	87.22%	0.30%	89.32%	97.54%
H2O		20.00%	6.75%	99.97%	6.75%	99.98%	99.98%	6.75%	6.75%	99.91%	2.25%	99.68%	0.60%	0.00%
Nitrogen		4.90%	5.71%	0.00%	5.71%	0.00%	0.00%	5.71%	5.71%	0.00%	5.98%	0.00%	5.71%	1.18%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		3.62%	4.22%	0.00%	4.22%	0.00%	0.00%	4.22%	4.22%	0.00%	4.43%	0.00%	4.25%	1.14%
SO2		0.10%	0.12%	0.00%	0.12%	0.00%	0.00%	0.12%	0.12%	0.01%	0.12%	0.02%	0.12%	0.14%
VAPOR														
MOLAR FLOW RATE	lbmol/hr	18,025.3	15,644.0	-	15,644.0	-	-	10,297.5	5,346.4	-	9,822.4	-	10,453.4	790.0
MASS FLOW RATE	lb/hr	682,080	639,160	-	639,160	-	-	420,720	218,430	-	412,150	-	443,800	34,533
STD VOL. FLOW	MMSCFD	164.17	142.48	-	142.48	-	-	93.79	48.69	-	89.46	-	95.20	7.20
ACTUAL VOL. FLOW	ACFM	139,420	114,110	-	108,590	-	-	71,481.45	37,112.63	-	25,965.84	-	8,434.13	447.36
MOLECULAR WEIGHT	MW	37.84	40.86	-	40.86	-	-	40.86	40.86	-	41.96	-	42.46	43.71
DENSITY	lb/ft³	0.08	0.09	0.09	0.10	-	-	0.10	0.10	0.26	0.26	-	0.88	1.29
VISCOSITY	cP	0.0145	0.0150	0.0150	0.0154	-	-	0.0154	0.0154	0.0155	0.0155	-	0.0156	0.0114
LIGHT LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID														
MOLAR FLOW RATE	lbmol/hr	208.94	-	2,590.29	-	96,555.95	94,000.00	-	-	475.16	-	12.78	-	-
MASS FLOW RATE	lb/hr	3,765	-	46,687	-	1,740,100	1,694,000	-	-	8,572.15	-	231.40	-	-
STD VOL. FLOW	BPD	258	-	3,204	-	119,400	116,240	-	-	588	-	16	-	-
ACTUAL VOL. FLOW	GPM	7.66	-	93.40	-	3,508.54	3,376.20	-	-	17.12	-	0.46	-	-
DENSITY	lb/ft³	61.32	-	62.32	-	61.84	62.56	-	-	62.44	-	62.74	62.74	-
VISCOSITY	cP	0.4793	-	0.6874	-	0.5707	0.7606	-	-	0.7185	-	0.8172	0.8172	-
SURFACE TENSION	Dyne/Cm	66.34	-	69.96	-	68.21	70.83	-	-	70.30	-	70.96	70.96	-

STREAM NAME		To drier	3rd water KO	From drier / condenser inlet	Condenser outlet	Non-condensable vent	Rectifier bottoms to condenser	CO2 to pipeline	Redfrig compressor discharge	Refrig condenser out	Refrig subcooler out	Refrig to CO2 condenser	Refrig from CO2 condenser	Warm non-condensable vent
PFD STREAM NO.		12	11	14	15	24	22	21	100	101	102	103	104	26
VAPOR FRACTION	Molar	1.000	0.000	1.000	0.206	1.000	0.130	0.000	1.000	0.000	0.000	0.237	1.000	1.000
TEMPERATURE	°F	90	90	90	-23	-48	-56	82	167	110	45	-28	-27	69
PRESSURE	PSIA	371	371	369	364	361	120	2,015	222	215	212	21	21	356
MOLAR FLOW RATE	lbmol/hr	10,419.63	33.73	10,390.85	10,140.85	1,358.92	790.00	8,209.05	9,300.00	9,300.00	9,300.00	9,300.00	9,300.00	1,358.92
MASS FLOW RATE	lb/hr	443,190	616	442,670	432,020	47,923	34,533	358,830	410,100	410,100	410,100	410,100	410,100	47,923
ENERGY	Btu/hr	4.03E+07	-4.86E+05	4.02E+07	-1.68E+07	3.58E+06	-2.22E+06	-2.76E+06	7.15E+07	6.91E+06	-1.12E+07	-1.12E+07	4.49E+07	4.97E+06
COMPOSITON	Mol %													
CO2		89.61%	0.87%	89.86%	89.86%	39.11%	97.54%	97.54%	0.00%	0.00%	0.00%	0.00%	0.00%	39.11%
H2O		0.28%	99.06%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		5.73%	0.00%	5.75%	5.75%	35.90%	1.18%	1.18%	0.00%	0.00%	0.00%	0.00%	0.00%	35.90%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%	0.00%
Oxygen		4.26%	0.00%	4.27%	4.27%	25.00%	1.14%	1.14%	0.00%	0.00%	0.00%	0.00%	0.00%	25.00%
SO2		0.12%	0.06%	0.12%	0.12%	0.00%	0.14%	0.14%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR														
MOLAR FLOW RATE	lbmol/hr	10,419.6	-	10,390.8	2,087.2	1,358.9	102.5	-	9,300.0	-	-	2,203.0	9,300.0	1,358.9
MASS FLOW RATE	lb/hr	443,190	-	442,670	79,955	47,923	4,281	-	410,100	-	-	97,146	410,100	47,923
STD VOL. FLOW	MMSCFD	94.90	-	94.64	19.01	12.38	0.93	-	84.70	-	-	20.06	84.70	12.38
ACTUAL VOL. FLOW	ACFM	2,424.00	-	2,437.63	375.82	241.80	56.03	-	3,840.43	-	-	7,622.81	32,248.61	342.07
MOLECULAR WEIGHT	MW	42.53	-	42.60	38.31	35.27	41.77	-	44.10	-	-	44.10	44.10	35.27
DENSITY	lb/ft³	3.05	-	3.03	3.55	3.30	1.27	-	1.78	-	-	0.21	0.21	2.34
VISCOSITY	cP	0.0165	-	0.0165	0.0145	0.0147	0.0117	-	0.0103	-	-	0.0066	0.0066	0.0183
LIGHT LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	-	-	8,053.66	-	687.50	8,209.05	-	9,300.00	9,300.00	7,097.02	-	-
MASS FLOW RATE	lb/hr	-	-	-	352,060	-	30,252	358,830	-	410,100	410,100	312,950	-	-
STD VOL. FLOW	BPD	-	-	-	29,120	-	2,507	29,682	-	55,421	55,421	42,293	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	662.46	-	52.70	895.69	-	1,774.07	1,571.17	1,094.38	-	-
DENSITY	lb/ft³	-	-	-	66.26	-	71.57	49.95	-	28.82	32.54	35.65	-	-
MOLECULAR WEIGHT	MW	-	-	-	43.72	-	44.00	43.71	-	44.10	44.10	44.10	-	-
VISCOSITY	cP	-	-	-	0.1641	-	0.2239	0.0559	-	0.0835	0.1188	0.1792	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	15.36	-	20.19	0.88	-	4.81	9.03	14.28	-	-
HEAVY LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	33.73	-	0.00	-	-	-	-	-	-	(0.03)	-	-
MASS FLOW RATE	lb/hr	-	616.28	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	42	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	1.22	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	62.84	62.84	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	0.7751	0.7751	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	70.20	70.20	-	-	-	-	-	-	-	-	-	-	-

2.4.2.4. Gas Processing System Utilities

The following tables define the cooling water, natural gas, and electrical requirements for the Gas Processing System.

Table 2.4. 3: Case-4 Gas Processing System Cooling Water and Fuel Gas Requirements

COOLING WATER

REV	Equipment TAG NO	SERVICE	No. Installed	DUTY MMBTU/HR	INLET TEMP, F	OUTLET TEMP, F	FLOWRATE LB/HR
	EA-101	FG Comp 1 stg trim cooler	1	6.82	85	103	378,788
	EA-102	FG Comp 1 stg trim cooler	1	3.73	85	103	207,071
	EA-103	FG Comp 1 stg trim cooler	1	3.91	85	103	217,172
	EA-201	Refrig Condenser	1	64.55	85	100	4,303,030
	EB-101	Water Cooler	1	50.91	85	105	2,545,455
		TOTAL COOLING WATER		129.91			7,651,515

FUEL GAS FUEL GAS VALUE BASIS: 930 BTU/SCF (LHV)

REV	Equipment TAG NO	SERVICE	ONLINE FACTOR	COMPR HP	HEAT RATE BTU/HP-HR	DUTY MMBTU/HR	EFFICIENCY %	FLOWRATE (Peak) MMSCFD	FLOWRATE (Peak) SCFH	FLOW (Avg) MMSCFD
	FH-101	Mole sieve regeneration	61%			4.60	80%	0.148	6,183	0.091
		TOTAL FUEL GAS				4.60		0.148	6,183	0.091

Table 2.4. 4: Case-4 Gas Processing System Electrical Requirements

Number of Trains	Item Number	Service	Number Operating per train	Power (ea) including 0.95 motor eff	Total all trains
				(kW)	(kW)
1	EC-101	Flue Gas Compressor 1st Stage Aftercooler	1	73	73
1	EC-102	Flue Gas Compressor 2nd Stage Aftercooler	1	61	61
1	EC-103	Flue Gas Compressor 3rd Stage Aftercooler	1	61	61
1	GB-101	1st Stage	1	5847	5847
1		2nd Stage	1	6032	6032
1		3rd Stage	1	6042	6042
1	GB-102	1st Stage	1	5243	5243
1		2nd Stage	1	3025	3025
1	GA-101	Water pump	1	130	130
1	GA-103	CO2 Pipeline pump	1	685	685
		Total			**27200**

2.4.2.5. Gas Processing System Equipment

The Case-4 Gas Processing System equipment description is identical to Case-2 and is not repeated here. Refer to section 2.2.2.3 for the relevant GPS equipment description.

The equipment list for the Gas Processing System is provided in Appendix I, Section 9.1.4.2.

2.4.3. Case-4 Air Separation Unit Process Description and Equipment

The Case-4 Air Separation Unit process description is identical to Case-2 and is not repeated here. Refer to section 2.2.3 for the relevant ASU process description.

2.4.3.1. Process Description and Process Flow Diagrams

The Case-4 ASU process description and process flow diagram is identical to Case-2 and is not repeated here. Refer to section 2.2.3.1 for the ASU process flow diagram.

2.4.3.2. Material and Energy Balance

The Case-4 ASU material and energy balance is identical to Case-2 and is not repeated here. Refer to section 2.2.3.2 for the ASU material and energy balance.

2.4.3.3. Air Separation Unit Utility Summary

The Case-4 ASU utilities are identical to Case-2 and are not repeated here. Refer to section 2.2.3.3

2.4.3.4. Air Separation Unit Equipment

The Case-4 Air Separation Unit equipment description is identical to Case-2 and is not repeated here. Refer to section 2.2.3.4 for the relevant ASU equipment description.

2.4.4. Case-4 Balance of Plant Equipment and Performance

The balance of plant equipment described in this section includes the steam cycle performance and equipment, the draft system equipment, the cooling system equipment, and the material handling equipment (coal, limestone, and ash). Refer to Appendix I for equipment lists and Appendix II for drawings.

2.4.4.1. Steam Cycle Performance

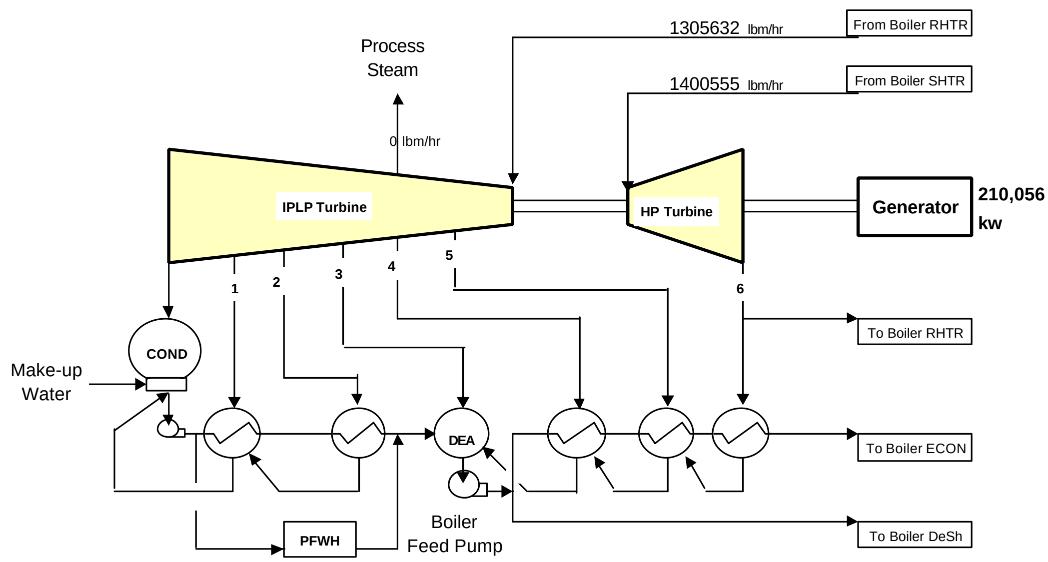
The steam cycle for Case-4 is shown schematically in Figure 2.4.7. The Mollier diagram which illustrates the process on enthalpy-entropy coordinates is the same as for Case-1 and is not repeated here.

The steam cycle arrangement and performance is very similar to Case-2 and 3. The steam cycle starts at the condenser hot well, which is a receptacle for the condensed steam from the exhaust of the steam turbine. The condensate flows to the suction of the condensate pumps, which increase the pressure of the fluid by a nominal 250 psi to transport it through the piping system and enable it to enter the open contact heater, or deaerator. The condensate passes through a gland steam condenser (SPE) first, followed in series by two low-pressure extraction feedwater heaters. The heaters successively increase the condensate temperature to a nominal 221°F by condensing and partially sub-cooling steam extracted from the LP steam turbine section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (now referred to as heater drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the condenser. The Case-4 condensate and feedwater system is arranged the same as for Cases 2 and 3 and differs from Case 1 in that there is a parallel low-pressure Feedwater Heater (PFWH - heated by flue gas) in a parallel feedwater stream with the two low-pressure extraction feedwater heaters as shown in Figure 2.4.7

The condensate entering the deaerator is heated and stripped of noncondensable gases by contact with the steam entering the unit. The steam is condensed and, along with the heated condensate, flows by gravity to a deaerator storage tank. The boiler feedwater pumps take suction from the storage tank and increase the fluid pressure to a nominal 2,200 psig. Both the condensate pump and boiler feed pump are electric motor driven. The boosted condensate flows through three more high-pressure feedwater heaters, increasing in temperature to 470°F at the entrance to the boiler economizer section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the deaerator.

Within the CMB boiler the feedwater is evaporated and finally superheated. The high-pressure superheated steam leaving the finishing superheater (1,400,555 lbm/hr of steam at 1,815 psia and 1,000 °F) is expanded through the high-pressure turbine. Reheat steam (1,305,632 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,000 °F. These conditions (temperatures, pressures) represent common steam cycle operating conditions for existing utility scale CFB power generation systems in use today. The reheated steam expands through the intermediate and low-pressure turbines before exhausting to the condenser. The condenser pressure used for Case-4 and all other cases in this study was 3.0 in. Hga.

The steam turbine performance analysis results show that the generator produces 210,056 kW output and the steam turbine heat rate is about 8,275 Btu/kWh. The generator output, turbine heat rate and condenser losses are slightly higher for Case-4 than for Case-2 and 3. This is a direct result of the slight increase in the PFWH heat absorption, which reduces extraction flows to the low-pressure extraction feedwater heaters and increases LP turbine power output.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Steam Turbine Power Output	729	Boiler Heat Input	1707
Process Steam Heat Loss	0	BFP & CP Input	12
Condenser Loss	991	**Total Energy Input**	1720
Total Energy Output	1720	**In - Out**	0

Turbine Heat Rate 8275 (Btu/kwhr)

Figure 2.4. 7: Case-4 Steam Cycle Schematic and Performance

2.4.4.2. Steam Cycle Equipment

This section provides a brief description of the major equipment (steam turbine, condensate and feedwater systems) utilized for the steam cycle of this case.

Steam Turbine:

The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the boiler passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheated steam flows through the reheat stop valves and intercept valves and enters the IP section at 465 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser.
The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for economic unit cycling.

Condensate and Feedwater Systems:

The function of the condensate system is to pump condensate from the condenser hot well to the deaerator, through the gland steam condenser and the LP feedwater heaters. The Case-4 condensate and feedwater system is very similar to Case 2 and 3 but differs from Case 1 in that there is a parallel low-pressure Feedwater Heater (PFWH) heated by flue gas in a parallel feedwater stream with the traditional extraction feedwater heaters. This PFWH is part of the Boiler scope of supply.

The system consists of one main condenser; two 50 percent capacity, motor-driven vertical condensate pumps; one gland steam condenser; two LP heaters, and one deaerator with a storage tank. Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump feedwater from the deaerator storage tank to the boiler economizer. Two motor-driven boiler feed pumps are provided to pump feedwater through the three stages of HP feedwater heaters. Pneumatic flow control valves control the recirculation flow. In addition, the suctions of the boiler feed pumps are equipped with startup strainers, which are utilized during initial startup and following major outages or system maintenance.

2.4.4.3. Other Balance of Plant Equipment

The systems for draft, solids handling (coal, limestone, and ash), cooling, electrical, and other BOP systems are described in this section for Case-4.

Draft System:

The flue gas is moved through the boiler, baghouse and other Boiler Island equipment with the draft system. The draft system includes the Gas Recirculation (GR) fans, the fluidizing gas blowers, the induced draft (ID) Fan, and the associated ductwork and expansion joints. This case has no traditional stack as the flue gas generated is supplied to the gas processing system where the CO_2 is purified and liquefied for sequestration or usage. The fans, and blowers are driven with electric motors and controlled to operate the unit in a balanced draft mode with the cyclone inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

Recirculated flue gas from the GR fan is mixed with oxygen from the ASU to provide a combustion oxidant stream, which is split into several flow paths.

Combustion gases exit the furnace and flow through a single inverted cyclone, which separates out ash and partially burned fuel particles. These solids are recycled back to the furnace, passing through J-valves, or seal pots, located below the cyclone. The solids leaving the seal pot are then returned directly to the combustor.

The gas exiting the cyclone passes directly to the tubular oxygen preheater (there is no convection pass for Case-4) and then exit the CMB steam generator to the baghouse for fine particulate capture. The flue gas leaving the baghouse is further cooled in a PFWH which is a low temperature economizer section and finally in a spray water cooler to about 100 ^{o}F. The gases are drawn through the CMB, baghouse, PFWH, and spray cooler with the Induced Draft Fan and then are recirculated to the CFB or discharged to the Gas Processing System.

The following fans and blowers are provided with the scope of supply of the Oxygen-fired CMB steam generator:

- Gas Recirculation fan; which provides recirculated flue gas to be mixed with oxygen from the ASU such that the mixed oxidant stream contains about 70 percent by volume oxygen. This fan is a centrifugal type unit, supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.4.5). The electric power required for the electric motor drive is 344 kW.

Table 2.4. 5: Gas Recirculation Fan Specification

Gas Analysis			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	182560	219072
Gas Inlet Temperature	(Deg F)	112.2	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	17.41	
Pressure Rise	(in wg)	75.0	97.5

- Induced draft fan; a centrifugal unit supplied with electric motor drive and inlet damper (see Table 2.4.6). The electric power required for the electric motor drive is 515 kW.

Table 2.4. 6: Induced Draft Fan Specification

Gas Analysis			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	642174	770609
Gas Inlet Temperature	(Deg F)	100.0	
Inlet Pressure	(psia)	13.64	
Outlet Pressure	(psia)	14.70	
Pressure Rise	(in wg)	29.5	38.4

- Fluidizing gas blowers; centrifugal units that provide recirculated flue gas for cooling and sealing the seal pots, and for assisting in the conveyance of cyclone bottoms (see Table 2.4.7). The electric power required for the electric motor drive is 209 kW.

Table 2.4. 7 Fluidizing Gas Blower Specification

Gas Analysis			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	37354	44825
Gas Inlet Temperature	(Deg F)	112.2	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	23.70	
Pressure Rise	(psia)	9.0	11.7

- Transport air blowers; centrifugal units that provide air for pneumatic transport of cool Bauxite from the MBHE bottom to the top of the combustor (see Table 2.4.8). The electric power required for the electric motor drive is 1,865 kW.

Table 2.4. 8 Transport Air Blower Specification

Gas Analysis			
Oxygen	(wt percent)	2.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	367188	440626
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	20.53	
Pressure Rise	(in wg)	161.5	210.0

Ducting and Stack:

There is no stack included in Case-4 as is true for Cases 2 and 3 also. The flue gas product leaving the Boiler Island, which is rich in CO_2, is delivered to the Gas Processing System (GPS), where the CO_2 stream is further purified for sequestration or usage. The impurities removed in the GPS, primarily nitrogen and oxygen, are vented to atmosphere.

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1/4" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the three silos.

Technical Requirements and Design Basis:

- Coal burn rate:
- – Maximum coal burn rate = 163,729 lbm/h = 81.9 tph plus 10 percent margin = 90 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
- – Average coal burn rate = 140,000 lbm/h = 70 tph (based on MCR rate multiplied by an 85 percent capacity factor)
- – Coal delivered to the plant by unit trains:
- – One and one-half unit trains per week at maximum burn rate
- – One unit train per week at average burn rate
- – Each unit train shall have 10,000 tons (100-ton cars) capacity
- – Unloading rate = 9 cars/hour (maximum)
- – Total unloading time per unit train = 11 hours (minimum)
- – Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
- – Reclaim rate = 300 tph
- – Storage piles with liners, run-off collection, and treatment systems:
- – Active storage = 6,600 tons (72 hours at maximum burn rate)
- – Dead storage = 50,000 tons (30 days at average burn rate)

Table 2.2. 9: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	90
Active Storage, tons	6,600
Dead Storage, tons	50,000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,000-ton silo to accommodate 3 days operation.

Bottom Ash Removal:

Bottom ash, or bed drain material, constitutes approximately two-thirds of the solid waste material discharged by the CFB steam generator. This bottom ash is discharged through a complement of two bed coolers (any one of which must be able to operate at 100 percent load on the design coal). The stripper/coolers cool the bed material to a temperature in the range of 300 ^oF (design coal) to a maximum of 500 ^oF (worst fuel) prior to discharge via rotary valves to the bed material conveying system. The steam generator scope terminates at the outlets of the rotary valves.

Fly Ash Removal:

Fly ash comprises approximately one-third of the solid waste discharged from the CMB steam generator. Approximately 8 percent of the total solids (fly ash plus bed material) is separated out in the oxygen heater hoppers; 25 percent of the total solids is carried in the gases leaving the steam generator en route to the baghouse. Fly ash is removed from the stack gas through a baghouse filter. Particulate conditions are as follows:

Design Specification for Particulate Removal System:

- Total solids to particulate removal system (stream 6, Figure 2.4.1) = 12,049 lbm/h
- Particle size distribution of particulate matter leaving cyclone (streams 3, 5, 6, Figure 2.4.1), see Table 2.4.10.

Table 2.4. 10: Particle Size Distribution

% Wt. Less	**Diameter** (Micron, µ)
100	192
99	160
90	74
80	50
70	37
60	30
50	24
40	16
30	12
20	8
10	4
1	< 4

- Solids leaving particulate removal system (stream 7, Figure 2.2.1) meet applicable environmental regulations, see Table 2.4.11.

Table 2.4. 11: Fly Ash Removal Design Summary

Design Parameter	**Value**
Flue Gas Temperature, ^{o}F	341
Flue Gas Flow Rate, lbm/h	685,849
Flue Gas Flow Rate, acfm	165,278
Particulate Removal, lbm/h	12,049
Particulate Loading, grains/acf	8.505

Ash Handling:

The function of the ash handling system is to provide the equipment required for conveying, preparing, storing, and disposing the bottom ash and fly ash that is produced on a daily basis by the boiler. The scope of the system is from the bag filter hoppers, oxygen heater hopper collectors, and bottom ash hoppers to the truck filling stations.

The fly ash collected in the bag filter and the oxygen heater is conveyed to the fly ash storage silo. A pneumatic transport system using low-pressure air from a blower provides the transport mechanism for the fly ash. Fly ash is discharged through a wet unloader, which conditions the fly ash and conveys it through a telescopic unloading chute into a truck for disposal.

The bottom ash from the boiler is drained from the bed, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. Ash from the fluidized bed ash coolers is drained to a complement of screw coolers, which discharge the cooled ash to a drag chain conveyor for transport to a surge bin. The latter is within the boiler scope of supply.

The cooled ash is pneumatically conveyed to the bottom ash silo from the surge bin. The silo is sized for a nominal holdup capacity of 36 hours of full-load operation (1,200 tons capacity). At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately 30 truck loads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.4.12: Ash Handling System Design Summary

Design Parameter	Value
Fly Ash from Baghouse, lbm/h	12,049
Ash from Boiler, lbm/h	65,500
Ash temperature, ^oF	520

Circulating Water System:

The function of the circulating water system is to supply cooling water to condense the main turbine exhaust steam. The system consists of two 50 percent capacity vertical circulating water pumps, a multi-cell mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Condenser Analysis:

The condenser system analysis is detailed in Table 2.4.13.

Table 2.4.13: Condenser Analysis

Item	Value	Units
Pressure	3.0	in. Hga
M stm-in	1,021,639	lbm/h
T stm-in	115.1	°F
P stm-in	1.474	psia
H stm-in	1051.7	Btu/lbm
M drain-in	76,525	lbm/h
H drain-in	89.7	Btu/lbm
H condensate	83	Btu/lbm
M condensate	1,098,164	lbm/h
Q condenser	990.8	10^6 Btu/h

Waste Treatment System:

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment is housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge de-watering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water are provided. A 200,000-gallon storage tank provides a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide distributed control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building

- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Plant Layout and Plot Plan:

The Case-4 plant is arranged functionally to address the flow of material and utilities through the plant site. A plan view of the boiler, power-generating components, and overall site plan for the entire plant is shown in Appendix II.

2.4.5. Case-4 Overall Plant Performance and CO_2 Emissions

Overall plant performance and emissions for Case-4 are summarized in Table 2.4.14. The Case-1 (Base Case) values are also listed along side for comparison purposes.

Boiler efficiency for Case-4 is calculated to be 93.66 percent (HHV basis) as compared to 89.46 percent for the Base Case. The improvement is primarily due to the reduced dry gas loss resulting from the oxygen firing. Refer to Section 2.2.5 for a discussion of why the dry gas loss is reduced with oxygen firing.

The steam cycle thermal efficiency including the boiler feed pump debit is about 41.25 percent as compared to 41.9 percent for Case-1. The slight reduction is due to some low level heat recovery, which is required with the oxygen-fired system.

The net plant heat rate and thermal efficiency for Case-4 are calculated to be 13,900 Btu/kWh and 24.6 percent respectively (HHV basis).

Auxiliary power for Case-4 is 77,888 kW (about 37.1 percent of generator output). The large auxiliary power increase, as compared to the Base Case, is due primarily to the large power requirement of the cryogenic based ASU and the gas compression requirement in the Gas Processing System of Case-4.

The resulting net plant output for Case-4 is 132,168 kW or about 68 percent of the Base Case output.

Carbon dioxide emissions for Case-4 are 27,659 lbm/hr or about 0.21 lbm/kWh on a normalized basis. This represents about 10 percent of the Case-1 normalized CO_2 emissions and a CO_2 avoided value of 1.79 lbm/kWh.

Table 2.4. 14: Case-4 Overall Plant Performance and Emissions

		CFB Air Fired (Case 1)	CMB Cryogenic O_2 Fired (Case 4)
Auxiliary Power Listing	(Units)		
Induced Draft Fan	(kW)	2285	515
Primary Air Fan	(kW)	2427	n/a
Secondary Air Fan	(kW)	1142	n/a
Fluidizing Air Blower	(kW)	920	209
Transport Air Fan	(kW)	n/a	1865
Gas Recirculation Fan	(kW)	n/a	344
Coal Handling, Preparation, and Feed	(kW)	300	294
Limestone Handling and Feed	(kW)	200	196
Limestone Blower	(kW)	150	147
Ash Handling	(kW)	200	196
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	152
Boiler Feed Pump	(kW)	3715	3715
Condensate Pump	(kW)	79	79
Circulating Water Pump	(kW)	1400	1889
Cooling Tower Fans	(kW)	1400	1889
Steam Turbine Auxiliaries	(kW)	200	207
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719
Transformer Loss	(kW)	470	472
Subtotal	(kW)	16007	12888
	(frac. of Gen. Output)	0.077	0.061
Auxiliary Power Summary			
Traditional Power Plant Auxiliary Power	(kW)	16007	12888
Air Separation Unit or Fuel Compressor	(kW)	n/a	37800
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a
CO_2 Removal System Auxiliary Power	(kW)	n/a	27200
Total Auxiliary Power	(kW)	16007	77888
	(frac. of Gen. Output)	0.077	0.371
Output and Efficiency			
Main Steam Flow	(lbm/hr)	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8275
OTM System Expander Generator Output	(kW)	n/a	n/a
Gas Turbine Generator Output		n/a	n/a
Steam Turbine Generator Output	(kW)	209041	210056
Net Plant Output	(kW)	193034	132168
	(frac. of Case-1 Net Output)	1.00	0.68
Boiler Efficiency (HHV)[1]	(fraction)	0.8946	0.9366
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855	1820
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	16.6
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855	1836

[1] Boiler Heat Output / (Qcoal-HHV + Qcredits)
[2] Required for GPS Desiccant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4

		CFB Air Fired (Case 1)	CMB Cryogenic O_2 Fired (Case 4)
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	13894
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.2456
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.69
CO_2 Emissions			
CO_2 Produced	(lbm/hr)	385427	379959
CO_2 Captured	(lbm/hr)	0	352380
Fraction of CO2 Captured	(fraction)	0.00	0.93
CO_2 Emitted	(lbm/hr)	385427	27579
Specific CO_2 Emissions	(lbm/kwhr)	2.00	0.21
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.10
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.79

2.5. Case-5: Air Fired CMB with CO_2 Capture utilizing Regenerative Carbonate Process

This section describes an advanced coal-fired power plant utilizing an atmospheric pressure Circulating Moving-Bed (CMB) type steam generator, providing steam for a subcritical steam plant, with the design modified to capture CO_2 within the fluidized bed. This CO_2 stream is then further processed in a Gas Processing System to produce a CO_2 product suitable for usage or sequestration. The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

The basic CO_2 capture concept for Case-5 is to create a high CO_2 content product stream within an air fired CMB type system. The high CO_2 content flue gas stream can then be further processed into a high purity CO_2 end product for various uses or sequestration. This concept utilizes an air fired CMB type system including a high temperature regenerative carbonate process to capture CO_2 from the products of combustion. All the energy rejected from the carbonate regeneration process is recovered in the steam cycle at high temperature such that there is no efficiency penalty associated with CO_2 capture for this process. The only significant energy penalty for this case is that associated with CO_2 compression in the Gas Processing System. Additionally, since the unit is air fired, it avoids the use of costly and energy intensive cryogenic type ASU systems as were used in Cases 2, 3, and 4. The trade off of course is a more complex boiler process. The concept is explained in detail below.

A brief performance summary for this plant reveals the following information. The Case-5 plant produces a net plant output of about 161 MW. The net plant heat rate and thermal efficiency are calculated to be 11,307 Btu/kWh and 30.2 percent respectively (HHV basis). Carbon dioxide emissions are about 0.01 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.5.4.

2.5.1. Case-5 Boiler Island Process Description and Equipment

This section describes the Boiler Island processes for Case-5 and includes a simplified process flow diagram (PFD), material and energy balance and equipment description. The equipment description includes only the major components included within the Boiler Island.

It should be emphasized that the CO_2 capture process described for this case is only conceptual at this time. A significant effort encompasing experimental work related to reaction rates, regeneration cycles, and fine particulate removal would be necessary to continue development of this capture process.

2.5.1.1. Process Description and Process Flow Diagrams

Figure 2.5.1 shows a simplified process flow diagram for the Boiler Island of the Case-5 air-fired CMB with CO_2 capture utilizing a regenerative carbonate process. Figures 2.5.2 and 2.5.3 show the equipment arrangement. This process description briefly describes the function of the major equipment and systems included within the Boiler Island. Complete data for all state points are shown in Table 2.5.1.

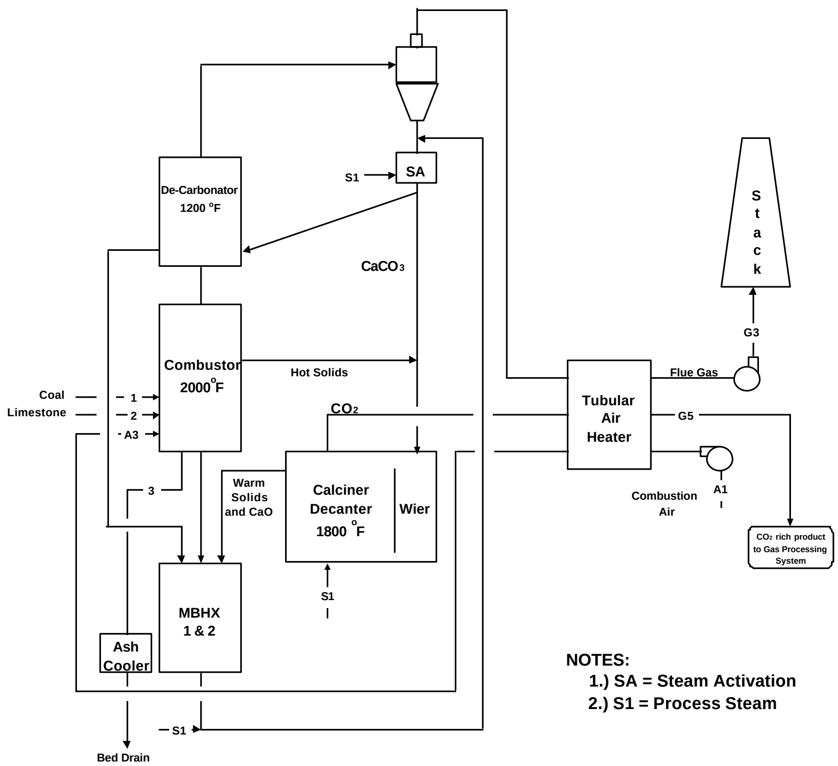


Figure 2.5. 1: Case-5: Simplified Boiler Island Gas Side Process Flow Diagram

In this concept coal or another high carbon content fuel (Stream 1) is reacted with air (Stream A3) and limestone (Stream 2) in the Combustor section of the Circulating Moving Bed (CMB) system at high temperature (~2,000 °F). The combustion air (Stream A3) is provided from the air fans and is preheated in the tubular air heater. There are three primary reactor areas in the Boiler Island system, the combustor, the de-carbonator, and the calciner. The purpose and operation of these reactors and the remaining Boiler Island equipment is described below.

Combustor:

The purpose of the Combustor is to combust the coal and any residual carbon contained in the recycled solids and also to heat solids for other process purposes. The combustor is air fired and is controlled to operate at about 2,000 °F. The flue gas leaving the combustor and entering the plenum area has a composition typical of any air fired coal combustor (e.g., Case-1). The combustor is designed to have a high superficial gas velocity in order to entrain most of the solids. The gas/solids mixture leaving the combustor enters a plenum area, where the gas velocity is reduced.

Combustor bed temperature is maintained at an optimum level for sulfur capture and combustion efficiency by controlling the flow of 1,200 °F solids leaving the De-Carbonator and entering the combustor.

Plenum:

The purpose of the plenum area is to drop out the coarse solid particles from the gas stream and transfer them to the moving bed heat exchanger and calciner. The plenum area is located in the gas flow stream between the combustor and the de-carbonator. The larger solid particles are disengaged from the gas as the gas velocity is decreased. These larger solids fall by gravity onto the moving bed heat exchanger (MBHE). Some of the particles also flow into the calciner. The flue gas and fine solids mixture leaving the plenum enter the de-carbonator.

De-Carbonator:

The purpose of the de-carbonator is to absorb most of the CO_2 contained in the incoming flue gas stream from the plenum. The CO_2 is absorbed with recycled CaO that is contained in the solids stream supplied from MBHE #2. The solids leaving MBHE#2 are originate from the solids stream drained from the Calciner. The de-carbonator is controlled to operate at about 1,200 ^{o}F. The CO_2 contained in the flue gas is removed in this area by reacting with CaO that is introduced into the reactor as described above. This reaction is shown below.

$$CaO + CO_2 \rightarrow CaCO_3 + Heat$$

The reactions are fast as indicated by tests conducted in-house. Conversion efficiency of the CaO to $CaCO_3$ is in excess of 20% per pass. The reaction is exothermic and the heat of reaction is transfered to the MBHE by heating the cool solids recycled from the bottom of the MBHE #2. The de-carbonator bed temperature is maintained at an optimum level for CO_2 capture by balancing solids flow between cool MBHE solids entering the de-Carbonator bed, and hot solids leaving the bed.

Particulate Removal:

The gas/solids mixture leaving the de-carbonator enters the particulate removal system. This is a very high efficiency particulate removal system, which utilizes a ring cone seperator where almost all the solids are disengaged from the flue gas. The solids removed from the flue gas are rich in $CaCO_3$. The flue gas leaving the particulate removal system is ducted to the tubular air heater where it is cooled by exchanging heat with the incoming combustion air. The solids stream leaving the particulate removal system is mixed with cool recycle solids from the MBHE's. The combined solids stream is then introduced to a proprietary Steam Activation (SA) process and finally split into two streams (1) a recycle stream to the de-carbonator and (2) a stream flowing to the calciner.

Calciner:

The calciner is designed to separate the captured CO_2 from the entering solids stream that is rich in $CaCO_3$, thereby regenerating the CaO. The calciner has two solids streams entering, a process steam stream (S1) entering, a regenerated solids stream leaving, and a captured CO_2 product stream leaving. The solids stream leaving the steam activation process and entering the calciner contains $CaCO_{3,}$ as described above. A second stream of hot solids (2,000 ^{o}F), from the combustor, is also introduced into the calciner to provide the heat for the reaction. The flow of this stream is controlled to maintain the calciner at about 1,800 ^{o}F. It is this stream of hot solids that provides the heat necessary for regeneration of the CaO. Under these conditions the following reaction occurs.

$$CaCO_3 + Heat \rightarrow CaO + CO_2$$

The CO_2 gas that is released in the calciner flows through a particulate removal device and then to the tubular air heater where it is cooled by exchanging heat with the incoming combustion air. The solids streams leaving the calciner and particulate removal device, at about 1,800 ^{o}F, are discharged into a second moving bed heat exchanger (MBHE #2) dedicated to cooling the calciner solids stream. For simplicity the PFD (Figure 2.5.1) shows only one MBHE (MBHE#1 & #2).

Moving Bed Heat Exchangers:

The heat removed from the solids streams flowing through the moving bed heat exchangers (MBHE's) is used to generate steam for the power cycle. The moving bed heat exchangers (MBHE #1 and #2) are designed to cool the entering solids streams by evaporating, superheating, and reheating steam for the power cycle. The MBHE's contain all the pressure parts for the steam cycle in the Boiler Island. The moving bed heat exchangers are not fluidized and contain several immersed tube bundles, which cool the hot solids that are supplied from several locations within the system. The tube bundles in the MBHE's utilize spiral-finned surface and include superheater, reheater, evaporator and economizer sections. Very high heat transfer rates are obtained in the MBHE's due to the conduction heat transfer mechanism between the solids and the tubes.

The solids streams leaving the MBHE's are transported to the top of the unit using steam as the transport medium, combined with solids leaving the particulate removal system, and introduced to a steam activation process. The solids leaving the steam activation process are split into two streams. One stream is introduced at the top of the de-carbonator to complete the cycle. The second stream flows to the calciner for regeneration.

Ash Removal:

Draining hot solids from the combustor through two water-cooled ash coolers controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is provided by feedwater from the final extraction feedwater heater of the steam cycle. The heated water leaving the ash cooler is then combined with water from the economizer located in the convection pass to feed the steam drum.

Process Steam:

Process steam is used in three places in the system, primarily for reactivation of sorbent solids and for transport of solids from the MBHE's discharge lines to the top of the unit to continue recirculation. The total amount of process steam is relativly small amounting to about 4.1 percent of the main steam flow. It is extracted from the steam turbine at the #4 feedwater heater extraction point at about 173 psia and 737 ^oF and sprayed down to saturation conditions with feedwater before usage.

2.5.1.2. Material and Energy Balance

Table 2.5.1 shows the Boiler Island material and energy balance for Case-5. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-5 simplified PFD for the Boiler Island (Figure 2.5.1). This performance was calculated at MCR conditions for this unit.

The MCR (maximum continuous rating) condition for Case-5 is defined as high-pressure turbine inlet conditions of 1,400,555 lbm/hr, 1,815 psia, and 1,000 ^oF and intermediate-pressure turbine inlet conditions of 1,304,737 lbm/hr 469 psia 1,000 ^oF. These conditions were very similar to those used for the Base Case (Case-1), differing only in the reheat flow, where for Case-1 1,305, 632 lbm/hr was used. The slight reduction in reheat steam flow for this case is the result of a small process steam extraction from the steam turbine for Case-5.

The Case-5 boiler was fired with 20 percent excess air, the same as was used for the Base Case. The resulting boiler efficiency calculated for Case-5 was 94.03 percent (HHV basis). The boiler efficiency is improved as compared to Case-1 as a result of a lower exit gas temperature.

Table 2.5. 1: Case-5 Boiler Island Gas Side Material and Energy Balance

Constituent	Units	Innputs				Outputs		
		1	2	A1	S1	G3	G5	3
		Coal	Limestone	Combustion Air	Steam	Flue Gas	CO_2 Product	Bed Drain
CO_2	(lb/hr)					18899	359088	
H_2	"	5850						
O_2	"	5178		379147		70487		
N_2	"	2392		1256320		1271276		
H_2O(gas)	"			21262	73157	132201	21243	
H_2O(liquid)	"	6538						
CaCO3	"		29927					629
CaO	"							9708
$CaSO_4$	"							16282
Carbon	"	101675						2033
S(solid)	"	3834						
Ash	"	38392	1575					39967
Coal	"	163859						
Total Gas & Liquid					73157	1492863	380331	
Total Solids		163859	31502					68620
Total Flow		**163859**	**31502**	**1656729**	**73157**	**1492863**	**380331**	**68620**
Temp	^oF	80	80	80	360	214	500	520
Press	(psia)	14.7	14.7	14.7	14.7	14.7	14.7	14.7
Hs (Sensible Heat)	(MMBtu/hr)				9.3	52.6	5.0	6.4
Hf (Heat of Formation)	"	-299.0	-166.1	-120.3	-422.5	-836.1	-1504.0	-399.0
Total Energy	"	-299.0	-166.1	-120.3	-413.2	-783.5	-1498.9	-392.5

2.5.1.3. Boiler Island Equipment

This section describes major equipment included in the Boiler Island for Case-5. The major components include the Combustor vessel, Plenum area, De-Carbonator vessel, Calciner vessel, particulate removal system, seal pots, fuel feed system, fuel silos, sorbent feed system, sorbent silo, external moving bed heat exchangers (MBHE #1 & #2), superheater, reheater, evaporator, economizer, ash coolers, high temperature air heater, and draft system.

Figures 2.5.2 and 2.5.3 show general arrangement drawings of the Case-5 CMB boiler. The plan area for the Case-5 Boiler Island is about 88 percent of that for Case-1. Similarly, the building volume for Case-5 is about 77 percent of that for Case-1. The complete Boiler Island Equipment List for Case-5 is shown in Appendix I. Appendix II shows several additional drawings of the Boiler (key plan view, boiler plan view, side elevation, and various sectional views).

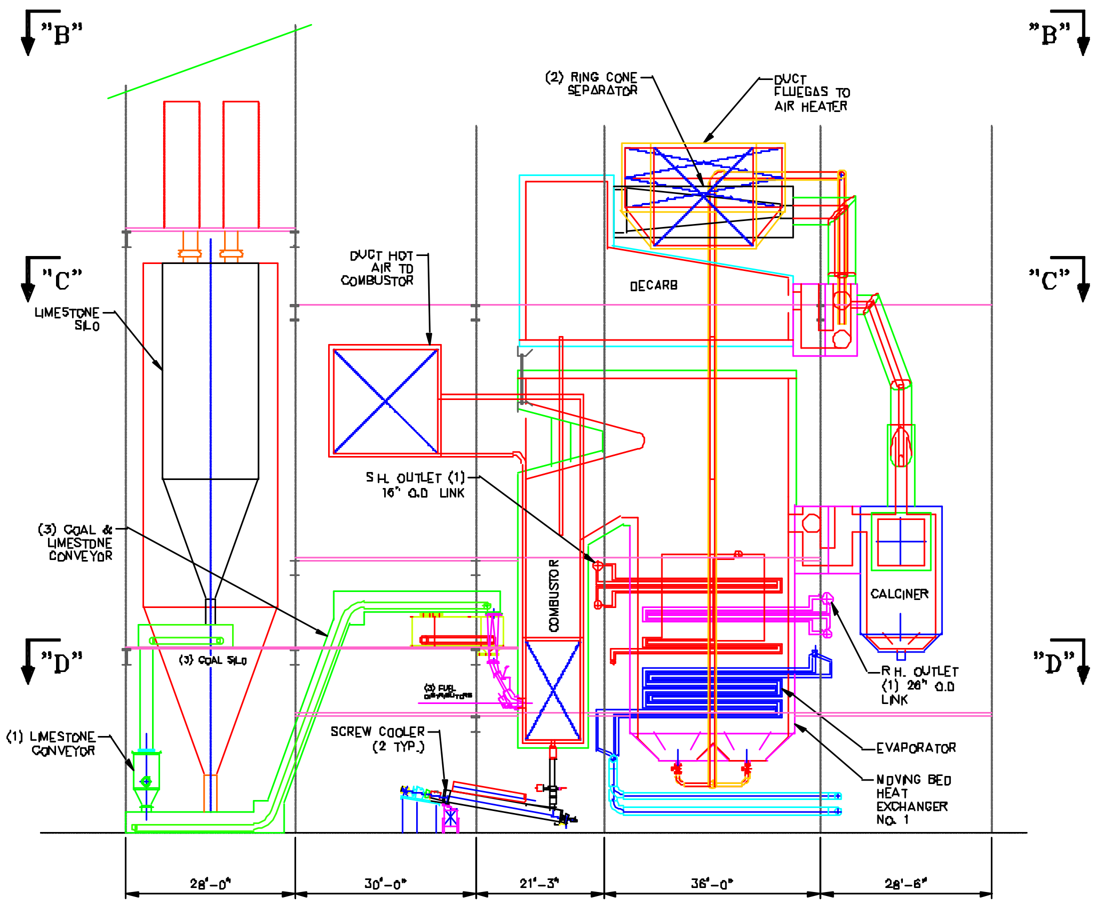


Figure 2.5. 2: Case-5 Boiler Island General Arrangement Drawing – Side Elevation

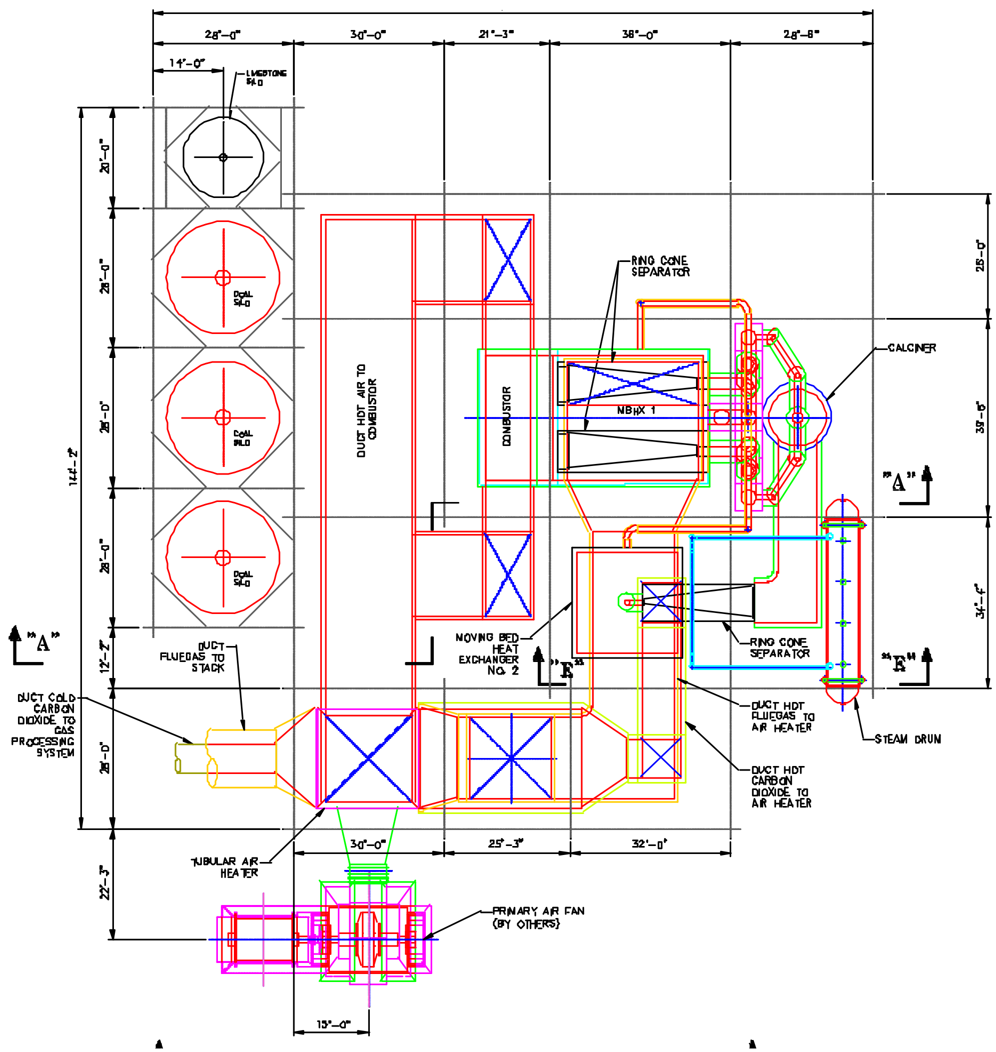


Figure 2.5. 3: Case-5 Boiler Island General Arrangement Drawing – Plan View

Combustor:

The combustor vessel is designed to react the oxygen contained in combustion air stream with the feed coal and the carbon contained in the recycle solids, thus producing a flue gas/solids mixture leaving the combustor and entering a plenum area. The combustor vessel for Case-5 is about 10 ft wide, 26 ft deep and 56 ft high. Crushed fuel, sorbent, recycle solids and combustion air are fed to the lower portion of the combustor.

The combustor vessel is constructed in the same fashion as the Case 4 and 6 combustors. It can be described as a rectangular refractory lined vessel with vertical walls and an arched outlet that leads to a plenum area. The lower and upper regions are formed with a multilayer refractory liner without any waterwall panels. The lower combustor has penetrations for the admission of fuel, sorbent, and recycle bed material. These penetrations are similar to those used for other cases in this study.

Plenum:

The plenum area is located directly above MBHE #1 and is designed to reduce the velocity of the gas/solids mixture stream leaving the combustor such that the coarse particulate drops out of the gas stream and onto the top of MBHE #1. The plenum area for Case-5 is a tapered area that starts at about 25 ft wide and 26 ft deep at the entrance and ends at about 44 ft wide and 26 ft deep at the exit and is about 17 ft high. The plenum is constructed with a multilayer refractory lining without any waterwall panels in a similar fashion as the combustor. The fine particulate remains entrained with the gas leaving the plenum and flows to the de-carbonator vessel.

De-Carbonator:

The de-carbonator is designed to absorb most of the CO_2 contained in the incoming flue gas stream with CaO containing recycle solids supplied from the MBHE. The de-carbonator vessel for Case-5 is a tapered vessel that starts at about 44 ft wide and 26 ft deep at the entrance and ends at about 14 ft wide and 26 ft deep at the exit and is about 28 ft high. This vessel requires penetrations for recycle solids from the MBHE and the particulate removal system, as well as a solids stream from the de-carbonator to the combustor.

The de-carbonator is constructed in the same fashion as the combustor. It can be described as a tapered rectangular refractory lined vessel with vertical walls. It is formed with a multilayer refractory liner without any waterwall panels.

Particulate Removal:

Flue gas and entrained solids exit the upper de-carbonator vessel and enter the ring cone separators. These are extremely high efficiency particle separation devices.

Calciner:

The calciner is designed to separate the captured CO_2 from the entering $CaCO_3$ rich stream thereby regenerating the CaO. The calciner vessel for Case-5 is cylindrical and has an inside diameter of about 12-ft and a height of about 24-ft. This vessel requires penetrations for recycle solids from the particulate removal system, as well as an entering steam stream.

The calciner is constructed in the same fashion as the combustor. It can be described as a cylindrical refractory lined vessel with vertical walls. It is formed with a multilayer refractory liner without any waterwall panels.

CO_2 and entrained solids exit the calciner vessel and enter the ring cone separator where the hot CO_2 is separated from the fine CaO particles.

Seal Pots:

The seal pots for Case-5 are of the same design as in other cases. The seal pot is a device that provides a pressure seal between the de-carbonator, which is at relatively high pressure, and the ring cone separator that is at near atmospheric pressure. The seal pot is a non-mechanical valve, which moves solids collected back to the de-carbonator. The seal pot is constructed of steel plate with a multiple layer refractory lining with fluidizing nozzles located along the bottom to assist solids flow. Some of the solids flow directly from the seal pot back to the de-carbonator while other solids are diverted through a hydrodynamic valve. The diverted solids flow through the calciner.

Fuel Feed System:

The fuel feed system for Case-5 is very similar to the system used for the other cases. It is designed to transport prepared coal from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, fuel feeders, feeder isolation valves, and fuel piping to the combustor.

Sorbent Feed System:

The limestone feed system for Case-5 is the same as for the other cases. The limestone feed system pneumatically transports prepared limestone from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, rotary feeders, blower, and piping from the blower to the reducer injection ports.

Ash Coolers:

The ash cooler design for Case-5 is the same as for the other cases as the ash flow is nearly identical in all cases except for Case-6. Draining hot solids from the combustor through two water-cooled ash coolers controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is provided by feedwater from the final extraction feedwater heater of the steam cycle. The heated water leaving the ash cooler is then combined with water from the economizer located in the convection pass to feed the steam drum.

Convection Pass:

There is no traditional convection pass containing pressure parts in Case-5, and the gas streams leaving the ring cone separators located at the outlet of the de-carbonator vessel and calciner are ducted directly to the High Temperature Air Heater for heat recovery.

Moving Bed Heat Exchangers:

There are two external moving bed heat exchangers for Case-5. The moving bed heat exchangers are not fluidized and contain several immersed tube bundles, which cool the hot solids leaving several areas before the cooled solids return to the upper part of the de-carbonator. The tube bundles in the MBHE's utilize spiral-finned surface and include superheater, reheater, evaporator and economizer sections. MBHE #1 is located directly below the plenum area and includes superheater, reheater, and evaporator sections. The MBHE #1 vessel for Case-5 is about 25 ft wide, 26 ft deep and 35 ft high. MBHE #2 is located to the side of MBHE #1 and includes evaporator and economizer sections. The MBHE #2 vessel for Case-5 is about 20 ft wide, 20 ft deep and 45 ft high. Very high heat transfer rates are obtained in the MBHE's due to the conduction heat transfer mechanism between the solids and tubes. The MBHE's are bottom supported and are constructed using steel plate refractory lined enclosure walls. They are rectangular in cross section with hopper shaped bottoms. The solids move through the beds by gravity at a design velocity of about 100 ft/hr. The cooled solids leaving the MBHE's are fed to the de-carbonator.

Superheater:

The superheater is divided into two major sections in MBHE #1. Saturated steam leaving the steam drum supplies the horizontal low temperature superheater. Steam leaving the low temperature section flows through a de-superheating spray station and thenontothe finishing superheater section. Both sections are located in MBHE #1. There are no superheater banks located in the convective pass for Case-5. The steam leaving the finishing superheater is piped to the high-pressure turbine where it is expanded to reheat pressure and then returned to the low temperature reheat section of MBHE #1.

Reheater:

The reheater, also located in MBHE #1, is designed as a single section. The steam is supplied to the reheater inlet header from the de-superheating spray station, which is fed from the high-pressure turbine exhaust. The reheater is a horizontal section comprised of spiral-finned tubing and located between the superheat finishing section and the low temperature superheat section. There are no reheater banks located in the convection pass for Case-5. The steam leaving the reheater is returned to the intermediate pressure turbine where it continues its expansion through the intermediate and low-pressure turbines before being exhausted to the condenser.

Evaporator:

The evaporator sections for Case-5 are located in the lower part of MBHE #1 and the upper part of MBHE #2. The evaporators are comprised of two banks of spiral-finned horizontal tubes, which evaporate high-pressure boiler feedwater. It is located just below the low temperature superheater section in MBHE #1 and at the top of MBHE #2. The water/steam mixture exiting the evaporator tube banks is supplied to the steam drum through risers where the steam and water phases are separated. The feedwater supplying the evaporators is piped from the steam drum through circulating water pumps and is comprised of a combination of separated saturated water and subcooled water from the economizer.

Economizer:

The economizer section for Case-5 is located in the lower part of MBHE #2. The economizer is comprised of four banks of spiral-finned horizontal tubes, which heat high-pressure boiler feedwater. The water exiting the economizer tube banks is supplied to the steam drum. The feedwater supplying the economizer is piped from the final extraction feedwater heater and the ash coolers.

Draft System:

The flue gas is moved through the Boiler Island equipment with the draft system. The draft system includes the primary air (PA) Fan, the induced draft (ID) Fan, the associated ductwork, and expansion joints. The ID and PA fans are driven with electric motors and are controlled to operate the unit in a balanced draft mode with the de-carbonator vessel outlet streams maintained at a slightly negative pressure (typically, -0.5 inwg).

High Temperature Air Heater:

A compact tubular regenerative air heater is used to cool the two gas streams leaving the de-carbonator and calciner by preheating the combustion air stream prior to combustion in the system. This is a high temperature segmented air heater and is considered a development item.

2.5.2. Case-5 Gas Processing System Process Description and Equipment

This purpose of this system is to process the flue gas stream leaving the Case-5 Boiler Island to provide a liquid CO_2 product stream of suitable purity for an EOR application.

The Case-5 CO_2 capture system is designed for about 95 percent CO_2 capture. This is possible for this case, since the CO_2 rich stream provided from the Boiler Island is comprised of only CO_2 and H_2O vapor. It does not have N_2 or O_2 impurities and, therefore, does not require the rectification process that was used in Cases 2, 4 and 6. Cost and performance estimates were developed for all the systems and equipment required to cool, clean, compress and liquefy the CO_2 to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO_2 specification for EOR, given previously in Table 2.0.1, was used as the basis for the CO_2 capture system design.

2.5.2.1. Process Description

The following describes a CO_2 recovery system that compresses and then cools a CO_2 rich gas stream captured from the calciner reactor section of an advanced air-fired CMB boiler to a pressure high enough so CO_2 can be liquefied. The resulting liquid CO_2 is pumped to a high pressure, so it can be economically transported for sequestration or usage. Pressure in the transport pipeline will be maintained above the critical pressure of CO_2 to avoid 2-phase flow.

The key process parameters (pressures, temperatures, duties etc.) are shown in the material and energy balance tables and will not be repeated here except in selected instances.

Figure 2.5.4 shows the Flue Gas Compression and Liquefaction process flow diagram.

Three-Stage Gas Compression System:

Please refer to Figure 2.5.4 (drawing D 12173-05001-0).

The compression section, where the CO_2 rich stream is compressed to 206 psia by a three-stage centrifugal compressor, includes Gas Compressor GB-2301. This three-stage compressor set includes a series of gas coolers (aftercoolers) located after each compression stage. Following the third stage aftercoolers, the stream is then further cooled in a propane chiller to a temperature of –25 °F. Note that both the trim cooling water and water for the propane condenser comes from the cooling tower. Following the compression and liquefaction steps, the pressure of the liquid is boosted to 2018 psia by CO_2 Pipeline Pump GA-2301. This stream is now available for sequestration or usage.

The volumetric flow to the compressor inlet is about 67,000 ACFM. The discharge pressures of the stages have been balanced to give reasonable power distribution and discharge temperatures across the various stages. They are:

- 1st Stage 35 psia
- 2nd Stage 90 psia
- 3rd Stage 206 psia

Power consumption for this large compressor has been estimated assuming adiabatic efficiency of 75 percent.

The hot gas stream from each compressor stage is first cooled in an air cooler to 120 ^oF in Flue Gas Compressor 1st/2nd/3rd Stage Air Coolers (EC-2301A-C, EC-2302A-B, EC-2303). The gas is then further cooled by water-cooled heat exchangers to 95 ^oF in Flue Gas Compressor 1st/2nd Stage Aftercoolers (EA-2301A/B and EA-2302). The gas compressor's 3rd stage Aftercooler (EA-2303) cools the gas to 90 ^oF to reduce the size of the dryers. Due to their large size, many of these heat exchangers consist of multiple shells. Because of highly corrosive conditions, the process side of the coolers must be stainless steel.

Because the flue gas stream leaving the Boiler Island is nearly saturated, some water condenses out in the three aftercoolers. The sour condensate is separated from the gas in knockout drums (FA-2300/1/2/4) equipped with mist eliminator pads. The condensate from these drums is drained to the cooling tower or to waste water treatment. To prevent corrosion, these drums have a stainless steel liner.

Gas Drying:

Please refer to Figure 2.5.4 (drawing D 12173-05001-0).

It is necessary to dry the CO_2 stream to meet the product specification. Flue gas leaving the 3rd stage discharge knockout drum (FA-2304) is fed to Flue Gas Drier PA-2351, where additional moisture is removed. A molecular sieve drier has been selected.

The performance of a fixed-bed drier improves as pressure increases. This favors locating the drier at the discharge of the compressor. However, as the operating pressure of the drier increases, so does the design pressure of the equipment. This favors low-pressure operation. But, at low pressure the diameter or number of the drier vessels grows, increasing the cost of the vessel. For this design the drier has been optimally located downstream of the 3rd stage compressor. The CO_2 Drier system consists of four molecular sieve beds. One vessel is on line while the others are being regenerated. The flow direction is down during operation and up during regeneration.

The drier is regenerated with drier outlet gas. After regeneration, heating is stopped while the gas flow continues. This cools the bed down to the normal operating range. The regeneration gas and the impurities contained in it are vented to the atmosphere after cooling and condensation of water vapor.

Regeneration of a molecular sieve bed requires relatively high temperature and, because HP steam pressure may fluctuate, a gas-fired heater has been specified for this service.

Flue Gas Filter FD-102 has been provided at the drier outlet to remove any fines that the gas stream may pick up from the desiccant bed.

CO_2 Condensation:

Please refer to Figure 2.5.4 (drawing D 12173-05001-0).

From the CO_2 Drier, the gas stream is cooled down further to -26 ^oF with propane refrigeration in CO_2 Condenser EA-2304A-F.

CO_2 Pumping and CO_2 Pipeline:

Please refer to Figure 2.5.4 (drawing D 12173-05001-0).

The CO_2 product must be increased in pressure to 2,000 psig. A multistage heavy-duty pump (GA-2301) is required for this service. This is a highly reliable derivative of an API-class boiler feed-water pump.

It is important that the pipeline pressure be always maintained above the critical pressure of CO_2 such that single-phase (dense-phase) flow is guaranteed. Therefore, the pressure in the line should be controlled with a pressure controller with the associated control valve located at the destination end of the line.

2.5.2.2. Process Flow Diagrams

One process flow diagram is shown below for these systems:

- Figure 2.5.4 (drawing D 12173-05001-0) CO_2 Compression and Liquefaction PFD

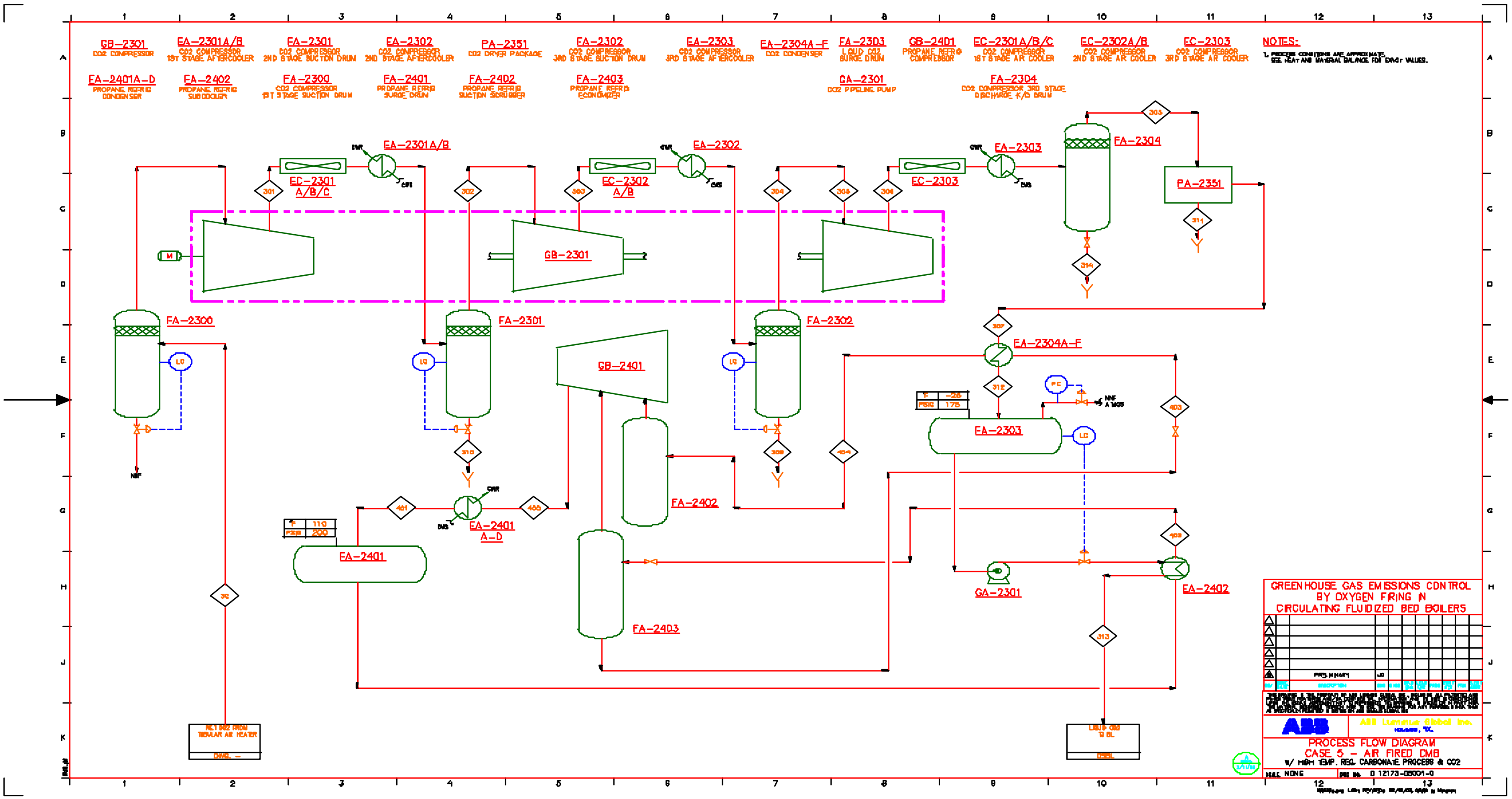


Figure 2.5. 4: Case-5 CO_2 Compression and Liquefaction Process Flow Diagram

2.5.2.3. Material and Energy Balance

Table 2.5.2 shows the material and energy balance for the Case-5 Gas Processing System.

Table 2.5. 2: Case-5 Gas Processing System Material & Energy Balance

STREAM NAME		To liquefaction	First stage discharge	To second stage	First stage water KO	2nd stage discharge	To 3rd stage	2nd stage water KO	From 3rd stage	To drier	3rd stage water KO
PFD STREAM NO.		**1**	**301**	**302**	**310**	**303**	**304**	**309**	**306**	**305**	**314**
VAPOR FRACTION	Molar	1.000	1.000	1.000	0.000	1.000	1.000	0.000	1.000	1.000	0.000
TEMPERATURE	°F	135.0	293	95	95	292	95	95	249	90	90
PRESSURE	PSIA	14.7	35	29	29	90	84	84	206	200	200
MOLAR FLOW RATE	lbmol/hr	9,338	9,338.47	8,398.21	940.25	8,398.21	8,244.09	154.13	8,244.09	8,192.78	51.31
MASS FLOW RATE	lb/hr	380,330	380,330	363,370	16,957	363,370	360,580	2,785	360,580	359,650	932
ENERGY	Btu/hr	4.43E+07	5.82E+07	3.66E+07	-1.36E+07	5.22E+07	3.54E+07	-2.23E+06	4.70E+07	3.36E+07	-7.44E+05
COMPOSITON	Mol %										
CO2		87.37%	87.37%	97.15%	0.07%	97.15%	98.96%	0.22%	98.96%	99.58%	0.54%
H2O		12.63%	12.63%	2.85%	99.93%	2.85%	1.04%	99.78%	1.04%	0.42%	99.46%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
SO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR											
MOLAR FLOW RATE	lbmol/hr	9,338.5	9,338.5	8,398.2	-	8,398.2	8,244.1	-	8,244.1	8,192.8	-
MASS FLOW RATE	lb/hr	380,330	380,330	363,370	-	363,370	360,580	-	360,580	359,650	-
STD VOL. FLOW	MMSCFD	85.05	85.05	76.49	-	76.49	75.08	-	75.08	74.62	-
ACTUAL VOL. FLOW	ACFM	67,234	35,735.79	28,430	-	12,407.03	9,448.75	-	4,915.57	3,732.54	-
MOLECULAR WEIGHT	MW	40.73	40.73	43.27	-	43.27	43.74	-	43.74	43.90	-
DENSITY	lb/ft³	0.09	0.18	0.21	0.21	0.49	0.64	-	1.22	1.61	-
VISCOSITY	cP	0.0148	0.0194	0.0149	0.0149	0.0210	0.0152	-	0.0202	0.0154	-
LIGHT LIQUID											
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID											
MOLAR FLOW RATE	lbmol/hr	-	-	-	940.25	-	-	154.13	-	-	51.31
MASS FLOW RATE	lb/hr	-	-	-	16,957.03	-	-	2,785.38	-	-	931.58
STD VOL. FLOW	BPD	-	-	-	1,164	-	-	191	-	-	64
ACTUAL VOL. FLOW	GPM	-	-	-	33.86	-	-	5.56	-	-	1.85
DENSITY	lb/ft³	-	-	-	62.44	-	-	62.48	-	-	62.70
VISCOSITY	cP	-	-	-	0.7185	-	-	0.7513	-	-	0.7831
SURFACE TENSION	Dyne/Cm	-	-	-	70.31	-	-	70.21	-	-	70.46

STREAM NAME		From drier/ To condenser	Water from drier	From condenser	From product pump	To pipeline	Refrig compressor discharge	From refrig condenser	From subcooler	Refrig to CO2 condenser	Refrig from CO2 condenser
PFD STREAM NO.		**307**	**311**	**312**	**308**	**313**	**400**	**401**	**402**	**403**	**404**
VAPOR FRACTION	Molar	1.000	0.726	0.000	0.000	0.000	1.000	0.000	0.000	0.209	0.992
TEMPERATURE	°F	90	380	-25	-10	82	167	110	51	-33	-33
PRESSURE	PSIA	195	195	190	2,018	2,015	222	215	212	19	19
MOLAR FLOW RATE	lbmol/hr	8,157.97	34.80	8,157.97	8,157.97	8,157.97	10,000	10,000	10,000	9,320.67	9,320.67
MASS FLOW RATE	lb/hr	359,030	627	359,030	359,030	359,030	440,960	440,960	440,960	411,010	411,010
ENERGY	Btu/hr	3.35E+07	7.64E+04	-2.44E+07	-2.20E+07	-4.17E+06	7.70E+07	7.44E+06	-1.04E+07	-1.42E+07	4.37E+07
COMPOSITON	Mol %										
CO2		100.00%	0.00%	100.00%	100.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%
H2O		0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
SO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR											
MOLAR FLOW RATE	lbmol/hr	8,158.0	25.3	-	-	-	10,000.0	-	-	1,947.1	9,250.6
MASS FLOW RATE	lb/hr	359,030	455	-	-	-	440,960	-	-	85,861	407,920
STD VOL. FLOW	MMSCFD	74.30	0.23	-	-	-	91.08	-	-	17.73	84.25
ACTUAL VOL. FLOW	ACFM	3,822.23	18.18	-	-	-	4,134.11	-	-	7,501.54	35,640.18
MOLECULAR WEIGHT	MW	44.01	18.02	-	-	-	44.10	-	-	44.10	44.10
DENSITY	lb/ft³	1.57	0.42	-	-	-	1.78	-	-	0.19	0.19
VISCOSITY	cP	0.0155	0.0154	-	-	-	0.0103	-	-	0.0065	0.0065
LIGHT LIQUID											
MOLAR FLOW RATE	lbmol/hr	-	-	8,157.97	8,157.97	8,157.97	-	10,000	10,000	7,373.58	70.02
MASS FLOW RATE	lb/hr	-	-	359,030	359,030	359,030	-	440,960	440,960	325,150	3,087.83
STD VOL. FLOW	BPD	-	-	29,786	29,786	29,786	-	59,593	59,593	43,941	417
ACTUAL VOL. FLOW	GPM	-	-	663.34	652.43	881.44	-	1,907.60	1,704.14	1,130.39	10.73
DENSITY	lb/ft³	-	-	67.48	68.61	50.78	-	28.82	32.26	35.86	35.86
MOLECULAR WEIGHT	MW	-	-	44.01	44.01	44.01	-	44.10	44.10	44.10	44.10
VISCOSITY	cP	-	-	0.1738	0.1593	0.0622	-	0.0835	0.1152	0.1849	0.1849
SURFACE TENSION	Dyne/Cm	-	-	15.92	13.90	0.86	-	4.81	8.64	14.66	14.66
HEAVY LIQUID											
MOLAR FLOW RATE	lbmol/hr	-	9.55	-	-	-	-	-	-	0.00	(0.00)
MASS FLOW RATE	lb/hr	-	172.02	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	12	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	0.40	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	53.74	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	0.1385	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	39.29	-	-	-	-	-	-	-	-

2.5.2.4. Gas Processing System Utilities

The following tables define the cooling water, natural gas, and electrical requirements for the Case-5 Gas Processing System.

Table 2.5. 3: Case-5 Gas Processing System Cooling Water and Fuel Gas Requirements

COOLING WATER

REV	Equipment TAG NO	SERVICE	No. Installed	DUTY MMBTU/HR	INLET TEMP, F	OUTLET TEMP, F	FLOWRATE LB/HR
	EA-101	FG Comp 1 stg trim cooler	1	7.27	85	103	404,040
	EA-102	FG Comp 1 stg trim cooler	1	3.73	85	103	207,071
	EA-103	FG Comp 1 stg trim cooler	1	3.36	85	103	186,869
	EA-201	Refrig Condenser	1	70.41	85	100	4,693,939
		TOTAL COOLING WATER		84.77			5,491,919

FUEL GAS FUEL GAS VALUE BASIS: 930 BTU/SCF (LHV)

REV	Equipment TAG NO	SERVICE	ONLINE FACTOR	COMPR HP	HEAT RATE BTU/HP-HR	DUTY MMBTU/HR	EFFICIENCY %	FLOWRATE (Peak) MMSCFD	FLOWRATE (Peak) SCFH	FLOW (Avg) MMSCFD
	FH-101	Mole sieve regeneration	72%			7.90	80%	0.255	10,618	0.183
		TOTAL FUEL GAS				7.90		0.255	10,618	0.183

Table 2.5. 4: Case-5 Gas Processing System Electrical Requirements

Number of Trains	Item Number	Service	Number Operating per train	Power (ea) including 0.95 motor eff (kW)	Total all trains (kW)
1	EC-101	Flue Gas Compressor 1st Stage Aftercooler	1	90	90
1	EC-102	Flue Gas Compressor 2nd Stage Aftercooler	1	49	49
1	EC-103	Flue Gas Compressor 3rd Stage Aftercooler	1	35	35
1	PA-2352	Drier Package	1	347	347
1	GB-101	1st Stage	1	4309	4309
1		2nd Stage	1	4824	4824
1		3rd Stage	1	3586	3586
1	GB-102	1st Stage	1	4606	4606
1		2nd Stage	1	4291	4291
1	GA-103	CO2 Pipeline pump	1	741	741
		Total			**22878**

2.5.2.5. Gas Processing System Equipment

The equipment list for the Case-5 Gas Processing System is provided in Appendix I, Section 9.1.5.2.

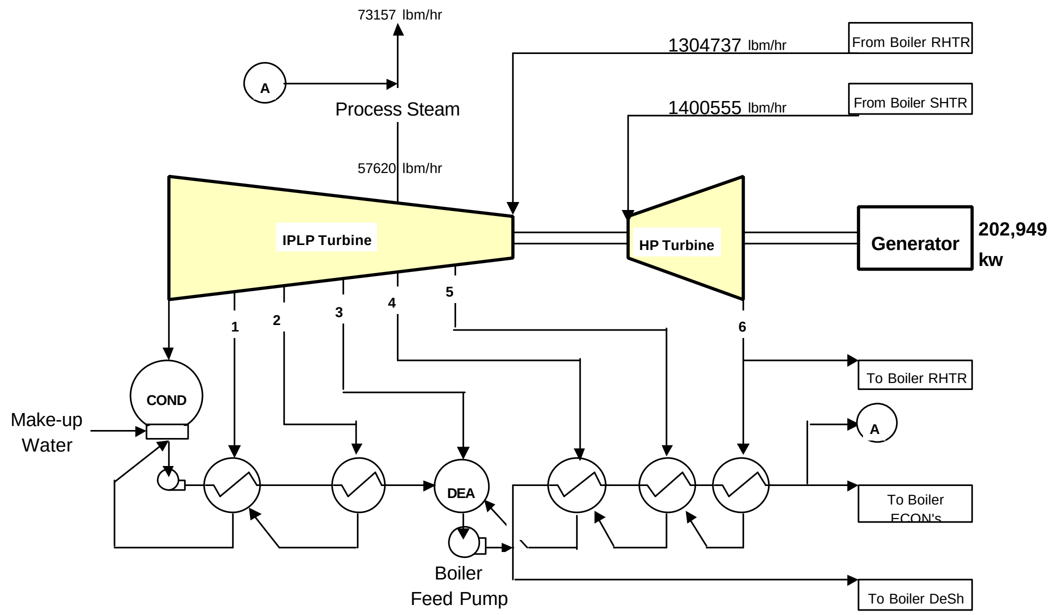
2.5.3. Case-5 Balance of Plant Equipment and Performance

The balance of plant equipment described in this section includes the steam cycle performance and equipment, the draft system equipment, the cooling system equipment, and the material handling equipment (coal, limestone, and ash). Refer to Appendix I for equipment lists and Appendix II for drawings.

2.5.3.1. Steam Cycle Performance

The steam cycle for Case-5 is shown schematically in Figure 2.5.5. The Mollier diagram which illustrates the process on enthalpy - entropy coordinates is the same as for Case-1 and is not repeated here. The steam cycle arrangement and performance is slightly different than Cases 2, 3, and 4. In this case a small amount of low-pressure process steam, which is required for solids transport and a steam activation process in the Boiler Island, is extracted from the low-pressure turbine and de-superheated. No process steam was used in Cases 1-4.

The high-pressure turbine expands 1,400,555 lbm/hr of steam at 1,800 psia and 1,005 ^{o}F. Reheat steam (1,304,737 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,005 ^{o}F. The condenser pressure used for Case-5 and all other cases in this study was 3.0 in. Hga. The steam turbine performance analysis results show the generator produces 202,949 kW output and the steam turbine heat rate is about 8,397 Btu/kWh. The generator output and condenser losses are slightly lower than for the other cases primarily due to the process steam requirement. Turbine heat rate is somewhat higher for Case-5 than other cases also as a result of the process steam requirement



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Steam Turbine Power Output	704	Boiler Heat Input	1673
Process Steam Heat Loss	81	BFP & CP Input	12
Condenser Loss	899	**Total Energy Input**	1685
Total Energy Output	1685	**In - Out**	0

Turbine Heat Rate 8396.7 (Btu/kwhr)

Figure 2.5. 5: Case-5 Steam Cycle Schematic and Performance

2.5.3.2. Steam Cycle Equipment

This section provides a brief description of the major equipment (steam turbine, condensate and feedwater systems) utilized for the steam cycle of this case.

Steam Turbine:

The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the boiler passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheated steam flows through the reheat stop valves and intercept valves and enters the IP section at 465 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. A small amount of steam is extracted for process steam as required in the Boiler Island for this case. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser.

The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for economic unit cycling.

Condensate and Feedwater Systems:

The function of the condensate system is to pump condensate from the condenser hot well to the deaerator, through the gland steam condenser and the LP feedwater heaters. The system consists of one main condenser; two 50 percent capacity, motor-driven vertical condensate pumps; one gland steam condenser; two LP heaters, and one deaerator with a storage tank.

Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump feedwater from the deaerator storage tank to the boiler economizer. Two motor-driven boiler feed pumps are provided to pump feedwater through the three stages of HP feedwater heaters. Pneumatic flow control valves control the recirculation flow. In addition, the suctions of the boiler feed pumps are equipped with startup strainers, which are utilized during initial startup and following major outages or system maintenance.

2.5.3.3. Other Balance of Plant Equipment

The systems for draft, solids handling (coal, limestone, and ash), cooling, electrical, and other BOP systems are described in this section for Case-5.

Draft System:

The flue gas is moved through the boiler, air heater and other Boiler Island equipment with the draft system. The draft system includes the primary air fans, the fluidizing air blowers, the induced draft (ID) Fan, the transport air blowers, the associated ductwork and expansion joints and the Stack, which disperses the flue gas leaving the system to the atmosphere. The induced draft, primary air fans, transport air blowers, and fluidizing air blowers are driven with electric motors and controlled to operate the unit in a balanced draft mode with the ring cone separator inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

Combustion gases exit the furnace and flow through ring cone separators, which separate out ash and partially burned fuel particles. These solids are recycled back to the furnace, passing through J-valves, or seal pots, located below the separators.

The gas exiting the ring cone separators passes directly to the tubular air preheater and then exits the CFB steam generator. The gases are drawn through the system with the Induced Draft Fan and then are discharged to atmosphere through the Stack.

The following fans and blowers are provided with the scope of supply of the CMB steam generator:

- Primary air fan, which provides forced draft primary airflow. This fan is a centrifugal type unit, supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.5.5). The electric power required for the electric motor drive is 2,015 kW.

Table 2.5. 5: Primary Air Fan Specification

Gas Analysis			
Oxygen	(wt percent)	22.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	1622958	1947550
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	16.00	
Pressure Rise	(in wg)	36.0	45.0

- The Transport Air Blower provides transport air for solids transport. This fan is a centrifugal type unit supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.5.6). The electric power required for the electric motor drive is 244 kW.

Table 2.5. 6: Transport Air Blower Specification

Gas Analysis			
Oxygen	(wt percent)	22.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	16567	19881
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	23.44	
Pressure Rise	(in wg)	242.0	314.6

- Induced draft fan, a centrifugal unit supplied with electric motor drive and inlet damper (see Table 2.5.7). The electric power required for the electric motor drive is 7,679 kW.

Table 2.5. 7: Induced Draft Fan Specification

Gas Analysis			
Oxygen	(wt percent)	4.72	
Nitrogen	"	85.16	
Water Vapor	"	8.86	
Carbon Dioxide	"	1.27	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	1492863	1791436
Gas Inlet Temperature	(Deg F)	150.0	
Inlet Pressure	(psia)	10.66	
Outlet Pressure	(psia)	14.70	
Pressure Rise	(in wg)	112.0	145.6

Ducting and Stack:

One stack is provided with a single 19.5-foot-diameter FRP liner. The stack is constructed of reinforced concrete, with an outside diameter at the base of 70 feet. The stack is 480 feet high for adequate dispersion of criteria pollutants, to assure that ground level concentrations are within regulatory limits. Table 2.5.8 shows the stack design parameters.

Table 2.5.8: Stack Design Summary

Design Parameter	Value
Flue Gas Temperature, ^{o}F	150
Flue Gas Flow Rate, lbm/h	1,492,863
Flue Gas Flow Rate, acfm	373,000
Particulate Loading, grains/acfm	nil

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1/4" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the three silos.

Technical Requirements and Design Basis

- Coal burn rate:
  - Maximum coal burn rate = 163,859 lbm/h = 81.9 tph plus 10 percent margin = 90 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
  - Average coal burn rate = 140,000 lbm/h = 70 tph (based on MCR rate multiplied by an 85 percent capacity factor)
  - Coal delivered to the plant by unit trains:
  - One and one-half unit trains per week at maximum burn rate
  - One unit train per week at average burn rate
  - Each unit train shall have 10,000 tons (100-ton cars) capacity
  - Unloading rate = 9 cars/hour (maximum)
  - Total unloading time per unit train = 11 hours (minimum)
  - Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
  - Reclaim rate = 300 tph
  - Storage piles with liners, run-off collection, and treatment systems:
  - Active storage = 6,600 tons (72 hours at maximum burn rate)
  - Dead storage = 50,000 tons (30 days at average burn rate)

Table 2.5. 9: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	92
Active Storage, tons	6,600
Dead Storage, tons	50,000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,000-ton silo to accommodate 3 days operation.

Bottom Ash Removal:

Bottom ash, or bed drain material, constitutes all of the solid waste material discharged by the CMB steam generator. This bottom ash is discharged through a complement of two bed coolers (any one of which must be able to operate at 100 percent load on the design coal). The stripper/coolers cool the bed material to a temperature in the range of 300 ^{o}F (design coal) to a maximum of 500 ^{o}F (worst fuel) prior to discharge via rotary valves to the bed material conveying system. The steam generator scope terminates at the outlets of the rotary valves.

Fly Ash Removal:

There is no significant amount of fly ash leaving the Boiler Island in this case. All ash collected in the ring cone separators (extremely high efficiency particulate collection devices) is recycled within the boiler island such that all ash leaves the system as bottom ash.

Ash Handling:

The function of the ash handling system is to convey, prepare, store, and dispose of the bottom ash produced on a daily basis by the boiler. The scope of the system is from the bottom ash hoppers to the truck filling stations.

The bottom ash from the boiler is drained from the ash coolers, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. Ash from the fluidized-bed ash coolers is drained to a complement of screw coolers, which discharge the cooled ash to a drag chain conveyor for transport to a surge bin. The ash is pneumatically conveyed to the bottom ash silo from the surge bin.
The silos are sized for a nominal holdup capacity of 36 hours of full-load operation (1,140 tons capacity) per each. At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately 30 truck loads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.5.10: Ash Handling System Design Summary

Design Parameter	Value
Ash from Boiler, lbm/h	68,620
Ash temperature, ^oF	520

Circulating Water System:

The function of the circulating water system is to supply cooling water to condense the main turbine exhaust steam. The system consists of two 50 percent capacity vertical circulating water pumps, a multi-cell mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Condenser Analysis:

The condenser system analysis is detailed in Table 2.5.11.

Table 2.5.11: Condenser Analysis

Item	Value	Units
Pressure	3.0	in. Hga
M stm-in	927040	lbm/h
T stm-in	115.1	°F
P stm-in	1.474	psia
H stm-in	1051.7	Btu/lbm
M drain-in	110,141	lbm/h
H drain-in	89.7	Btu/lbm
M drain-in	73,157	lbm/h
M make-up	83	Btu/lbm
H condensate	83	Btu/lbm
M condensate	1,094,841	lbm/h
Q condenser	899.4	10^6 Btu/h

Waste Treatment System:

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment is housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge de-watering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water are provided. A 200,000-gallon storage tank provides a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide distributed control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building

- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Plant Layout and Plot Plan:

The Case-5 plant is arranged functionally to address the flow of material and utilities through the plant site. A plan view of the boiler, power-generating components, and overall site plan for the entire plant is shown in Appendix II.

2.5.4. Case-5 Overall Plant Performance and CO_2 Emissions

Overall plant performance and emissions for Case-5 are summarized in Table 2.5.12. The Case-1 (Base Case) values are also listed along side for comparison purposes.

Boiler efficiency for Case-5 is calculated to be 94.03 percent (HHV basis) as compared to 89.46 percent for the Base Case. The improvement is primarily due to the reduced dry gas loss resulting from additional low-level heat recovery.

The steam cycle thermal efficiency including the boiler feed pump debit is about 40.65 percent as compared to 41.89 percent for Case-1. The slight reduction is due to the small amount of process steam, which is required with the Case-5 system.

The net plant heat rate and thermal efficiency for Case-5 are calculated to be 11,307 Btu/kWh and 30.19 percent respectively (HHV basis).

Auxiliary power for Case-5 is 41,764 kW (about 20.5 percent of generator output). The large auxiliary power increase, as compared to the Base Case, is due primarily to the large power requirement of the gas compression equipment in the Gas Processing System of Case-5.

The resulting net plant output for Case-5 is 161,184 kW or about 84 percent of the Base Case output.

Carbon dioxide emissions for Case-5 are 968 lbm/hr or about 0.01 lbm/kWh on a normalized basis. This represents less than 1 percent of the Case-1 normalized CO_2 emissions and a CO_2 avoided value of 1.99 lbm/kWh.

Table 2.5. 12: Case-5 Overall Plant Performance and Emissions

		CFB Air Fired (Case 1)	CMB Air Fired HT Carb (Case 5)
Auxiliary Power Listing	(Units)		
Induced Draft Fan	(kW)	2285	7679
Primary Air Fan	(kW)	2427	2015
Secondary Air Fan	(kW)	1142	n/a
Fluidizing Air Blower	(kW)	920	n/a
Transport Air Fan	(kW)	n/a	244
Gas Recirculation Fan	(kW)	n/a	n/a
Coal Handling, Preperation, and Feed	(kW)	300	293
Limestone Handling and Feed	(kW)	200	217
Limestone Blower	(kW)	150	163
Ash Handling	(kW)	200	205
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	n/a
Boiler Feed Pump	(kW)	3715	3756
Condensate Pump	(kW)	79	79
Circulating Water Pump	(kW)	1400	1436
Cooling Tower Fans	(kW)	1400	1436
Steam Turbine Auxilliaries	(kW)	200	187
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719
Transformer Loss	(kW)	470	456
Subtotal	(kW)	16007	18886
	(frac. of Gen. Output)	0.077	0.093
Air Separation Unit	(kW)	n/a	n/a
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a
CO2 Removal System Auxiliary Power	(kW)	n/a	22878
Total Auxilary Power	(kW)	16007	41764
	(frac. of Gen. Output)	0.077	0.206
Output and Efficiency			
Main Steam Flow	(lbm/hr)	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8397
OTM System Expander Generator Output	(kW)	n/a	n/a
Steam Turbine Generator Output	(kW)	209041	202949
Net Plant Output	(kW)	193034	161184
	(frac. of Case-1 Net Output)	1.00	0.84
Simplified Boiler Efficiency (HHV)[1]	(fraction)	0.8946	0.9217
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855	1815
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	7.9
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855	1822

[1] Boiler Heat Output / Qcoal (HHV)
[2] Required for GPS Desicant Regen in Cases 2-7 and ASU in Cases 2-4

Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	11307
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.3019
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.85
			359090.4
CO_2 Emissions			
CO_2 Produced	(lbm/hr)	385427	359998
CO_2 Captured	(lbm/hr)	0	359030
Fraction of CO2 Captured	(fraction)	0.00	1.00
CO_2 Emitted	(lbm/hr)	385427	968
Specific CO_2 Emissions	(lbm/kwhr)	2.00	0.01
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.00
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.99

2.6. Case-6: Oxygen Fired CMB with Oxygen Transport Membrane and CO_2 Capture

This section describes an oxygen-fired Circulating Moving Bed subcritical steam plant designed to produce a stack gas having a high concentration of CO_2. The plant utilizes an integrated Oxygen Transport Membrane (OTM) for the oxygen supply. The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

The basic concept for Case-6 is similar to Case-4 in that combustion air is replaced with oxygen in a CMB thereby creating a high CO_2 content flue gas stream. The oxygen used in conjunction with Case-6 is produced from an advanced integrated Oxygen Transport Membrane (OTM) system as compared to a commercial cryogenic ASU system used in Case-4. The concept is explained in more detail below.

A brief performance summary for this plant reveals the following information. The Case-6 plant produces a net plant output of about 197 MW. The net plant heat rate and thermal efficiency are calculated to be 11,380 Btu/kWh and 30.0 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 0.15 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.6.5.

2.6.1. Case-6 Boiler Island Process Description and Equipment

2.6.1.1. Process Description and Process Flow Diagrams

Figure 2.6.1 shows a simplified process flow diagram for the Boiler Island of the Case-6 oxygen-fired CMB concept. Selected mass flow rates (lbm/hr) and temperatures (^oF) are shown on this figure. Complete data for all state points are shown in Table 2.6.1. In this concept coal or another high carbon content fuel (Stream 1) is reacted with a stream containing about 70 percent by volume of oxygen (Stream 19) in the falling solids combustor section of the Circulating Moving Bed (CMB) system. The oxygen (Streams 27, 18, 19) is provided from an advanced Oxygen Transport Membrane (OTM) system.

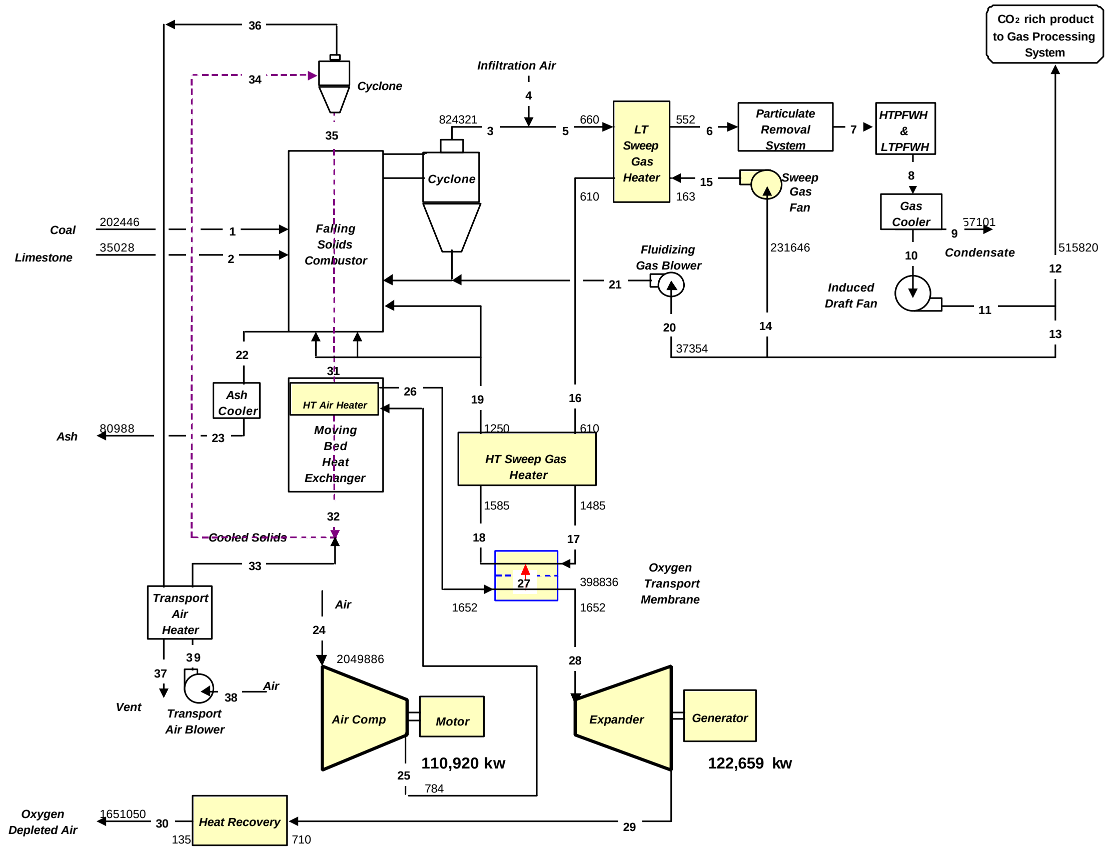


Figure 2.6. 1: Case-6 Simplified Boiler Island Gas Side Process Flow Diagram

The products of combustion leaving the combustor, flue gas comprised of primarily CO_2, H_2O vapor and unreacted hot solids with smaller amounts of N_2 and O_2, flow through a cyclone or another type of particulate removal device, where most of the entrained hot solids are removed and recirculated to the Combustor.

The temperature of the flue gas stream leaving the combustor and cyclone (Stream 3) is relatively cool (about 660 ^{o}F). The cooling of the combustor flue gas stream is identical to Case-4 and is accomplished by transferring heat from the flue gases in the combustor to a relatively cool stream of bauxite (Stream 35). The combustion products flow vertically up the combustor exchanging heat in a counter current fashion with the stream of bauxite solids flowing vertically down the combustor.

The bauxite is referred to as a “designer solid” in that it is optimally sized to accomplish this gas to solids and then solids to working fluid (steam/water) heat transfer in a highly effective manner. The bauxite particles are relatively dense, show very little attrition, and exhibit a high specific heat. Because of their size and density, they are not entrained with the gas and fine bed material but continue to fall to the bottom of the falling solids combustor. The bauxite particles are heated to about 2,000 ^{o}F by the time they reach the combustor bottom (Stream 31).

At this point they are transported in refractory lined connecting tubes into the Moving Bed Heat Exchanger (MBHE) located directly alongside the Combustor. An advantage of this system is that the heat transfer process to the steam cycle working fluid is completely separate from the combustion process. It also allows location of all the pressure parts in a common location (near ground level), thus minimizing interconnecting piping length and cost. The bauxite stream is cooled in the counterflow Moving Bed Heat Exchanger (MBHE) by exchanging heat with the power cycle working fluid (superheater, reheater, evaporator, and economizer), which is contained in spiral finned tubes within the MBHE enclosure walls. The bauxite particles leaving the MBHE (Stream 32) are at a temperature of about 530 ^{o}F. The bauxite particles are designed to be very free flowing as they move through the compact array of spiral finned tubing comprising the MBHE. Test results have confirmed this gravity-induced flowability of the particles against the spiral finned tube surfaces. The solids velocity in the MBHE is very slow (typically, 60-150 ft/hr) such that erosion and attrition is minimized.

The bauxite particles leaving the MBHE are then pneumatically transported in several parallel vertical pipes using hot air (Stream 34) as the transport medium back to the top of the combustor. At this location the hot air is separated from the bauxite in an array of small cyclones. The low temperature bauxite (Stream 35) then starts another cycle through the system.

The hot air leaving the small cyclones (Stream 36) is ducted to a tubular transport air heater provided to exchange heat with the incoming cool supply air (Stream 39). The cool air stream leaving the transport air heater (Stream 37) is then vented to atmosphere. The transport supply air (Stream 38) is boosted to the required pressure with the transport air blower. The pressure required for Stream 39 includes all system pressure drops including the static pressure required to lift the bauxite stream to the top of the combustor.

Draining hot bed solids through water-cooled fluidized bed ash coolers (Stream 22) controls solids inventory in the system, while recovering heat from the hot ash in an efficient manner. The cooling water used for the ash coolers is feedwater provided from the final extraction feedwater heater of the steam cycle.

The flue gas cooling system of Case-6, which is described below, is very similar to that of Case-2 and 4 except a traditional convection pass (used in Case-2) following the cyclone is not necessary due to the low gas temperature leaving the cyclone. The flue gas stream leaving the cyclone (Stream 3) is first cooled in a tubular Low Temperature Sweep Gas Heater (LTSGH).

The flue gas leaving the LTSGH (Stream 6) is cleaned of fine particulate matter in the baghouse and further cooled in a series of two Parallel Feedwater Heaters (HT-PFWH & LT-PFWH) by transferring heat to feedwater streams in parallel with both the HP and LP extraction feedwater heaters.

Finally, a direct contact water spray Gas Cooler is used to cool the gas before the flue gas enters the Induced Draft (ID) Fan (Stream 10). The gas cooler is used to cool the flue gas to the lowest temperature possible before recycling to minimize the power requirements for the boiler draft system (induced draft fan, fluidizing air blower, and sweep gas fan) and the product gas compression system. Some H_2O vapor is condensed in the Gas Cooler. This system is described in detail in Section 2.6.2 as it is considered a part of the Gas Processing System.

The flue gas leaving the ID Fan (Stream 11), comprised of mostly CO_2 and H_2O vapor with smaller amounts of O_2 and N_2, is split with about 48 percent of the flue gas going to the product stream (Stream 12) for further processing and the remainder recirculated to the CMB system. The quantity of recirculated gas (Stream 13) is about 52 percent of the product gas stream (Stream 12). About 86 percent of the recirculated gas provides the sweep gas for the OTM. After the sweep gas is mixed with the oxygen provided from the OTM the oxygen content in streams 18 and 19 is about 70 percent by volume. Stream 20, the remaining 14 percent of the recirculated flue gas, is used as fluidizing gas within the CMB system.

The OTM system includes a sweep gas system, and an air supply system. The sweep gas system includes a fan and two heat exchangers to preheat the sweep gas, which is provided to the OTM. The purpose of the purge gas (sweep gas) on the permeate side of the OTM is to increase the chemical potential, increase the oxygen flux, improve the overall process performance, and reduce the OTM cost. The air supply system includes an air compressor, high temperature air heater, gas expander, and heat recovery system.

The sweep gas system is described as follows. Stream 14 is referred to as sweep gas as it is ultimately used to sweep the oxygen from the OTM surface. The use of sweep gas helps to provide a low oxygen partial pressure on the low-pressure side of the membrane. This low oxygen partial pressure helps to minimize the cost of the OTM. The sweep gas is first pressurized in the sweep gas fan (Stream 15) to provide the required oxidant pressure entering the combustor. The pressurized sweep gas (Stream 15) is then heated to about 610 ^{o}F in the LTSGH and finally to about 1,485 ^{o}F in the tubular High Temperature Sweep Gas Heater (HTSGH). The sweep gas then picks up the oxygen (Stream 27) which was transported across the membrane.

The air supply system is described as follows. Atmospheric air (Stream 24) is first compressed in the electric motor driven air compressor to provide a high-pressure air stream, which ultimately supplies the OTM. The high air pressure is required to provide a high oxygen partial pressure on the airside of the OTM. This is advantageous because the OTM oxygen flux rate is proportional to the oxygen partial pressure difference across the membrane. The air leaving the compressor (Stream 25) is then heated in the High Temperature Air Heater (HTAH) to the temperature required by the membrane (about 1,652 ^{o}F). The high temperature air is supplied to the OTM and it gives up most (about 85 percent) of its oxygen (Stream 27) as the stream which crosses the membrane. The depleted oxygen stream (Stream 28) flows from the OTM to a gas expander. Stream 28 is then expanded to near atmospheric pressure in the gas expander while generating power in the associated generator. The gas expander power is greater than the air compressor power; and, therefore, the OTM system produces a net power output from the system. The gas stream leaving the expander (Stream 29) is then cooled against feedwater in two PFWH's (the HTPFWH is in a parallel water stream with the high-pressure extraction feedwater heaters and the LTPFWH is in parallel with the low pressure extraction feedwater heaters) before being discharged to the atmosphere. The oxygen transported through the membrane (Stream 27) is mixed with the heated sweep gas (Stream 17) to form the oxidant stream for the CMB combustor. The combustor oxidant stream (Streams 18 & 19) contains about 70 percent by volume of oxygen.

By using oxygen instead of air for combustion and by minimizing the amount of recirculated flue gas (sweep gas), the size and cost of many components (Combustor, Cyclone, Sweep Gas Heater, ductwork, fans and other equipment) can be reduced as compared to many other concepts for CO_2 capture with CFB systems as was shown previously in Case-2. Additionally, the OTM system eliminates the large power requirement of the cryogenic ASU, which improves overall plant efficiency significantly as shown in section 2.6.5.

2.6.1.2. Material and Energy Balance

Table 2.6.1 shows the Boiler Island material and energy balance for Case-6. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-6 simplified PFD for the Boiler Island (Figure 2.6.1). This performance was calculated at MCR conditions for this unit.

The MCR condition is defined as high-pressure turbine inlet conditions of 1,400,555 lbm/hr, 1,815 psia, and 1,000 ^{o}F and intermediate-pressure turbine inlet conditions of 1,371,446 lbm/hr 469 psia 1,000 ^{o}F. These conditions were very similar to those used for the Base Case and all other cases in this study although reheat flow was slightly higher in this case due to differences in low-level heat recovery arrangements. The boiler was fired with enough oxygen to leave about 3 percent by volume in Stream 3, the same as for the Base Case and Cases 2, 3, and 4. This oxygen requirement results in a stoichiometry of about 1.05 for Case-6. The resulting boiler efficiency calculated for Case-6 was 94.04 percent (HHV basis) with a LT sweep gas heater exit gas temperature of 552 ^{o}F and the LT-PFWH exit gas temperature of 135 ^{o}F. This boiler efficiency takes credit for the PFWH heat recovery.

Table 2.6. 1: Case-6 Boiler Island Gas Side Material and Energy Balance

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
O_2	(Lbm/hr)	6397		21896	4029	25925	25925	25925	25925		25925	25925	17039	8886	7652	7652
N_2	"	2956		11457	13346	24803	24803	24803	24803		24803	24803	16302	8501	7321	7321
H_2O	"	8078		80898	225	81123	81123	81123	81123	57101	24022	24022	15788	8234	7090	7090
CO_2	"			708629		708629	708629	708629	708629		708629	708629	465744	242885	209158	209158
SO_2	"			1440		1440	1440	1440	1440		1440	1440	947	494	425	425
H_2	"	7227														
Carbon	"	125618														
Sulfur	"	4737														
CaO	"															
$CaSO_4$	"															
$CaCO_3$	"		33276													
Ash	"	47433	1751													
		Coal	Limestone	Flue Gas	Infiltration Air	Flue Gas	Flue Gas	Flue Gas	Flue Gas	Condensate	Flue Gas	Flue Gas	Flue Gas	Recirc Gas	Sweep Gas	Sweep Gas
Total Gas	(Lbm/hr)			824321	17600	841921	841921	841921	841921		784820	784820	515820	269000	231646	231646
Total Solids	"	202446	35028													
Total Flow	"	202446	35028	824321	17600	841921	841921	841921	841921	57101	784820	784820 0	515820	269000	231646	231646 0
Temperature	(Deg F)	80.0	80.0	670.3	80	659.6	551.9	551.9	135.2	100	100.0	112.1	112.1	112.1	112.1	163.3
Pressure	(Psia)	14.7	14.700	14.700	14.700	14.447	14.231	13.779	13.707	14.700	13.635	14.700	14.700	14.700	14.700	19.892
hsensible	(Btu/lbm)			152.5	0.0	149.3	119.2	119.2	12.6		4.2	6.8	6.8	6.8	6.81	17.94
										19.960						
Chemical	10^6 Btu/hr	2241.9														
Sensible	10^6 Btu/hr	0.0	0.000	125.668	0.000	125.667	100.345	100.345	10.642	1.140	3.312	5.343	3.512	1.831	1.577	4.156
Latent	10^6 Btu/hr	0.000	0.000	84.943	0.237	85.180	85.180	85.180	85.180	0.000	25.223	25.223	16.578	8.645	7.445	7.445
Total Energy(1)	10^6 Btu/hr	2241.883	0.000	210.611	0.237	210.847	185.524	185.524	95.821	1.140	28.535	30.566	20.090	10.477	9.022	11.601

Constituent	(Units)	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
O_2	(Lbm/hr)	7652	7652	406488	406488	1234	1234			469219	469219	469219	398836	70383	70383	70383
N_2	"	7321	7321	7321	7321	1181	1181			1554428	1554428	1554428	0	1554428	1554428	1554428
H_2O	"	7090	7090	7090	7090	1143	1143			26239	26239	26239		26239	26239	26239
CO_2	"	209158	209158	209158	209158	33728	33728									
SO_2	"	425	425	425	425	69	69									
H_2	"															
Carbon	"							2512	2512							
Sulfur	"								0							
CaO	"							11187	11187							
$CaSO_4$	"							18105	18105							
$CaCO_3$	"								0							
Ash	"							49184	49184							
		Sweep Gas	Sweep Gas	Oxy + SG	Oxy + SG	Grease Gas	Grease Gas	Hot Ash Drain	Cool Ash Drain			OTM Air	Oxygen	OTM Exhaust	Heat Rec In	Heat Rec Out
Total Gas	(Lbm/hr)	231646	231646	630482	630482	37354	37354			2049886	2049886	2049886	398836	1651050	1651050	1651050
Total Solids	"				0			80988	80988							
Total Flow	"	231646	231646	630482	630482	37354	37354 0	80988	80988	2049886	2049886	2049886	398836	1651050	1651050	1651050
Temperature	(Deg F)	609.6	1484.8	1584.76	1249.8	112.1	194.6	2000.0	520.3	80.0	783.9	1652.0	1652.0	1652.0	709.9	135.2
Pressure	(Psia)	19.675	18.888	18.133	17.408	14.700	23.700	14.700	14.700	14.700	215.054	200.000	18.133	192.000	15.313	14.700
hsensible	(Btu/lbm)	127.26	380.09	384.53	291.64	6.81	24.92	545.33	95.39	0.00	175.44	410.70	387.11	416.40	158.98	13.74
Chemical	10^6 Btu/hr)							35.407	35.407							
Sensible	10^6 Btu/hr	29.480	88.047	242.438	183.871	0.254	0.931	44.166	7.726	0.000	359.640	841.881	154.392	687.489	262.479	22.682
Latent	10^6 Btu/hr	7.445	7.445	7.445	7.445	1.201	1.201	0.000	0.000	27.550	27.550	27.550	0.000	27.550	27.550	27.550
Total Energy(1)	10^6 Btu/hr	36.925	95.492	249.883	191.315	1.455	2.131	79.572	43.132	27.550	387.190	869.431	154.392	715.040	290.030	50.232

Constituent	(Units)	31	32	33	34	35	36	37	38	39
Air	(Lbm/hr)	0	0	476070	476070	0	476070	476070	476070	476070
Bauxite	"	4760698	4760698	0	4760698	4760698	0	0	0	0
Total Gas	(Lbm/hr)	0	0	476070	476070	0	476070	476070	476070	476070
Total Solids	"	4760698	4760698	0	4760698	4760698	0	0	0	0
Total Flow	"	4760698	4760698	476070	5236768	4760698	476070	476070	476070	476070
Temperature	(Deg F)	2000	530	430	521	521	521	240	80	147
Pressure	(Psia)			19.7	19.7	14.70	15.1	14.7	14.7	20.5
hsensible solids	(Btu/lbm)	545.3	97.8		95.6	95.6				
hsensible gas				85.997	108.647		108.647	39.058	0.000	16.408
Chemical	10^6 Btu/hr	0	0	0	0	0	0	0	0	0
Sensible	10^6 Btu/hr	2596.173	465.777	40.940	506.718	454.994	51.724	18.594	0.000	7.811
Latent	10^6 Btu/hr	0	0	0	0	0	0	0	0	0
Total Energy(1)	10^6 Btu/hr	2596.173	465.777	40.940	506.718	454.994	51.724	18.594	0.000	7.811

Notes:
(1) Energy Basis; Chemical based on Higher Heating Value (HHV); Sensible energy above 80F; Latent based on 1050 Btu/Lbm of water vapor

2.6.1.3. Boiler Island Equipment

This section describes major equipment included in the Boiler Island for Case-6. This equipment is very similar to what was described for Case-4 however, there are enough significant differences that it is also described here. Figures 2.6.2 and 2.6.3 show general arrangement drawings of the Case-6 CMB boiler. The complete Equipment List for Case-6 is shown in Appendix I. Appendix II shows several drawings of the Boiler (key plan, boiler plan view, side elevation, and various section views). The major components include the falling solids combustor, ash coolers, fuel feed system, sorbent feed system, bauxite recycle system, cyclone, seal pots, external moving bed heat exchanger (MBHE), superheater, reheater, economizer, sweep gas heater, baghouse, parallel feedwater heaters (PFWH's), gas cooler, and draft system.

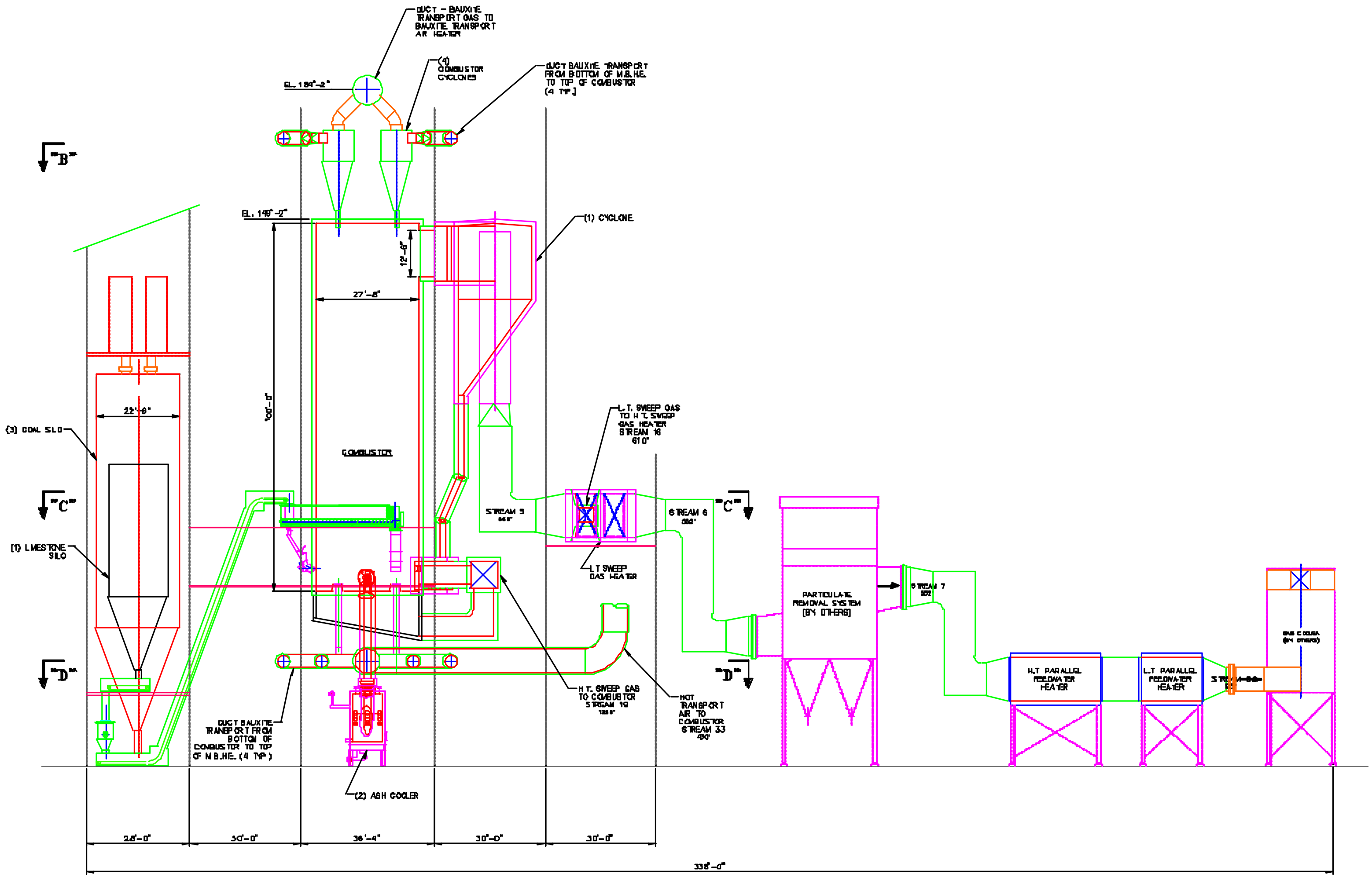


Figure 2.6. 2: Case-6 Boiler Island General Arrangement Drawing – Side Elevation

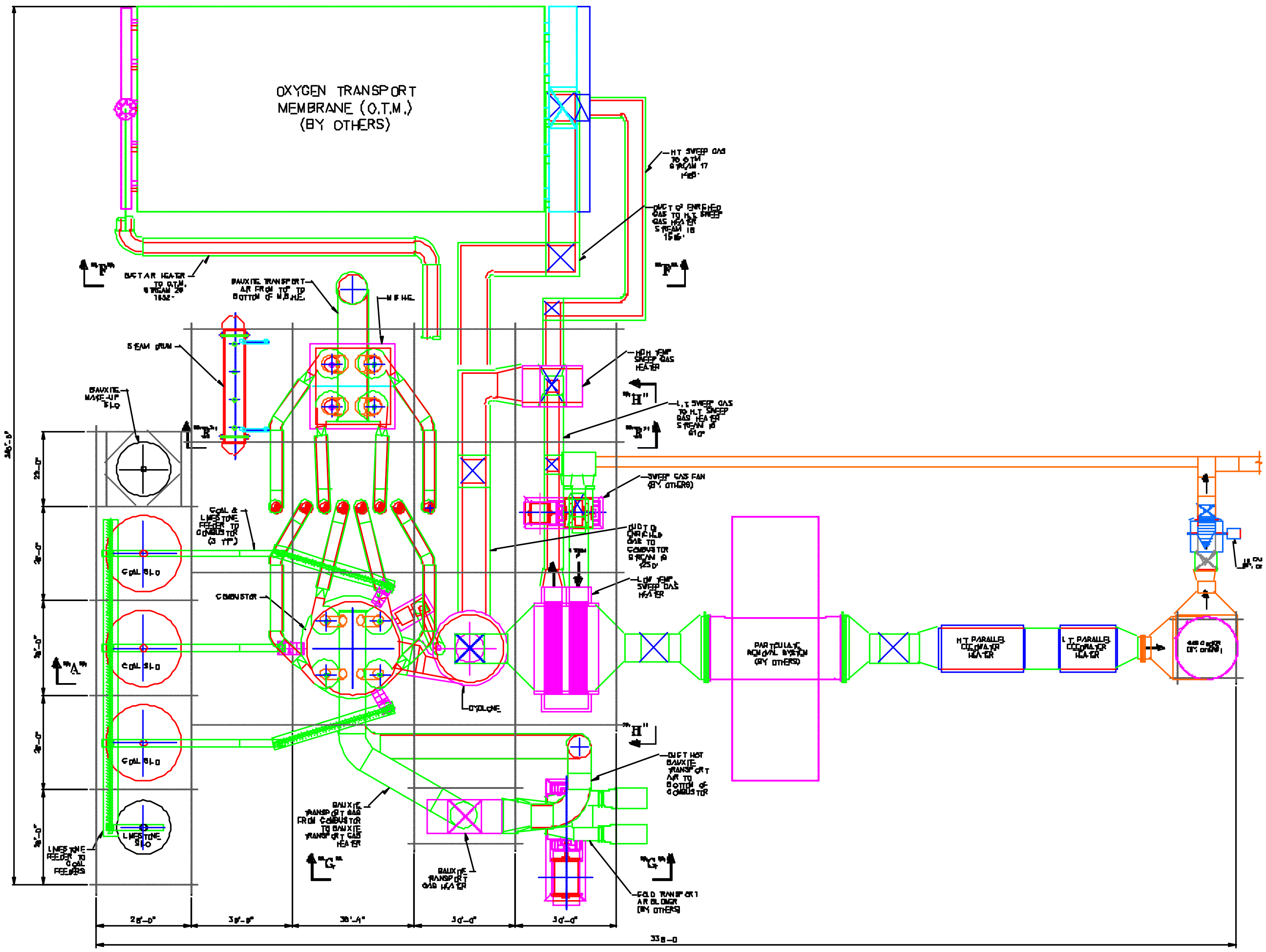


Figure 2.6. 3: Case-6 Boiler Island General Arrangement Drawing – Plan View

Falling Solids Combustor:

The combustor size is increased for Case-6 as compared to Case-4 due to increased coal input. The cylindrical combustor for Case-6 is about 28 ft in diameter and 100 ft high. Thus, the plan area for Case-6 is about 122 percent of the Case-4 plan area. This plan area increase is a nearly direct ratio of fuel heat input quantities. As compared to Case-1, the furnace plan area is about 43 percent as large. Crushed fuel, sorbent, and recycle solids are fed to the lower portion of the combustor. Primary "air" (actually a mixture of oxygen and recycled flue gas) is fed to the combustor bottom through a grid plate with secondary "air" supplied higher up in the lower combustor region. Cooled bauxite (530 ^oF) leaving the moving bed heat exchanger is transported to the top of the combustor where it is fed and distributed. The bauxite provides an intermediate heat transfer material. As the cool bauxite particles fall downward in the combustor and the hot flue gas and entrained bed material moves upward, in counter current fashion, the flue gas and bed material transfers some of their heat to the bauxite. The flue gas is cooled to about 660 ^oF at the combustor outlet (Cyclone inlet). Similarly, the bauxite particles, which exit the combustor at the bottom, are heated to about 2,000 ^oF. The bauxite particles are sized large enough that they are not entrained by the flue gas but small enough to provide the proper heat transfer in the combustor and MBHE. The hot bauxite particles leave the lower combustor region and then enter the moving bed heat exchanger, described below, where heat is transferred to the steam cycle working fluid.

The combustor is constructed significantly differently than the Case-1 combustor, or the Case-2/3 combustor but similar to Case-4. It can be described as a cylindrical refractory lined vessel with vertical walls. The lower and upper combustor regions are formed with a multilayer refractory liner without any waterwall panels. The lower combustor has penetrations for the admission of fuel, sorbent, recycled bed material, and oxidant. These penetrations are similar to those used for Case-1, 2, 3, and 4. Additionally, the hot bauxite must be removed from the lower combustor. This is done with a series of bauxite drain tubes connecting the lower combustor to the MBHE. Combustion occurs throughout the lower combustor, which is filled with bauxite particles and normal bed material. The upper combustor section is a cylindrical straight walled section formed with a multilayer refractory lining. The combustor flue gas is cooled exclusively by transferring heat to the bauxite particles, as mentioned above. The combustor bed temperature is maintained at an optimum level for sulfur capture and combustion efficiency by modulating the flow of cooled bauxite into the upper combustor. The bauxite stream entering the upper combustor is distributed evenly across the plan area to ensure proper heat transfer.

Fuel Feed System:

The fuel feed system for Case-6 is the same as for Case-1, 2, 3, and 4. It is designed to transport prepared coal from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, conveyors, fuel feeders, feeder isolation valves, and fuel piping to the furnace.

Sorbent Feed System:

The limestone feed system for Case-6 is the same as for Case-1. The limestone feed system pneumatically transports prepared limestone from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, rotary feeders, blowers, and piping from the blowers to the furnace injection ports.

Bauxite Recycle System:

The bauxite recycle system is designed to transport the cooled bauxite particles leaving the moving bed heat exchanger to the top of the falling solids combustor in an energy efficient manner. The particles are then fed into the combustor and provide an indirect heat transfer medium. The dense phase pneumatic system uses ambient air as a transport medium. The air is pressurized as required in the transport air blower and then preheated in the Transport Air Heater. The preheated air then transports the cooled bauxite particles leaving the MBHE to the top of the falling solids combustor. The bauxite particles are separated from the air in a simple cyclone, and then fed to the combustor. The air stream leaving the cyclone is cooled by exchanging heat with the cool incoming air stream in the Transport Air Heater and is then exhausted to atmosphere.

Ash Cooler:

The ash cooler design for Case-6 is the same as for Case-1. Draining hot solids through a water-cooled ash cooler controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is feedwater from the final extraction feedwater heater of the steam cycle.

Cyclone:

The flue gas and entrained solids exit the upper combustor at about 660 ^oF and enter the cyclone (inverted design). Only one cyclone is required for Case-6 because of the reduced gas flow resulting from the oxygen firing. The gas temperature is also significantly reduced. The cyclone is shaped like a cylindrical cone constructed from steel plate. The solids are separated from the flue gas in the cyclone and fall into a seal pot. Well over 99 percent of the entrained solids are captured in the cyclone. The flue gas leaving the cyclone is then ducted directly to the Oxygen Heater, as there is no convection pass required for Case-6.

Seal Pot:

The seal pot for Case-6 is of the same design as in Case-1 although smaller since less solids are recirculated due to the reduced gas flow in Case-6. The seal pot is a device that provides a pressure seal between the combustor, which is at high pressure, and the cyclone that is at atmospheric pressure. It is designed to move solids collected in the cyclone back to the combustor. The seal pot for Case-6 is constructed of steel plate with fluidizing nozzles located along the bottom. All of the solids in this case flow directly from the seal pot back to the combustor.

Convection Pass:

Because the temperature of the flue gas leaving the cyclone is so low in this case (660 ^oF), there is no convection pass and the flue gas leaving the cyclone is simply ducted directly to the Low Temperature Sweep Gas Heater.

Moving Bed Heat Exchanger:

The external heat exchanger for Case-6 is a moving bed as was used in Case-2 and Case-4 rather than a fluidized bed as was used in Case-1. The moving bed heat exchanger is not fluidized and contains several immersed tube bundles, which cool the hot bauxite particles leaving the lower combustor. The tube bundles in the MBHE are both bare tube and finned and include high temperature air heater, superheater, reheater, evaporator, and economizer sections. Very high heat transfer rates are obtained in the MBHE due to the conduction heat transfer between the solids and the tubes. The solids moving through the heat exchanger in this case are bauxite particles as opposed to typical bed material used in Case-2.

The MBHE is constructed using steel plate multilayer refractory lined enclosure walls. It is rectangular in cross section with a hopper shaped bottom. The solids move through the bed by gravity at about 150 ft/hr.

High Temperature Air Heater:

The high temperature air heater (HTAH) is located within the MBHE and is designed to heat the air leaving the air compressor of the OTM system to about 1,650 ^{o}F as required by the OTM. The heat for the air heating is provided from hot bauxite particles that flow by gravity from the Falling Solids Combustor into the MBHE. The HTAH is a tubular heat exchanger design with air inside the finned tubes and hot bauxite particles surrounding the outside of the tubes. The slowly moving bauxite particles leaving the HTAH then flow to the finishing superheater section located directly beneath the HTAH. The hot high pressure (~200 psia) air leaving the HTAH is supplied to the OTM for oxygen separation.

Superheater:

The superheater is divided into two major sections. Saturated steam leaving the steam drum supplies the horizontal low temperature section that is followed by the finishing superheater section located in the external moving bed heat exchanger. Both of these sections utilize spiral-finned tubes. There are no superheater banks located in the convection pass for Case-6, as there is no convection pass required in this design. Steam leaving the low temperature superheater is piped to the de-superheating spray station for outlet steam temperature control and then to the finishing superheater section. Steam leaving the finishing superheater is piped to the high-pressure turbine where it is expanded to reheat pressure and then returned to the low temperature reheat section of the MBHE.

Reheater:

The reheater is also divided into two sections, a low temperature section followed by a finishing section. The steam from the high-pressure turbine exhaust is supplied to the low temperature reheater. The low temperature and finishing sections are horizontal spiral-finned sections located in the MBHE. There are no reheater banks located in the convection pass for Case-6 as there is no convection pass required in this design. The steam leaving the low temperature reheater is piped to the de-superheating spray station and then to the finishing reheat section. Both sections are located in the external moving bed heat exchanger. The steam leaving the finishing reheater is returned to the intermediate pressure turbine where it continues its expansion through the intermediate and low-pressure turbines before being exhausted to the condenser.

Evaporator:

The evaporator section for Case-6 is located in the middle portion of the MBHE. The evaporator is comprised of three banks of horizontal spiral finned tubes, which evaporate high-pressure boiler feedwater. The water/steam mixture exiting the evaporator tube banks is supplied to the steam drum where the steam and water phases are separated. The separated steam supplies the low temperature superheater section. The feedwater supplying the evaporator is piped from the steam drum through circulating water pumps and is comprised of a combination of separated saturated water and subcooled water from the economizer.

Economizer:

The economizer section for Case-6 is also located in the lower MBHE. The economizer is comprised of two banks of horizontal tubes, which heats high-pressure boiler feedwater. The water exiting the economizer tube banks is supplied to the steam drum. The feedwater supplying the economizer is piped from the final extraction feedwater heater and the ash cooler.

Low Temperature Sweep Gas Heater:

A tubular regenerative low temperature sweep gas heater is used to cool the flue gas leaving the cyclone by pre-heating the sweep gas stream prior to supplying it to the high temperature sweep gas heater. The sweep gas is comprised of recirculated clean flue gas, which has been cooled against feedwater and increased to the required pressure by the ID fan and Sweep Gas fan. The sweep gas flows through the inside of the tubes and the flue gas, leaving the cyclone, flows over the outside of the tubes such that any ash contained in the flue gas stream can be easily removed from the outside of the tubes with conventional sootblowers.

High Temperature Sweep Gas Heater:

A tubular high temperature sweep gas heater is used to cool the oxidant stream being supplied to the CMB combustor while heating the sweep gas being supplied to the OTM. The sweep gas is comprised of recirculated clean flue gas from the low temperature sweep gas heater. The sweep gas flows through the inside of the tubes and the oxidant flows over the outside of the tubes.

Baghouse:

The fine particulate matter for Case-6 is removed from the cooled flue gas leaving the oxygen heater in a baghouse. The baghouse for Case-6 is much smaller than for Case-1 due to the reduced gas flow (about 43 percent of the Case-1 flow). The ash collected in the baghouse is supplied to the ash handling system. This system is described further in Section 2.6.4.2 under Balance of Plant Equipment.

Parallel Feedwater Heaters:

The Parallel Feedwater Heaters (PFWH) included in the Boiler Island of Case-6 are used to efficiently recover additional heat in the steam cycle for this case as shown in Figures 2.6.4 and 2.6.5. The feedwater flow for the high temperature PFWH (Figure 2.6.4) is in parallel with the top three extraction feedwater heaters (Heaters #4, #5 and #6) included in the steam cycle. The feedwater flow for the low temperature PFWH (Figure 2.6.5) is in parallel with the bottom two extraction type feedwater heaters (Heaters #1 and #2) included in the steam cycle.

The PFWH's are used because in Case-6 the gas temperature leaving the LT Sweep Gas Heater (LTSGH) is significantly higher than the gas temperature leaving the Air Heaters of Case-1 (552 ^{o}F Vs 275 ^{o}F). This occurs because the ratio of air to flue gas is higher in Case-1 than the ratio of sweep gas to flue gas in Case-6 making the Air heater of Case-1 more effective than the LTSGH of Case-6.

The PFWH heat exchangers are constructed similarly to economizer heat exchanger banks used in Heat Recovery Steam Generator units. The tubes used are heavily finned since the gas is clean and the enclosure walls are insulated steel liners. The two PFWH units cool the flue gas to about 135 ^{o}F.

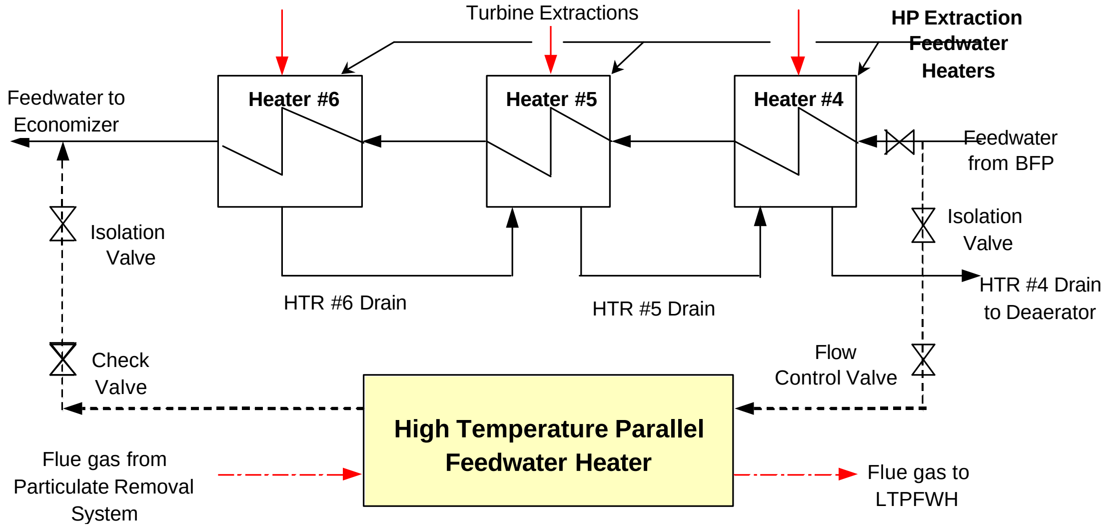


Figure 2.6. 4: Case-6 High Temperature Parallel Feedwater Heater Arrangement

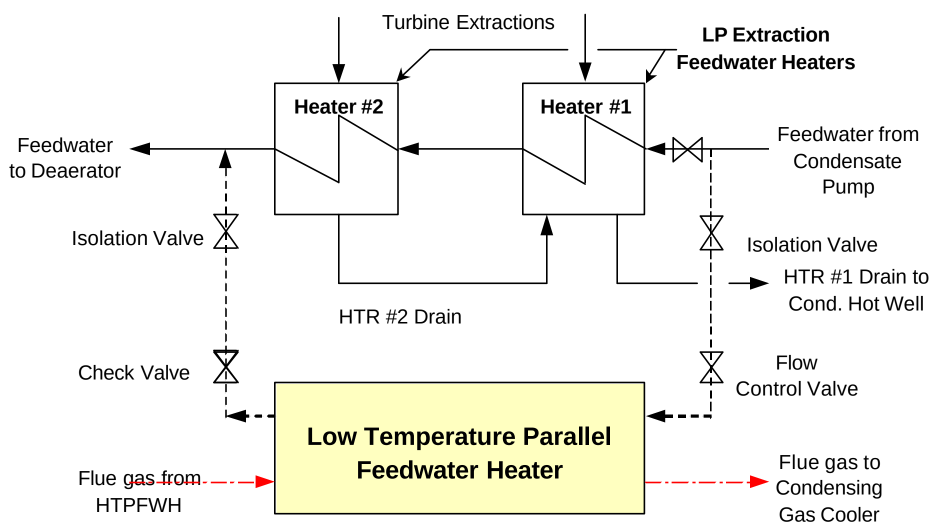


Figure 2.6. 5: Case-6 Low Temperature Parallel Feedwater Heater Arrangement

Gas Cooler:

The gas cooler of Case-6 is used to cool the flue gas leaving the low temperature PFWH to as low a temperature as possible in order to minimize the power requirements of the Boiler Draft System and the Gas Processing System. The Gas Cooler is a direct contact, water spray type of system. Some of the water vapor contained within the flue gas stream also condenses out in this cooler. This cooler is designed to cool the flue gas to 100^{o}F. This equipment is described further in Section 2.6.2 as it is considered a part of the Gas Processing System.

Draft System:

The flue gas is moved through the Boiler Island equipment with the boiler draft system. The boiler draft system includes the sweep gas fan, the fluidizing gas blower, the induced draft (ID) Fan, the associated ductwork, and expansion joints. The induced draft fan, sweep gas fan, and fluidizing gas blower are driven with electric motors and controlled to operate the unit in a balanced draft mode with the cyclone inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

2.6.2. Case-6 Gas Processing System Process Description and Equipment

The Case-6 Gas Processing System (GPS) is basically a scaled up version of the Case-4 GPS. The gas flow to the Case-6 GPS system is increased by about 22 percent as compared to Case-4. The gas analysis is nearly identical to that of Case-4 although not exactly the same since the OTM provides pure Oxygen to the Boiler Island system as compared to the 99 percent pure Oxygen provided by the Case-4 cryogenic ASU.

The purpose of this system is to processes the flue gas stream leaving the oxygen-fired Boiler Island to provide a liquid CO_2 product stream of suitable purity for an EOR application.

The Case-6 CO_2 capture system is designed for about 94 percent CO_2 capture. Cost and performance estimates were developed for all the systems and equipment required to cool, purify, compress and liquefy the CO_2, to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO_2 specification for EOR, given in Table 2.0.1 was used as the basis for the CO_2 capture system design.

A very low concentration of oxygen, in particular, is specified for meeting current pipeline operating practices, due to the corrosive nature of the oxygen. Hence, for Case-6, whereby the final CO_2 liquid product was found to contain about 11,600 ppmv of O_2, the design of the transport pipe to an EOR site for example would have to take this characteristic under consideration.

The nitrogen concentration specified in Table 2.0.1 is < 300 ppmv. It should be noted that according to Charles Fox of Kinder Morgan (Fox, 2002), this specification is very conservative as his company specifies a maximum nitrogen concentration of 4 percent (by volume) to control the minimum miscibility pressure. In Case 6 the nitrogen concentration in the liquid product was 9,800 ppmv. The exact reasoning behind the very low nitrogen specification listed in Table 2.0.1 is not clear.

2.6.2.1. Process Description

The following describes a CO_2 recovery system that cools and then compresses a CO_2 rich flue gas stream from an oxygen-fired CFB boiler to a pressure high enough so CO_2 can be liquefied. The resulting liquid CO_2 is passed through a CO_2 Stripper to reduce the N_2 and O_2 content to levels that are optimized with respect to energy consumption. Then the liquid CO_2 is pumped to a high pressure so it can be economically transported for sequestration or usage. Pressure in the transport pipeline will be maintained above the critical pressure of CO_2 to avoid 2-phase flow. The overhead gas from the CO_2 Stripper is vented to the atmosphere.

The key process parameters (pressures, temperatures, duties, etc.) are shown in the material and energy balance tables and will not be repeated here except in selected instances.

Figure 2.6.6 (Refer to Section 2.6.2.2) shows the Flue Gas Cooling process flow diagram and Figure 2.6.7 shows the Flue Gas Compression and Liquefaction process flow diagram.

Flue Gas Cooling:

Please refer to Figure 2.6.6 (drawing D 12173-06001-0).

The feed to the Gas Processing System is the flue gas stream that leaves the PFWH of the Boiler Island. At this point, the flue gas is near the dew point of H_2O. All of the flue gas leaving the boiler is cooled to 100 °F in Gas Cooler DA-101 that operates slightly below atmospheric pressure. A significant amount of water condenses out in this cooler. Excess condensate is blown down to the cooling water system. A single vessel has been provided for this cooler.

The Gas Cooler is configured in a packed tower arrangement where the flue gas is contacted with cold water in countercurrent fashion. Warm water from the bottom of the contactor is recycled back to the top of the contactor by Water Pump GA-101 after first cooling it in an external water cooled heat exchanger, Water Cooler EB-101 (plate and frame exchanger). The cooling water for this exchanger comes from the new cooling tower.

Because the flue gas may carry a small amount of fly ash, the circulating water is filtered in Water Filter FD-101A-F to prevent solids build-up in the circulating water. Condensate blowdown is filtered and is taken out downstream of the filter. However, the stream is not cooled and is split off before EB-101. Thus the heat load to the cooling tower is minimized.

From the Gas Cooler the gas stream then is boosted in pressure by the ID fan followed by a split of the gas into two streams. This design was developed to minimize the length of ducting operating at a slight vacuum and to minimize the temperature of the gas being recycled back to the boiler. The mass flow rate of the gas recirculation stream is about 52 percent of the flow rate of the product gas stream, which proceeds to the gas compression area. The recycle stream is sized to provide oxygen content of about 70 percent by volume in the oxidant stream supplying the boiler. The Gas Cooler minimizes the volumetric flow rate to, and the resulting power consumption of, the Flue Gas Compression equipment located downstream.

Three-Stage Gas Compression System:

Please refer to Figure 2.6.7 (Drawing D 12173-06002-0).

The compression section, where CO_2 is compressed to 365 psig by a three-stage centrifugal compressor, includes Flue Gas Compressor GB-101. After the aftercoolers, the stream is then chilled in a propane chiller to a temperature of –21 °F. Note that both the trim cooling water and water for the propane condenser comes from the cooling tower. At this pressure and temperature, about 80 mole percent of the stream can be condensed. The flash vapors contain approximately 80 weight percent of the inlet oxygen and nitrogen, but also about 13.7 weight percent of the CO_2. Therefore, a rectifier tower has been provided to reduce the loss of CO_2 to an acceptable level (about 6 weight percent). Then the pressure of the liquid is boosted to 2,000 psig by CO_2 Pipeline Pump GA-103. This stream is now available for sequestration or usage.

The volumetric flow to the compressor inlet is about 87,000 ACFM and only a single frame is required. The discharge pressures of the stages have been balanced to give reasonable power distribution and discharge temperatures across the various stages. They are:

- 1st Stage 28 psig
- 2nd Stage 108 psig
- 3rd Stage 365 psig

Power consumption for this large compressor has been estimated assuming adiabatic efficiency of 75 percent.

The hot gas from each stage is first cooled in an air cooler to 120 ^oF (Flue Gas Compressor 1st/2nd/3rd Stage Aftercooler EC-101/2/3) and then further cooled by a water-cooled heat exchanger to 95 ^oF (Flue Gas Compressor 1st/ 2nd Stage Trim Cooler EA-101/2). The flue gas compressor 3rd stage cooler (EA-103A/B) cools the gas to 90 ^oF to reduce the size of the dryers. Due to their large size, many of these heat exchangers consist of multiple shells. Because of highly corrosive conditions, the process side of the coolers must be stainless steel.

Because the flue gas stream leaving DA-101 is saturated, some water condenses out in the three aftercoolers. The sour condensate is separated in knockout drums (FA-100/1/2/3) equipped with mist eliminator pads. Condensate from these drums is drained to the cooling tower or to waste water treatment. To prevent corrosion, these drums have a stainless steel liner.

Gas Drying:

Please refer to Figure 2.6.7 (Drawing D 12173-06002-0).

It is necessary to dry the CO_2 stream to meet the product specification. Flue gas leaving the 3rd stage discharge knockout drum (FA-103) is fed to Flue Gas Drier FF-101 A/B where additional moisture is removed. An alumina bed drier has been selected.

The performance of a fixed-bed drier improves as pressure increases. This favors locating the drier at the discharge of the compressor. However, as the operating pressure of the drier increases, so does the design pressure of the equipment. This favors low-pressure operation. But, at low pressure the diameter or number of the drier vessels grows, increasing the cost of the vessel. Having to process the recycle gas from the rectifier condenser cooling would also increase the diameter of the vessel. However, this is less than 10 percent of the forward flow. For this design the drier has been optimally located downstream of the 3rd stage compressor. The CO_2 Drier system consists of two vessels; FF-101 A/B. One vessel is on line while the other is being regenerated. Flow direction is down during operation and up during regeneration.

The drier is regenerated with the noncondensable vent gas from the rectifier after it exits the third stage discharge trim cooler in a simple once through scheme. During regeneration, the gas is heated in Regeneration Heater FH-101 before passing it through the exhausted drier. After regeneration, heating is stopped while the gas flow continues. This cools the bed down to the normal operating range. The regeneration gas and the impurities contained in it are vented to the atmosphere.

Regeneration of an alumina bed requires relatively high temperature and, because HP steam pressure may fluctuate, a gas-fired heater has been specified for this service.

Flue Gas Filter FD-102 has been provided at the drier outlet to remove any fines that the gas stream may pick up from the desiccant bed.

CO_2 Condensation and Stripping:

Please refer to Figure 2.6.7 (Drawing D 12173-06002-0).

From the CO_2 Drier, the gas stream is cooled down further to -21 °F with propane refrigeration in CO_2 Condenser EA-104 A-F. From EA-104 the partially condensed flue gas stream continuesontoCO$_2$ Rectifier DA-102.

At this pressure and temperature, 80 mole percent of the stream can be condensed. The flash vapors contain approximately 80 weight percent of the inlet oxygen and nitrogen, but also 12 weight percent of the CO_2. Therefore, as mentioned, a rectifier tower has been provided to reduce the loss of CO_2 to an acceptable level. The pressure of the liquid is boosted to 2,000 psig by CO_2 Pipeline Pump GA-103 for delivery to a sequestration or usage location.

The vapors in the feed to the rectifier contain the nitrogen and the oxygen that flashed from the liquid CO_2. To keep the CO_2 loss to the minimum, the rectifier also has an overhead condenser, CO_2 Rectifier Condenser EA-107. This is a floodback type condenser installed on top of the Rectifier. It cools the overhead vapor from the tower down to –48 °F. The condensed CO_2 acts as cold reflux in the CO_2 Rectifier.

Taking a slipstream from the inert-free liquid CO_2 from the Rectifier bottoms and letting it down to the Flue Gas Compressor 3rd stage suction pressure cools EA-107. At this pressure, CO_2 liquid boils at –55 °F thus providing the refrigeration necessary to condense some of the CO_2 from the Stripper overhead gas. The process has been designed to achieve at least 94 percent CO_2 recovery. The vaporized CO_2 from the cold side of EA-107 is fed to the suction of the Flue Gas Compressor 3rd stage.

Any system containing liquefied gas such as CO_2 is potentially subject to very low temperatures if the system is depressurized to atmospheric pressure while the system contains cryogenic liquid. If the CO_2 Rectifier (and all other associated equipment that may contain liquid CO_2) were to be designed for such a contingency, it would have to be made of stainless steel. However, through proper operating procedures and instrumentation such a scenario can be avoided, and low temperature carbon steel (LTCS) can be used instead. Our choice here is LTCS. However, the condenser section will be made from stainless steel.

CO_2 Pumping and CO_2 Pipeline:

Please refer to Figure 2.6.7 (Drawing D 12173-06002-0).

The CO_2 product must be increased in pressure to 2,000 psig. A multistage heavy-duty pump (GA-103) is required for this service. This is a highly reliable derivative of an API-class boiler feed-water pump.

It is important that the pipeline pressure be always maintained above the critical pressure of CO_2 such that single-phase (dense-phase) flow is guaranteed. Therefore, pressure in the line should be controlled with a pressure controller and the associated control valve located at the destination end of the line.

Offgas:

Please refer to Figure 2.6.7 (Drawing D 12173-06002-0).

The vent gas from the CO_2 Rectifier overhead is at high pressure and there is an opportunity for power recovery using turbo-expanders. Because the gas cools down in the expansion process, there is also an opportunity for cold recovery. Heat recovery from the stream after let down via an expander was examined and it was determined that the amount of duty that could be recovered without the carbon dioxide in the stream freezing was small. Thus heat recovery could not be justified. The offgas leaves the Rectifier at –48 ^{o}F approximately. The refrigeration recovery to condense CO_2 was the best use for this cold since it also produces a reasonable temperature regeneration gas for the dryers.

2.6.2.2. Process Flow Diagrams

Two process flow diagrams are shown below for these systems:

- Figure 2.6.6 (drawing D 12173-06001-0) Flue Gas Cooling PFD
- Figure 2.6.7 (drawing D 12173-06002-0) CO_2 Compression and Liquefaction PFD

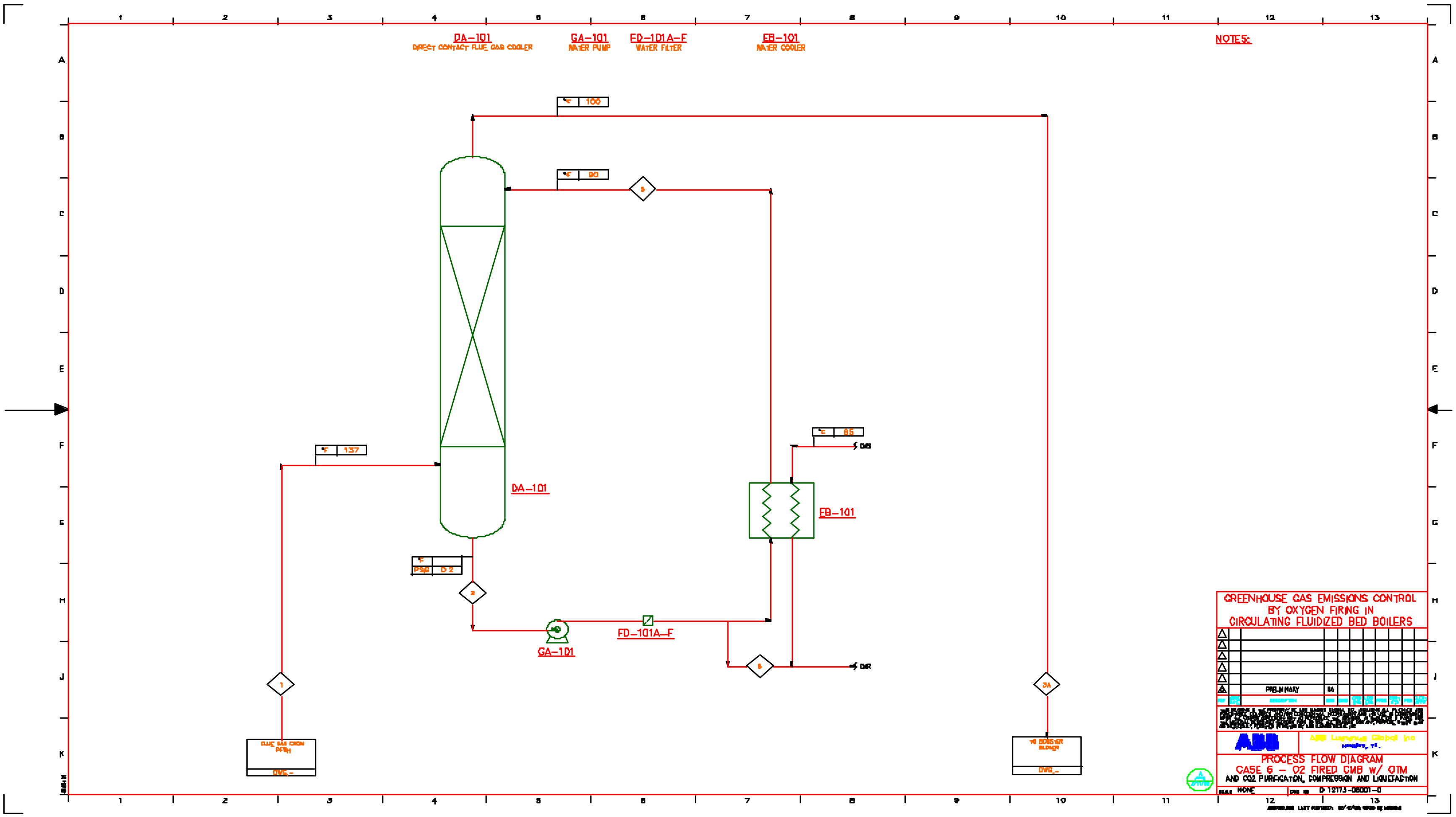


Figure 2.6. 6: Case-6 Flue Gas Cooling Process Flow Diagram

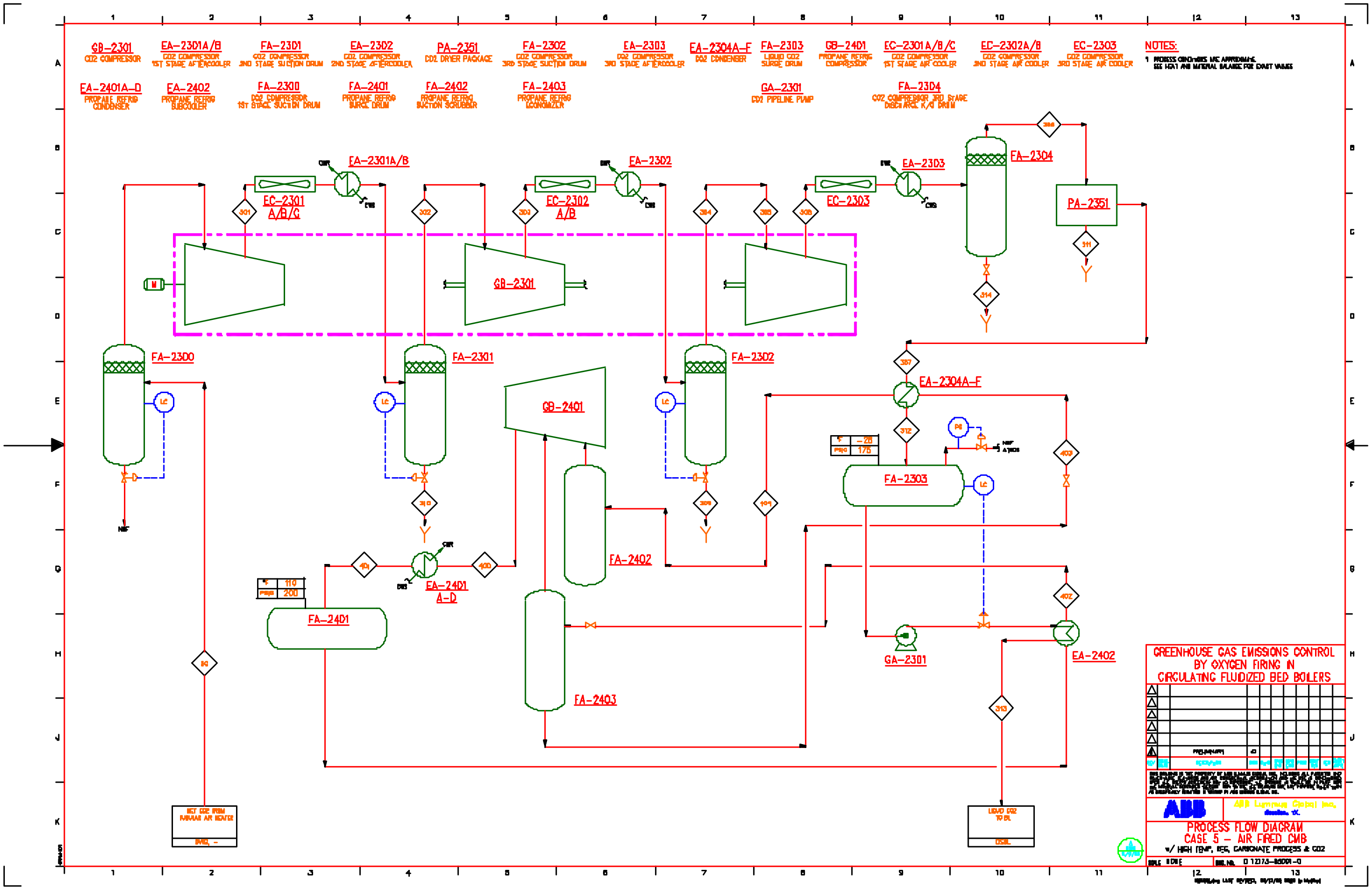


Figure 2.6. 7: Case-6 CO_2 Compression and Liquefaction Process Flow Diagram

2.6.2.3. Material and Energy Balance

Table 2.6.2 contains the overall material and energy balance for the Flue Gas Cooling System and the CO_2 Compression and Liquefaction System for Case-6 described above. It is based on 94 percent recovery of CO_2. Please refer to the Process Flow Diagrams shown in the previous section for stream numbers shown in this table.

Table 2.6. 2: Case-6 Gas Processing System Material & Energy Balance

STREAM NAME		To quench columns	From Quench columns	Excess water	From Large blowers	Quench water out	Quench water in	To liquefaction train	To boiler	To Train A liquefaction	First water KO	To 2nd stage	2nd water KO	To 3rd stage	Recycle from condenser	To drier
PFD STREAM NO.		**1**	**3a**	**6**	**3b**	**2**	**5**	**3c**	**3d**	**4**	**7**	**8**	**9**	**10**	**25**	**12**
VAPOR FRACTION	Molar	0.980	1.000	0.000		0.000	0.000	1.000			0.000	1.000	0.000	1.000	1.000	1.000
TEMPERATURE	°F	135.0	100	117		117	90	112			95	95	86	86	-48	90
PRESSURE	PSIA	13.7	14	55		14	45	15			38	38	116	116	117	373
MOLAR FLOW RATE	lbmol/hr	22,323	19,150.99	3,169.34		118,171.77	115,000.00	12,588.25			611.25	11,977.00	15.71	12,761.61	975.00	12,720.56
MASS FLOW RATE	lb/hr	841,920	784,747	57,118		2,129,697	2,072,524	515,820			11,028	504,792	284	543,995	42,649	543,245
ENERGY	Btu/hr	-3.19E+09	-2.86E+09	-3.87E+08		-1.44E+10	-1.41E+10	-1.88E+09			-7.48E+07	-1.82E+09	-1.93E+06	-1.96E+09	-1.63E+08	-1.96E+09
COMPOSITON	Mol %															
CO2		72.13%	84.07%	0.02%		0.02%	0.02%	84.08%			0.00%	88.36%	0.30%	90.39%	97.72%	90.68%
H2O		20.17%	6.96%	99.97%		99.97%	99.98%	6.96%			0.00%	2.22%	99.67%	0.59%	0.00%	0.28%
Nitrogen		3.97%	4.62%	0.00%		0.00%	0.00%	4.62%			0.00%	4.86%	0.00%	4.63%	0.98%	4.65%
Ammonia		0.00%	0.00%	0.00%		0.00%	0.00%	0.00%			0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%		0.00%	0.00%	0.00%			0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		3.63%	4.23%	0.00%		0.00%	0.00%	4.23%			0.00%	4.44%	0.00%	4.26%	1.16%	4.27%
SO2		0.10%	0.12%	0.00%		0.00%	0.00%	0.12%			0.00%	0.12%	0.02%	0.12%	0.14%	0.12%
VAPOR																
MOLAR FLOW RATE	lbmol/hr	21,884.7	19,151.0	-		-	-	12,588.2			-	11,977.0	-	12,761.6	975.0	12,720.6
MASS FLOW RATE	lb/hr	834,026	784,747	-		-	-	515,820			-	504,792	-	543,995	42,649	543,245
STD VOL. FLOW	MMSCFD	199.32	174.42	-		-	-	114.65			-	109.08	-	116.23	8.88	115.86
ACTUAL VOL. FLOW	ACFM	168,991.99	139,897.16	-		-	-	87,153.52			-	31,210.73	-	10,288.89	552.10	2,939.98
MOLECULAR WEIGHT	MW	38.11	40.98	-		-	-	40.98			-	42.15	-	42.63	43.74	42.71
DENSITY	lb/ft³	0.08	0.09	-		-	-	0.10			-	0.27	-	0.88	1.29	3.08
VISCOSITY	cP	0.0145	0.0149	-		-	-	0.0153			-	0.0154	-	0.0156	0.0114	0.0164
LIGHT LIQUID																
MOLAR FLOW RATE	lbmol/hr	-	-	-		-	-	-			-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-		-	-	-			-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-		-	-	-			-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-		-	-	-			-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-		-	-	-			-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-		-	-	-			-	-	-	-	-	-
VISCOSITY	cP	-	-	-		-	-	-			-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-		-	-	-			-	-	-	-	-	-
HEAVY LIQUID																
MOLAR FLOW RATE	lbmol/hr	438.05	-	3,169.34		118,171.77	115,000.00	-			611.25	-	15.71	-	-	-
MASS FLOW RATE	lb/hr	7,894	-	57,118.00		2,129,697	2,072,524	-			11,027.66	-	284.44	-	-	-
STD VOL. FLOW	BPD	542	-	3,919		146,135	142,212	-			757	-	20	-	-	-
ACTUAL VOL. FLOW	GPM	16.04	-	115.16		4,294.10	4,130.58	-			22.02	-	0.57	-	-	-
DENSITY	lb/ft³	61.35	-	61.84		61.83	62.56	-			62.44	-	62.74	-	-	-
VISCOSITY	cP	0.4835	-	0.5705		0.5708	0.7606	-			0.7185	-	0.8177	-	-	-
SURFACE TENSION	Dyne/Cm	66.44	-	68.20		68.21	70.83	-			70.30	-	70.97	-	-	-

STREAM NAME		3rd water ko	From drier / Condenser inlet	Condenser outlet	Non-condensable vent	Rectifier bottoms to condenser	CO_2 to pipeline	Refrig compressor discharge	Refrig condenser out	Refrig subcooler out	Refrig to CO_2 condenser	Refrig from CO_2 condenser	Warm non condensable
PFD STREAM NO.		11	14	15	24	22	21	100	101	102	103	104	26
VAPOR FRACTION	Molar	0.000	1.000	0.197	1.000	0.132	0.000	1.000	0.000	0.000	0.246	0.996	1.000
TEMPERATURE	°F	90	90	-22	-45	-55	82	92	110	48	-28	-28	56
PRESSURE	PSIA	373	360	355	352	120	2,015	100	215	212	21	21	347
MOLAR FLOW RATE	lbmol/hr	41.05	12,685.46	12,435.46	1,547.27	975.00	10,163.27	11,700.00	11,700.00	11,700.00	11,700.00	11,700.00	1,547.27
MASS FLOW RATE	lb/hr	750	542,613	531,920	55,405	42,649	444,563	515,935	515,935	515,935	515,935	515,935	55,405
ENERGY	Btu/hr	-5.04E+06	-1.96E+09	-1.99E+09	-1.12E+08	-1.68E+08	-1.73E+09	-5.23E+08	-5.92E+08	-6.14E+08	-6.14E+08	-5.44E+08	-1.11E+08
COMPOSITON	Mol %												
CO_2		0.89%	90.93%	90.93%	42.06%	97.72%	97.72%	0.00%	0.00%	0.00%	0.00%	0.00%	42.06%
H_2O		99.05%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	4.66%	4.66%	31.17%	0.98%	0.98%	0.00%	0.00%	0.00%	0.00%	0.00%	31.17%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%	0.00%
Oxygen		0.00%	4.28%	4.28%	26.77%	1.16%	1.16%	0.00%	0.00%	0.00%	0.00%	0.00%	26.77%
SO_2		0.06%	0.12%	0.12%	0.00%	0.14%	0.14%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR													
MOLAR FLOW RATE	lbmol/hr	-	12,685.5	2,446.5	1,547.3	128.5	-	11,700.0	-	-	2,876.5	11,654.9	1,547.3
MASS FLOW RATE	lb/hr	-	542,613	94,959	55,405	5,398	-	515,935	-	-	126,845	513,948	55,405
STD VOL. FLOW	MMSCFD	-	115.54	22.28	14.09	1.17	-	106.56	-	-	26.20	106.15	14.09
ACTUAL VOL. FLOW	ACFM	-	3,054.21	452.31	283.71	70.30	-	10,244.39	-	-	9,953.26	40,328.59	386.37
MOLECULAR WEIGHT	MW	-	42.77	38.81	35.81	42.00	-	44.10	-	-	44.10	44.10	35.81
DENSITY	lb/ft^3	-	2.96	3.50	3.25	1.28	-	0.84	-	-	0.21	0.21	2.39
VISCOSITY	cP	-	0.0164	0.0145	0.0146	0.0117	-	0.0087	-	-	0.0066	0.0066	0.0178
LIGHT LIQUID													
MOLAR FLOW RATE	lbmol/hr	-	-	9,988.94	-	846.49	10,163.27	-	11,700.00	11,700.00	8,823.51	45.05	-
MASS FLOW RATE	lb/hr	-	-	436,960.7	-	37,250.47	444,562.7	-	515,934.9	515,934.9	389,090.2	1,986.65	-
STD VOL. FLOW	BPD	-	-	36,140	-	3,086	36,771	-	69,724	69,724	52,582	268	-
ACTUAL VOL. FLOW	GPM	-	-	824.20	-	64.98	1,105.75	-	2,231.95	1,984.46	1,360.63	6.95	-
DENSITY	lb/ft^3	-	-	66.10	-	71.47	50.13	-	28.82	32.41	35.65	35.65	-
MOLECULAR WEIGHT	MW	-	-	43.74	-	44.01	43.74	-	44.10	44.10	44.10	44.10	-
VISCOSITY	cP	-	-	0.1624	-	0.2224	0.0568	-	0.0835	0.1172	0.1792	0.1792	-
SURFACE TENSION	Dyne/Cm	-	-	15.14	-	20.08	0.88	-	4.81	8.85	14.28	14.28	-
HEAVY LIQUID													
MOLAR FLOW RATE	lbmol/hr	41.05	-	-	-	0.00	-	-	-	-	-	0.00	-
MASS FLOW RATE	lb/hr	750.19	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	52	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	1.49	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft^3	62.85	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	0.7748	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	70.19	-	-	-	-	-	-	-	-	-	-	-

2.6.2.4. Gas Processing System Utilities

The following tables define the cooling water, natural gas, and electrical requirements for the Gas Processing System.

Table 2.6. 3: Case-6 Gas Processing System Cooling Water and Fuel Gas Requirements

GPS COOLING WATER - Case-6

REV	Equipment TAG NO	SERVICE	No. Installed	DUTY MMBTU/HR	INLET TEMP, F	OUTLET TEMP, F	FLOWRATE LB/HR
	EA-101	FG Comp 1 stg trim cooler	1	8.73	85	103	484,848
	EA-102	FG Comp 1 stg trim cooler	1	5.68	85	103	315,657
	EA-103	FG Comp 1 stg trim cooler	1	4.73	85	103	262,626
	EA-201	Refrig Condenser	1	80.55	85	100	5,369,697
	EB-101	Water Cooler	1	57.91	85	105	2,895,455
		TOTAL COOLING WATER		**157.59**			9,328,283

FUEL GAS FUEL GAS VALUE BASIS: 930 BTU/SCF (LHV)

REV	Equipment TAG NO	SERVICE	ONLINE FACTOR	COMPR HP	HEAT RATE BTU/HP-HR	DUTY MMBTU/HR	EFFICIENCY %	FLOWRATE (Peak) MMSCFD	FLOWRATE (Peak) SCFH	FLOW (Avg) MMSCFD
	FH-101	Mole sieve regeneration	61%			5.72	80%	0.185	7,688	0.113
		TOTAL FUEL GAS				5.72		0.185	7,688	0.113

Table 2.6. 4: Case-6 Gas Processing System Electrical Requirements

Number of Trains	Item Number	Service	Power (ea) including 0.95 motor eff	Total all trains
			(kW)	(kW)
1	EC-101	Flue Gas Compressor 1st Stage Aftercooler	90	90
1	EC-102	Flue Gas Compressor 2nd Stage Aftercooler	73	73
1	EC-103	Flue Gas Compressor 3rd Stage Aftercooler	75	75
1	GB-101	1st Stage	7217	7217
1		2nd Stage	7253	7253
1		3rd Stage	7393	7393
1	GB-102	1st Stage	6523	6523
1		2nd Stage	3777	3777
1	GA-101	Water pump	113	113
1	GA-103	CO2 Pipeline pump	920	920
		Total		**33434**

2.6.2.5. Gas Processing System Equipment

The equipment list for the Gas Processing System is provided in Appendix I, Section 9.1.6.2.

2.6.3. Case-6 Oxygen Transport Membrane System Process Description and Equipment

2.6.3.1. Process Description

A team led by Praxair has been developing Advanced Oxygen Transport Membranes (OTM), which can be integrated with power generation systems such as IGCC and oxygen fired combustion based systems to produce significantly lower-cost oxygen (Prasad, et al. 2002). This work is being carried out in partnership with the Department of Energy's National Energy Technology Center under Contract No. DE-FC26-99FT40437.

Prasad et al., describe the OTM technology as follows: "OTM technology is based on ceramic materials, which can rapidly transport oxygen ions at 600 – 1,000°C. Mixed conductors, which transport both oxygen ions and electrons, can be operated in a pressure driven mode, obviating the need for the costly electrodes and external circuits that are required for purely ionic conductors. Mixed conductors can be a single-phase material, which conducts both electrons and oxygen ions, or "dual Phase" materials (Figure 2.6.8) wherein two separate phases are used for transporting oxygen ions and electrons.

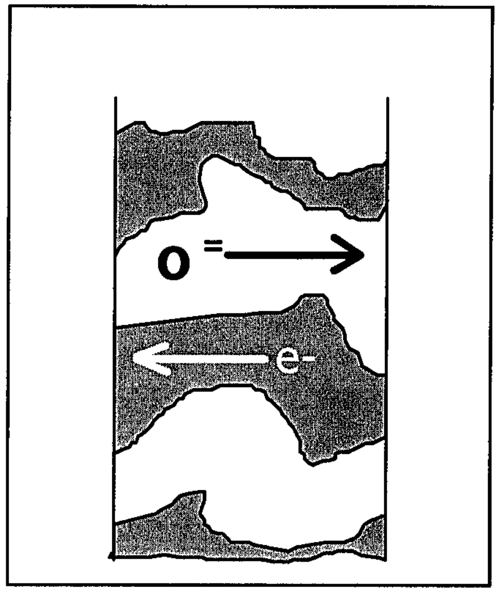


Figure 2.6.8 Schematic of the Dual Phase Oxygen Transport Membrane (from Prasad, et al., 2002)

The oxygen chemical potential difference across the membrane provides the driving force for oxygen transport. Oxygen atoms adsorb on the cathode (high oxygen partial pressure side of the membrane), dissociate into atoms/ions as they pick electrons. These ions travel from cathode to anode (the low oxygen partial pressure) by jumping through lattice sites and vacancies until they reach the anode side of the membrane. On the anode side, the oxygen ions give up their electrons to become atoms/molecules, which are then desorbed into the gas phase. Electrons from the anode side are carried through the membrane to the cathode side to complete the circuit. The rate of oxygen transport through such membranes is temperature sensitive, and can be very fast at high temperatures."

Prasad et al. Continue: "Ideally, the flux through the membrane is inversely proportional to the thickness, hence thin films can enable higher fluxes, leading to compact systems. These membranes also have infinite selectivity for oxygen over other gases, because only oxygen ions can occupy the lattice positions. The ability to produce pure oxygen at high permeation rates combined with the thermal integration enabled by high temperature operation results in significant benefits upon integration with the IGCC." This process uses a purge gas on the permeate side of the OTM to increase the chemical potential, increase the flux, and improve the overall process performance.

Prasad, et al. presented a brief status of the OTM development at the 2002 Pittsburgh Conference on Coal Science.

2.6.4. Case-6 Balance of Plant Equipment and Performance

The balance of plant equipment described in this section includes the steam cycle performance and equipment, the draft system equipment, the cooling system equipment, and the material handling equipment (coal, limestone, and ash). Refer to Appendix I for equipment lists and Appendix II for drawings.

2.6.4.1. Steam Cycle Performance

The steam cycle for Case-6 is shown schematically in Figure 2.6.9. The Mollier diagram which illustrates the process on enthalpy - entropy coordinates is the same as for Case-1 and is not repeated here. The steam cycle arrangement and performance for Case-6 is somewhat different than Case-4. There is significantly more low-level heat recovery for this case due to the arrangement of the OTM system.

The steam cycle starts at the condenser hot well, which is a receptacle for the condensed steam from the exhaust of the steam turbine. The condensate flows to the suction of the condensate pumps, which increase the pressure of the fluid by a nominal 250 psi to transport it through the piping system and enable it to enter the open contact heater, or deaerator. The condensate passes through a gland steam condenser (SPE) first, followed in series by two low-pressure extraction feedwater heaters. The heaters successively increase the condensate temperature to a nominal 221°F by condensing and partially sub-cooling steam extracted from the LP steam turbine section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (now referred to as heater drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the condenser. The Case-6 condensate and feedwater system includes a parallel low-pressure Feedwater Heater (LT PFWH - heated by flue gas) in a parallel feedwater stream with the two low-pressure extraction feedwater heaters as shown in Figure 2.6.9.

The condensate entering the deaerator is heated and stripped of noncondensable gases by contact with the steam entering the unit. The steam is condensed and, along with the heated condensate, flows by gravity to a deaerator storage tank. The boiler feedwater pumps take suction from the storage tank and increase the fluid pressure to a nominal 2200 psig. Both the condensate pump and boiler feed pump are electric motor driven. The boosted condensate flows through three more high-pressure feedwater heaters, increasing in temperature to 470°F at the entrance to the boiler economizer section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the deaerator. The high pressure feedwater system differs from Case 1 in that there is a parallel high-pressure Feedwater Heater (HT PFWH - heated by flue gas) in a parallel feedwater stream with the three high-pressure extraction feedwater heaters as shown in Figure 2.6.9.

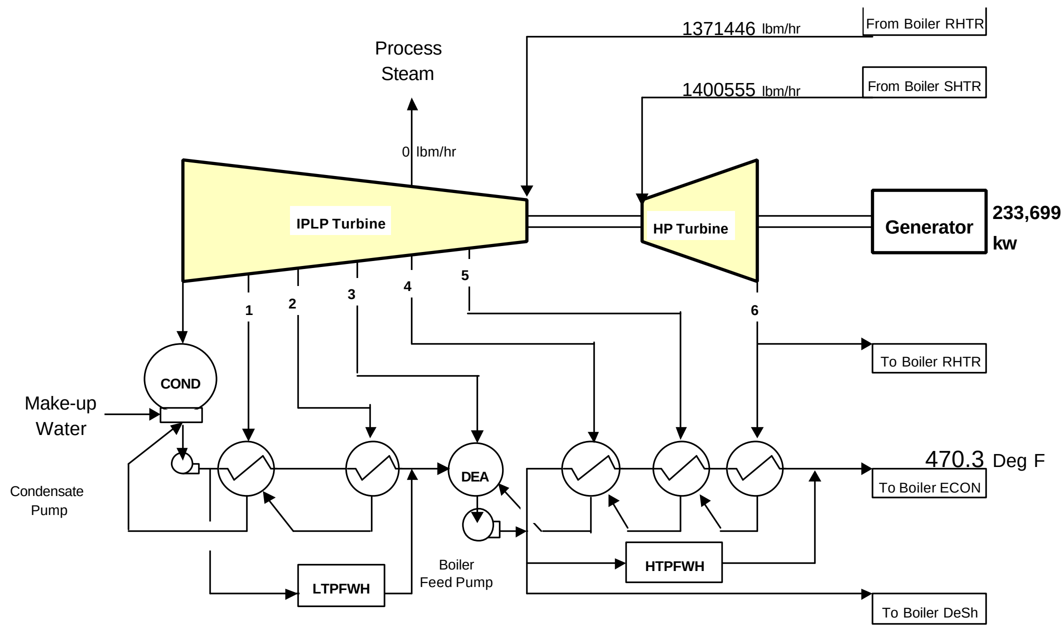
The high-pressure turbine expands 1,400,555 lbm/hr of steam at 1,800 psia and 1,005 ^{o}F, the same as for all cases. The reheat steam (1,371,446 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,005 ^{o}F. The reheat flow is higher due to heat recovery in parallel with the high pressure extraction feedwater heaters (HTPFWH).

The HTPFWH shown in Figure 2.6.9 is in reality two heat exchangers with parallel water streams. One recovers heat from the flue gas stream leaving the baghouse of the Boiler Island. The other recovers heat from the depleted air stream leaving the gas expander of the OTM system.

Similarly, The LTPFWH shown in Figure 2.6.9 is in reality two heat exchangers with parallel water streams. One recovers heat from the flue gas stream leaving the HTPFWH of the Boiler Island. The other recovers heat from the depleted air stream leaving the HTPFWH of the OTM system.

The condenser pressure used for Case-6 and all other cases in this study was 3.0 in. Hga. The steam turbine performance analysis results show the generator produces 233,669 kW output and the steam turbine heat rate is about 8,758 Btu/kWh.

The generator output, turbine heat rate and condenser losses are significantly higher for Case-6 than for the other cases. This is a direct result of the increased low level heat recovery in parallel with the extraction feedwater heaters which reduces extraction flows to the low-pressure and high pressure extraction feedwater heaters and increases the IP & LP turbine power output and condenser loss.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Steam Turbine Power Output	811	Boiler Heat Input	2014
Process Steam Heat Loss	0	BFP & CP Input	12
Condenser Loss	1215	**Total Energy Input**	2026
Total Energy Output	2026	**In - Out**	0

Turbine Heat Rate 8758 (Btu/kwhr)

Figure 2.6. 9: Case-6 Steam Cycle Schematic and Performance

2.6.4.2. Steam Cycle Equipment

This section provides a brief description of the major equipment (steam turbine, condensate and feedwater systems) utilized for the steam cycle of this case.

Steam Turbine:

The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the boiler passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheated steam flows through the reheat stop valves and

intercept valves and enters the IP section at 465 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser.
The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for economic unit cycling.

Condensate and Feedwater Systems:

The function of the condensate system is to pump condensate from the condenser hot well to the deaerator, through the gland steam condenser and the LP feedwater heaters. The Case-6 condensate and feedwater system includes parallel low-pressure Feedwater Heaters (PFWH - heated by flue gas) in parallel feedwater streams with the traditional extraction feedwater heaters. The PFWH's are part of the Boiler scope of supply.

The condensate system consists of one main condenser; two 50 percent capacity, motor-driven vertical condensate pumps; one gland steam condenser; two LP heaters, and one deaerator with a storage tank. Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump feedwater from the deaerator storage tank to the boiler economizer. Two motor-driven boiler feed pumps are provided to pump feedwater through the three stages of HP feedwater heaters. Pneumatic flow control valves control the recirculation flow. In addition, the suctions of the boiler feed pumps are equipped with startup strainers, which are utilized during initial startup and following major outages or system maintenance.

OTM Heat Recovery System:

The OTM heat recovery system is utilized to recover the remaining sensible heat from the expanded off gas from the OTM gas expander. Two heat exchangers (HT-PFWH and LT-PFWH) are specified in the following table:

Table 2.6.5 OTM Heat Recovery Heat Exchanger Specification

Item	(units)	HTPFWH	LTPFWH
T gas in	(Deg F)	710	306.7
T gas out	(Deg F)	307	135
T water in	(Deg F)	287	115
T water out	(Deg F)	470	221
P water	(psia)	2269	64
Water flow	(lbm/hr)	869082	673854
Gas flow	(lbm/hr)	1651050	1651050
Heat Duty	(10^6 Btu/hr)	168.2	71.6

Both heat exchangers are located in the depleted air stream (between Streams 29 and 30 of Figure 2.6.1) leaving the OTM system Gas Expander. The depleted air stream analysis is shown in the following table.

Table 2.6.6 OTM Heat Recovery System Gas Analysis

Gas Analysis (wt frac.)	
O2	0.0426
N2	0.9415
H2O	0.0159
Total	1.0000

2.6.4.3. Other Balance of Plant Equipment

The systems for draft, solids handling (coal, limestone, and ash), cooling, electrical, and other BOP systems are described in this section for Case-6.

Draft System:

The draft system includes the draft equipment for the boiler and the draft equipment for the OTM.

Boiler Draft System:
The flue gas is moved through the boiler, baghouse and other Boiler Island equipment with the draft system. The draft system includes the Sweep Gas (SG) fans, the fluidizing gas blowers, the induced draft (ID) Fan, and the associated ductwork and expansion joints. This case has no traditional stack as the flue gas generated in the boiler is supplied to the gas processing system where the CO_2 is purified and liquefied for sequestration or usage. The fans, and blowers are driven with electric motors and controlled to operate the unit in a balanced draft mode with the cyclone inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

Recirculated flue gas from the SG fan is mixed with oxygen from the OTM to provide a combustion oxidant stream, which is split into several flow paths.

Combustion gases exit the furnace and flow through a single inverted cyclone, which separates out ash and partially burned fuel particles. These solids are recycled back to the furnace, passing through J-valves, or seal pots, located below the cyclone. The solids leaving the seal pot are then returned directly to the combustor.

The gas exiting the cyclone passes directly to the low temperature sweep gas heater (there is no convection pass for Case-6) and then exit the CMB steam generator to the baghouse for particulate capture. The flue gas leaving the baghouse is further cooled in a HT-PFWH and a LT-PFWH, which are low temperature economizer sections in parallel water streams with extraction feedwater heaters, and finally in a spray water cooler to about 100 ^{o}F. The gases are drawn through the CMB, baghouse, PFWH, and spray cooler with the Induced Draft Fan and then are recirculated to the CFB or discharged to the Gas Processing System.

The following fans and blowers are provided with the scope of supply of the Oxygen-fired CMB steam generator:

- Sweep Gas fan, which provides recirculated flue gas to be mixed with oxygen supplied from the OTM such that the mixed oxidant stream contains about 70 percent oxygen. This fan is a centrifugal type unit, supplied with an electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.6.7).

Table 2.6. 7: Sweep Gas Fan Specification

Gas Analysis			
Oxygen	(wt percent	3.30	
Nitrogen	"	3.16	
Water Vapor	"	3.06	
Carbon Dioxide	"	90.29	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	231646	277975
Gas Inlet Temperature	(Deg F)	112.1	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	19.89	
Pressure Rise	(in wg)	143.8	187.0
Drive Motor Power	(kW)	795	

- Induced draft fan, a centrifugal unit supplied with an electric motor drive and inlet damper (see Table 2.6.8).

Table 2.6. 8: Induced Draft Fan Specification

Gas Analysis			
Oxygen	(wt percent	3.30	
Nitrogen	"	3.16	
Water Vapor	"	3.06	
Carbon Dioxide	"	90.29	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	784820	941784
Gas Inlet Temperature	(Deg F)	100.0	
Inlet Pressure	(psia)	13.64	
Outlet Pressure	(psia)	14.70	
Pressure Rise	(in wg)	29.5	38.4
Drive Motor Power	(kW)	626	

- Fluidizing gas blowers, centrifugal units that provide recirculated flue gas for cooling and sealing the seal pots, and for assisting in the conveyance of cyclone bottoms (see Table 2.6.9).

Table 2.6. 9 Fluidizing Gas Blower Specification

Gas Analysis			
Oxygen	(wt percent	3.30	
Nitrogen	"	3.16	
Water Vapor	"	3.06	
Carbon Dioxide	"	90.29	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	37354	44825
Gas Inlet Temperature	(Deg F)	112.1	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	23.70	
Pressure Rise	(psia)	9.0	11.7
Drive Motor Power	(kW)	209	

- Transport air blowers, centrifugal units that provide air for pneumatic transport of cool Bauxite from the MBHE bottom to the top of the combustor (see Table 2.6.10).

Table 2.6. 10 Transport Air Blower Specification

Gas Analysis			
Oxygen	(wt percent	2.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	476070	571284
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	20.53	
Pressure Rise	(in wg)	161.5	210.0
Drive Motor Power	(kW)	2409	

OTM Draft System:
Additionally, this case includes an air compressor and gas expander for support of the OTM air supply and energy recovery requirements.

Compressed air (about 215 psia) is provided to the OTM from an air compressor. The compressed air is then preheated in the MBHE and supplied to the OTM. Most of the oxygen contained in the compressed air stream (~85 percent) is transported across the membrane in the OTM to provide the oxidant for the boiler. The depleted oxygen stream leaving the OTM, still at high pressure (about 200 psia) is expanded to near atmospheric pressure in a gas expander, thus generating power to offset the air compressor power requirements.

- Air Compressor, centrifugal unit that provides air for the OTM (see Table 2.6.11).

Table 2.6. 11 OTM Air Compressor Specification

Gas Analysis		
Oxygen	(wt percent	22.89
Nitrogen	"	75.83
Water Vapor	"	1.28
Carbon Dioxide	"	0.00
Sulfur Dioxide	"	0.00
Total	"	100.00
Operating Conditions		
Mass Flow Rate	(lbm/hr)	2049886
Gas Inlet Temperature	(Deg F)	80.0
Inlet Pressure	(psia)	14.70
Outlet Pressure	(psia)	215.05
Pressure Ratio	(Pout / Pin)	14.6
Drive Motor Power	(kW)	110920

- Gas Expander, an axial unit that expands hot gas leaving the OTM (see Table 2.6.12).

Table 2.6. 12 OTM Gas Expander Specification

Gas Analysis		
Oxygen	(wt percent	4.26
Nitrogen	"	94.15
Water Vapor	"	1.59
Carbon Dioxide	"	0.00
Sulfur Dioxide	"	0.00
Total	"	100.00
Operating Conditions		
Mass Flow Rate	(lbm/hr)	1651050
Gas Inlet Temperature	(Deg F)	1652.0
Inlet Pressure	(psia)	192.00
Outlet Pressure	(psia)	15.31
Pressure Ratio	(Pout / Pin)	0.080
Generator Output Powe	(kW)	122659

Ducting and Stack:

There is no traditional stack included in Case-6 Boiler Island as is true for Cases 2, 3 and 4 also. The entire flue gas product leaving the Boiler Island, which is rich in CO_2, is delivered to the Gas Processing System (GPS) where the CO_2 stream is further purified to be suitable for sequestration or usage. The impurities removed in the GPS, primarily nitrogen and oxygen are vented to atmosphere.

There is a small stub stack, which vents the clean oxygen depleted air stream leaving the OTM system gas expander and heat recovery system.

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1/4" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the three silos.

Technical Requirements and Design Basis:

- Coal burn rate:
  - Maximum coal burn rate = 202,446 lbm/h = 101 tph plus 10 percent margin = 111 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
  - Average coal burn rate = 172,000 lbm/h = 86 tph (based on MCR rate multiplied by an 85 percent capacity factor)
  - Coal delivered to the plant by unit trains:
  - One and one-half unit trains per week at maximum burn rate
  - One unit train per week at average burn rate
  - Each unit train shall have 10,000 tons (100-ton cars) capacity
  - Unloading rate = 9 cars/hour (maximum)
  - Total unloading time per unit train = 11 hours (minimum)
  - Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
  - Reclaim rate = 300 tph
  - Storage piles with liners, run-off collection, and treatment systems:
  - Active storage = 8,200 tons (72 hours at maximum burn rate)
  - Dead storage = 80,000 tons (30 days at average burn rate)

Table 2.6. 13: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	101
Active Storage, tons	8,200
Dead Storage, tons	60,000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,000-ton silo to accommodate 3 days operation.

Bottom Ash Removal:

Bottom ash, or bed drain material, constitutes approximately two-thirds of the solid waste material discharged by the CFB steam generator. This bottom ash is discharged through a complement of two bed coolers (any one of which must be able to operate at 100 percent load on the design coal). The stripper/coolers cool the bed material to a temperature in the range of 300 ^{o}F (design coal) to a maximum of 500 ^{o}F (worst fuel) prior to discharge via rotary valves to the bed material conveying system. The steam generator scope terminates at the outlets of the rotary valves.

Fly Ash Removal:

Fly ash comprises approximately one-third of the solid waste discharged from the CMB steam generator. Approximately 8 percent of the total solids (fly ash plus bed material) is separated out in the sweep gas heater hoppers; 25 percent of the total solids is carried in the gases leaving the steam generator en route to the baghouse. Fly ash is removed from the stack gas through a baghouse filter. Particulate conditions are as follows:

Design Specification for Particulate Removal System:

- Total solids to particulate removal system (stream 6, Figure 2.6.1) = 14,788 lbm/h
- Particle size distribution of particulate matter leaving cyclone (streams 3, 5, 6, Figure 2.6.1), see Table 2.6.14.

Table 2.6. 14: Particle Size Distribution

% Wt. Less	**Diameter** (Micron, μ)
100	192
99	160
90	74
80	50
70	37
60	30
50	24
40	16
30	12
20	8
10	4
1	< 4

- Solids leaving particulate removal system (stream 7, Figure 2.6.1) meet applicable environmental regulations, see Table 2.6.15.

Table 2.6. 15: Fly Ash Removal Design Summary

Design Parameter	Value
Flue Gas Temperature, ^oF	552
Flue Gas Flow Rate, lbm/h	841,921
Flue Gas Flow Rate, acfm	235,492
Particulate Removal, lbm/h	14,788
Particulate Loading, grains/acf	7.326

Ash Handling:

The function of the ash handling system is to provide the equipment required for conveying, preparing, storing, and disposing the bottom ash and fly ash that is produced on a daily basis by the boiler. The scope of the system is from the bag filter hoppers, sweep gas heater hopper collectors, and bottom ash hoppers to the truck filling stations.

The fly ash collected in the bag filter and the sweep gas heater is conveyed to the fly ash storage silo. A pneumatic transport system using low-pressure air from a blower provides the transport mechanism for the fly ash. Fly ash is discharged through a wet unloader, which conditions the fly ash and conveys it through a telescopic unloading chute into a truck for disposal.

The bottom ash from the boiler is drained from the bed, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. Ash from the fluidized bed ash coolers is drained to a complement of screw coolers, which discharge the cooled ash to a drag chain conveyor for transport to a surge bin. The latter is within the boiler scope of supply.

The cooled ash is pneumatically conveyed to the bottom ash silo from the surge bin. The silo is sized for a nominal holdup capacity of 36 hours of full-load operation (1,200 tons capacity). At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately 30 truck loads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.6.16: Ash Handling System Design Summary

Design Parameter	Value
Flyash from Baghouse, lbm/h	14,788
Ash from Boiler, lbm/h	80,988
Ash temperature, ^oF	520

Circulating Water System:

The function of the circulating water system is to supply cooling water to condense the main turbine exhaust steam. The system consists of two 50 percent capacity vertical circulating water pumps, a multi-cell mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Condenser Analysis:

The condenser system analysis is detailed in Table 2.6.17.

Table 2.6.17: Condenser Analysis

Item	Value	Units
Pressure	3.0	in. Hga
M stm-in	1,253,838	lbm/h
T stm-in	115.1	°F
P stm-in	1.474	psia
H stm-in	1051.7	Btu/lbm
M drain-in	25,650	lbm/h
H drain-in	89.7	Btu/lbm
H condensate	83	Btu/lbm
M condensate	1,279,488	lbm/h
Q condenser	1215.5	10^6 Btu/h

Waste Treatment System:

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment is housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge de-watering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water are provided. A 200,000-gallon storage tank provides a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide distributed control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes

from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Plant Layout and Plot Plan:

The Case-6 plant is arranged functionally to address the flow of material and utilities through the plant site. A plan view of the boiler, power-generating components, and overall site plan for the entire plant is shown in Appendix II.

2.6.5. Case-6 Overall Plant Performance and CO_2 Emissions

The overall plant performance and emissions for Case-6 are summarized in Table 2.6.17. The values of Case-1 (Base Case) and Case-4 are also listed along side for comparison purposes. The Base Case is shown because it is the primary comparison case for all the CO_2 removal cases. Case-4 is shown because Case-6 and Case-4 differ only in the design and performance of the Oxygen supply system (OTM system for Case-6 and cryogenic ASU for Case-4).

The coal heat input for Case-6 is significantly higher than all the other cases. It is about 21 percent greater than for Case-1 and about 23 percent greater than for Case-4. This large increase for Case-6 was primarily due to the requirement for the CMB to provide the additional heat duty of the high temperature air heater as required by the OTM system. In all the other cases the boiler heat output was only that amount required to produce the MCR steam flow and conditions.

Boiler efficiency for Case-6 is calculated to be 94.04 percent (HHV basis) as compared to 89.46 percent for the Base Case. The improvement is primarily due to the reduction in dry gas loss resulting from the oxygen firing. Refer to Section 2.2.5 for a discussion of why the dry gas loss is reduced with oxygen firing. The boiler efficiency for Case-4 is 93.66 percent by comparison.

The Case-6 steam cycle thermal efficiency including the boiler feed pump debit is about 38.97 percent. This is lower than for Case-4 (41.25 percent) and compares to 41.89 percent for Case-1. The decrease as compared to Case-4 is due to additional low-level heat recovery for Cases 6 as required by the OTM system.

Auxiliary power for Case-6 is 48,004 kW (not including the OTM air compressor) or about 19.5 percent of generator output. The large auxiliary power decrease for Case-6 as compared to Case-4 is due primarily to the large power requirement of the cryogenic based ASU used in Case-4 as compared to the power producing OTM System of Case-6. The power requirement for the Gas Processing System of Case-6 is about 120 percent of that for Case-4. The total plant auxiliary power for Case-6 is about 122 percent of the Case-4 requirement, exclusive of the ASU and GPS requirements. This increase is primarily due to the increased gas flow for Case-6.

The resulting net plant output for Case-6 is about 102 percent of the Base Case output and about 150 percent of the Case-4 output.

The net plant heat rate and thermal efficiency for Case-6 are calculated to be 11,380 Btu/kWh and 29.99 percent, respectively (HHV basis). This thermal efficiency is about 85 percent of the Base Case efficiency and about 22 percent better than Case-4.

Carbon dioxide emissions for Case-6 are 29,217 lbm/hr or about 0.15 lbm/kWh on a normalized basis. This represents about 7 percent of the Case-1 normalized CO_2 emissions and a CO_2 avoided value of 1.85 lbm/kWh. As compared to Case-4 this represents about 76 percent of the Case-4 normalized CO_2 emissions.

Table 2.6. 18: Case-6 Overall Plant Performance and Emissions

		CFB Air Fired (Case 1)	CMB Cryogenic O_2 Fired (Case 4)	CMB with OTM O_2 Fired (Case 6)
Auxiliary Power Listing	**(Units)**			
Induced Draft Fan	(kW)	2285	515	626
Primary Air Fan	(kW)	2427	n/a	n/a
Secondary Air Fan	(kW)	1142	n/a	n/a
Fluidizing Air Blower	(kW)	920	209	209
Transport Air Fan	(kW)	n/a	1865	2409
Gas Recirculation Fan	(kW)	n/a	344	795
Coal Handling, Preperation, and Feed	(kW)	300	294	363
Limestone Handling and Feed	(kW)	200	196	242
Limestone Blower	(kW)	150	147	181
Ash Handling	(kW)	200	196	242
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	152	186
Boiler Feed Pump	(kW)	3715	3715	3715
Condensate Pump	(kW)	79	79	92
Circulating Water Pump	(kW)	1400	1889	2006
Cooling Tower Fans	(kW)	1400	1889	2006
Steam Turbine Auxilliaries	(kW)	200	207	253
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719	719
Transformer Loss	(kW)	470	472	525
Subtotal	(kW)	16007	12888	14570
	(frac. of Gen. Output)	0.077	0.061	0.062
Traditional Power Plant Auxiliary Power	(kW)	16007	12888	14570
Air Separation Unit or Fuel Compressor	(kW)	n/a	37800	n/a
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a	110920
CO_2 Removal System Auxiliary Power	(kW)	n/a	27200	33434
Total Auxilary Power	(kW)	16007	77888	158923
	(frac. of Gen. Output)	0.077	0.371	0.196
Output and Efficiency				
Main Steam Flow	(lbm/hr)	1400555	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8275	8758
OTM System Expander Generator Output	(kW)	n/a	n/a	122659
Gas Turbine Generator Output		n/a	n/a	n/a
Steam Turbine Generator Output	(kW)	209041	210056	233699
Net Plant Output	(kW)	193034	132168	197435
	(frac. of Case-1 Net Output)	1.00	0.68	1.02
Boiler Efficiency (HHV)[1]	(fraction)	0.8946	0.9366	0.9404
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855	1820	2242
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	16.6	4.8
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855	1836	2247

[1] Boiler Heat Output / (Qcoal-HHV + Qcredits)
[2] Required for GPS Desicant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4

		CFB Air Fired (Case 1)	CMB Cryogenic O_2 Fired (Case 4)	CMB with OTM O_2 Fired (Case 6)
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	13894	11380
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.2456	0.2999
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.69	0.84
CO_2 Emissions				
CO_2 Produced	(lbm/hr)	385427	379959	466301
CO_2 Captured	(lbm/hr)	0	352380	437084
Fraction of CO2 Captured	(fraction)	0.00	0.93	0.94
CO_2 Emitted	(lbm/hr)	385427	27579	29217
Specific CO_2 Emissions	(lbm/kwhr)	2.00	0.21	0.15
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.10	0.07
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.79	1.85

2.7. Case-7: Indirect Combustion of Coal via Chemical Looping and CO_2 Capture

This section describes an advanced coal-fired power plant utilizing an atmospheric pressure Chemical Looping steam generator. The Chemical Looping steam generator is designed to produce a high CO_2 content stream leaving the fluidized bed while providing steam for a subcritical steam plant. This is accomplished through the use of a solid oxygen carrier to provide oxygen for combustion of the fuel. The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

The basic CO_2 capture concept behind Case-7 is to indirectly replace combustion air with oxygen thereby creating a high CO_2 content flue gas stream that can be further processed into a high purity CO_2 end product for various uses or sequestration. A chemical looping concept is used to indirectly provide the oxygen for the combustion of coal rather than direct utilization of ambient air as was done in Case-1. The chemical looping concept utilizes a solid oxygen carrier to supply the oxygen to the combustion process without the large efficiency penalty associated with the cryogenic type Air Separation Units (Cases 2, 3, 4). Additionally the large investment costs associated with both cryogenic type Air Separation Units and Oxygen Transport Membrane type oxygen supply systems are avoided. The trade off of course is the more complex boiler process, which is explained below.

A brief performance summary for this plant reveals the following information. The Case-7 plant produces a net plant output of about 164 MW. The net plant heat rate and thermal efficiency are calculated to be 11,051 Btu/kWh and 30.9 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 0.01 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.7.4.

2.7.1. Case-7 Boiler Island Process Description and Equipment

This section describes the Boiler Island processes for Case-7 and includes a simplified process flow diagram (PFD), material and energy balance and equipment description. The equipment description includes only the major components included in the Boiler Island.

It should be emphasized that the chemical looping combustion process described for this case is only conceptual at this time. A significant effort encompasing experimental work related to reaction rates, solids regeneration cycles, and fine particulate removal would be necessary to continue development of this combustion process. Additionally, high temperature air heater development would also be required.

2.7.1.1. Process Description and Process Flow Diagrams

Figure 2.7.1 shows a simplified process flow diagram for the Case-7 Boiler Island which utilizes indirect combustion of coal via chemical and thermal looping to provide a high CO_2 content stream for processing and capture. This process description briefly describes the function of the major equipment and systems included within the Boiler Island. Selected mass flow rates (lbm/hr) and temperatures (^oF) are shown on this figure. Complete data for all state points are shown in Table 2.7.1.

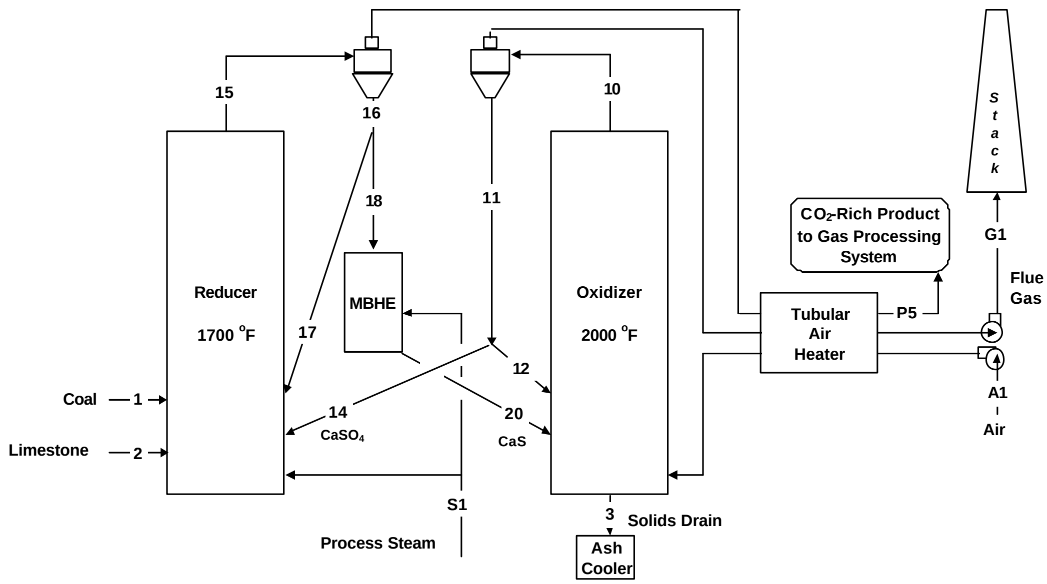


Figure 2.7. 1: Case-7: Simplified Boiler Island Gas Side Process Flow Diagram

Oxidizer:

The purpose of the oxidizer is to oxidize the incoming oxygen deficient bed solids (Stream 20), which is rich in CaS, with the incoming air stream. In this way the solids are used as a chemical looping oxygen carrier. The oxidizer operates at about 2,000 ^{o}F. The basic chemistry in the oxidizer is shown in the following reaction.

$$CaS + 2O_2 \rightarrow CaSO_4 + Heat$$

The oxidizer bed temperature is maintained at an optimum level for sulfur capture and combustion efficiency by balancing solids flow between an uncooled stream (Stream 12) flowing directly back to the reducer and a cooled stream (Stream 20) which flows through the MBHE and then to the oxidizer vessel.

Reducer:

The purpose of the reducer is to use the oxygen carried in with the solids (Stream 14), which is rich in $CaSO_4$, to combust the coal and the residual carbon contained in the recycle solids (Stream 17). The oxygen now carried by the solids ($CaSO_4$ in Stream 14) is reacted with the carbon and hydrogen contained within the coal (Stream 1) in the reducer vessel to form a high CO_2 content exhaust stream (Stream 15) which also includes water vapor and entrained solids. The reducer operates at about 1,700 ^{o}F. The basic chemistry in the reducer is shown in the following overall reactions.

$$CaSO_4 + 2C + heat \rightarrow 2CO_2 + CaS$$
$$CaSO_4 + 4H_2 + heat \rightarrow 4H_2O + CaS$$

The solids leaving the reducer, now rich in oxygen deficient (CaS) are separated from the gas (Stream 16) and returned to the Oxidizer (Stream 20) after passing through the MBHE to complete the solids loop.

Limestone (Stream 2) is continuously added to the reducer to remove sulfur from the coal. The limestone and the sulfur combine to form CaS in the reducer, which is used in the chemical looping reactions described above.

The temperature in the reducer is controlled to the proper level by splitting the flow of hot recirculated solids leaving the cyclone, between an uncooled stream (Stream 17) that flows directly back to the reducer and a stream (Stream 18) that goes to the Moving Bed Heat Exchanger (MBHE), where the solids are cooled before returning to the oxidizer.

Gas Cooling:

There is a high nitrogen content gas stream (Stream 10) leaving the oxidizer, which is cleaned of solids, cooled in a tubular air heater and finally exhausted to the atmosphere after passing through the Induced Draft (ID) fan.

Similarly, there is a high CO_2 content gas stream (Stream 15) leaving the reducer. This stream is cleaned of solids, cooled in a tubular air heater and provides the feed stream (Stream P5) to the Gas Processing System whereby a high purity CO_2 stream is produced and available for sequestration or usage.

The cooling of the two gas streams leaving the reactor vessels is done in a high temperature tubular air heater where the sensible heats of the high nitrogen content and high CO_2 content streams are transferred to the incoming air stream (Stream A1).

Particulate Removal:

Stream 15, flue gas comprised of primarily CO_2 and H_2O vapor and entrained hot solids, flows through a particulate removal device, where hot solids are removed and recirculated.

Stream 10, flue gas comprised of primarily N_2 and a small amount of O_2 and entrained hot solids, flows through a particulate removal device, where hot solids are removed and recirculated.

Moving Bed Heat Exchanger:

The purpose of the Moving Bed Heat Exchanger (MBHE) is to generate high-pressure, high-temperature steam for the power cycle while cooling the entering solids stream. The MBHE is designed to cool the entering solids stream (Stream 18) by evaporating, superheating, and reheating steam for the power cycle. The MBHE contains all the pressure parts in the Boiler Island. The moving bed heat exchangers are not fluidized and contain several immersed spiral-finned tube bundles, which cool the hot solids that are supplied from the particulate removal system at the top of the reducer. The tube bundles in the MBHE utilize spiral-finned surface and include superheater, reheater, evaporator and economizer sections. Very high heat transfer rates are obtained in the MBHE's due to the conduction heat transfer mechanism between the solids and tube.

The cooled solids stream (Stream 20) leaving the MBHE is transported to the oxidizer vessel to complete the solids loop. This stream is cooled to 600 ^{o}F in the MBHE.

Process Steam:

A small quantity of process steam (Stream S1) is introduced in the MBHE and the reducer. The purpose of this steam is to help initiate the reducer reactions. This steam is provided from an extraction point on the steam turbine.

Ash Removal:

Draining hot solids from the oxidizer through water-cooled fluidized bed ash coolers (Stream 3) controls solids inventory in the system while recovering heat from the hot ash. This stream is rich in $CaSO_4$ and innert Ash while also containing smaller amounts of $CaCO_3$, CaO, and Carbon. The cooling water used for the ash coolers is feedwater from the final extraction feedwater heater of the steam cycle. In this way the sensible heat of the ash is efficiently recovered in the steam cycle

2.7.1.2. Material and Energy Balance

Table 2.7.1 shows the Boiler Island material and energy balance for Case-7. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-7 simplified PFD for the Boiler Island (Figure 2.7.1). This performance was calculated at MCR conditions for this unit.

The MCR condition for Case-7 is defined as high-pressure turbine inlet conditions of 1,400,555 lbm/hr, 1,815 psia, and 1,000 ^oF and intermediate-pressure turbine inlet conditions of 1,304,710 lbm/hr 469 psia 1,000 ^oF. These conditions were very similar to those used for the Base Case (Case-1) differing only in the reheat flow where for Case-1 1,305,632 lbm/hr was used. The slight reduction in reheat steam flow for this case is the result of a small process steam extraction from the steam turbine for Case-7.

The Case-7 boiler was fired with 20 percent excess air same as for the Base Case. The resulting boiler efficiency calculated for Case-7 was 94.80 percent (HHV basis).

Table 2.7. 1: Case-7 Boiler Island Gas Side Material and Energy Balance

Constituent	Units	Input Streams				Output Streams		
		1	2	A1	S1	G1	P5	3
		Coal	Limestone	Combustion Air	Steam	Flue Gas	Product Gas	Solids Drain
C	(lb/hr)	101406						
H	"	5834						
O	"	5164						
N	"	2386						
S	"	3824						
CO_2	"					11807	383547	
H_2	"							
H_2O(gas)	"			20646	75309	20852	91343	
O_2	"			368177		56460	517	
N_2	"			1219968		1232168	2386	
H2O(liquid)	"	6521						
CaCO3	"		23878					239
CaO	"							6555
$Ca(OH)_2$	"							
$CaSO_4$	"							16239
CaS	"							
Carbon	"							806
Ash	"	38291	1257					39547
Coal	"	163425						
H_2O(gas)	"					42624		
Chem SubStream Solids	"	163425	25135					63386
Grand Total	**"**	**163425**	**25135**	**1608791**	**75309**	**1363911**	**477794**	**63386**
Temp	(^oF)	80	80	80	360	152	170	520
Press	(psia)	14.7	14.7	14.7	150	14.7	14.7	14.7
Hs (Sensible Heat)	(MMBtu/hr)				10	22.5	11.7	5.9
Hf (Heat of Formation)	"	-298.2	-132.5	-119.2	-435	-412.0	-2002.9	-378.5
Total Energy	"	-298.2	-132.5	-119.2	-425	-389.5	-1991.2	-372.6

2.7.1.3. Boiler Island Equipment

This section describes major equipment included in the Boiler Island for Case-7. The major components included in the Boiler Island include the Reducer vessel, Oxidizer vessel, ash coolers, fuel feed system, fuel silos, sorbent feed system, sorbent silo, particulate removal system, seal pots, external moving bed heat exchanger (MBHE), superheater, reheater, evaporator, economizer, high temperature air heater, and draft system.

Figures 2.7.2 and 2.7.3 show general arrangement drawings of the Case-7 CMB boiler. The plan area for the Case-7 Boiler Island is about 71 percent of that for Case-1. Similarly, the building volume for Case-7 is about 66 percent of that for Case-1. The complete Boiler Island Equipment List for Case-7 is shown in Appendix I. Appendix II shows several additional drawings of the Boiler (key plan view, boiler plan view, side elevation, and various sectional views).

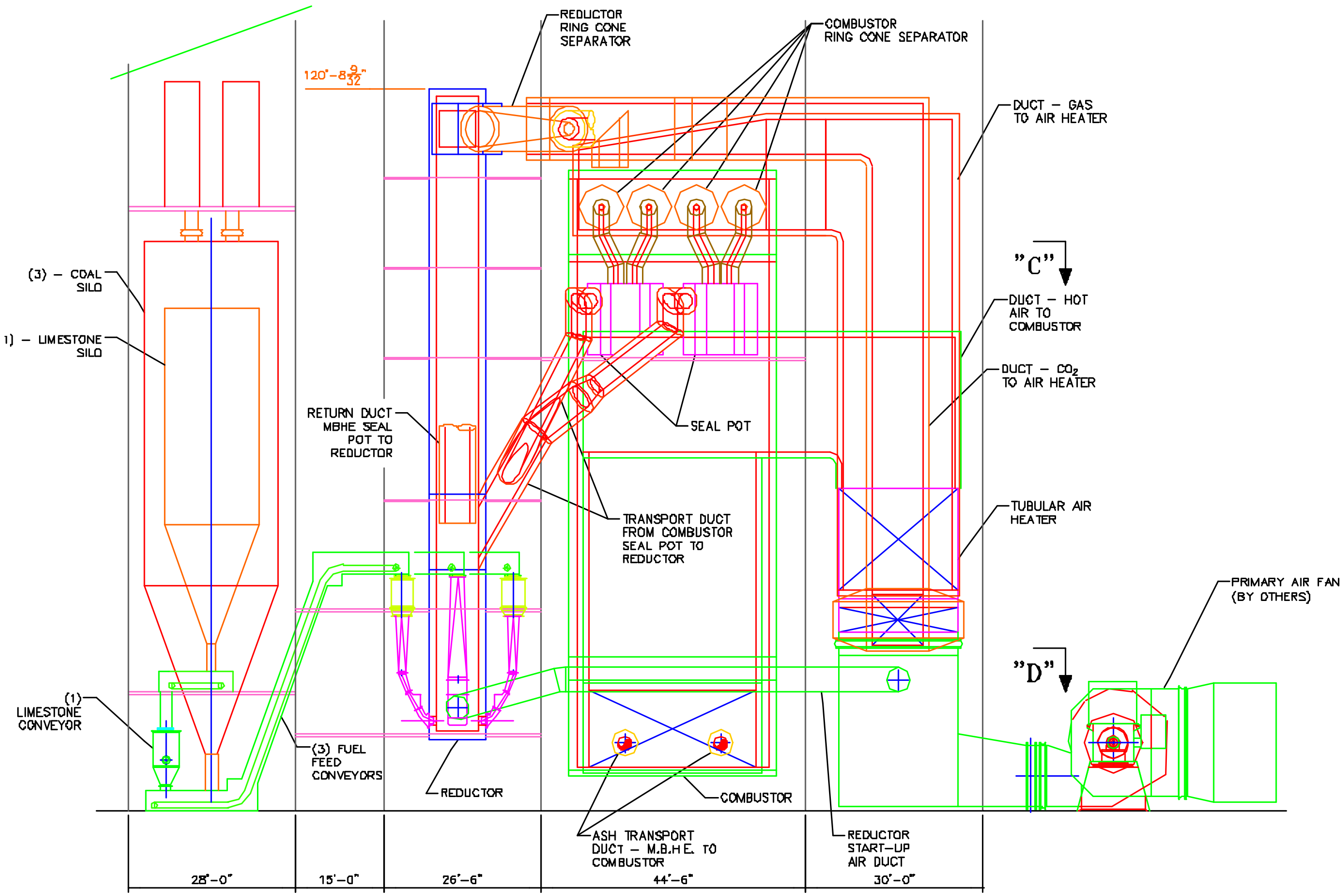


Figure 2.7. 2: Case-7 Boiler Island General Arrangement Drawing – Side Elevation

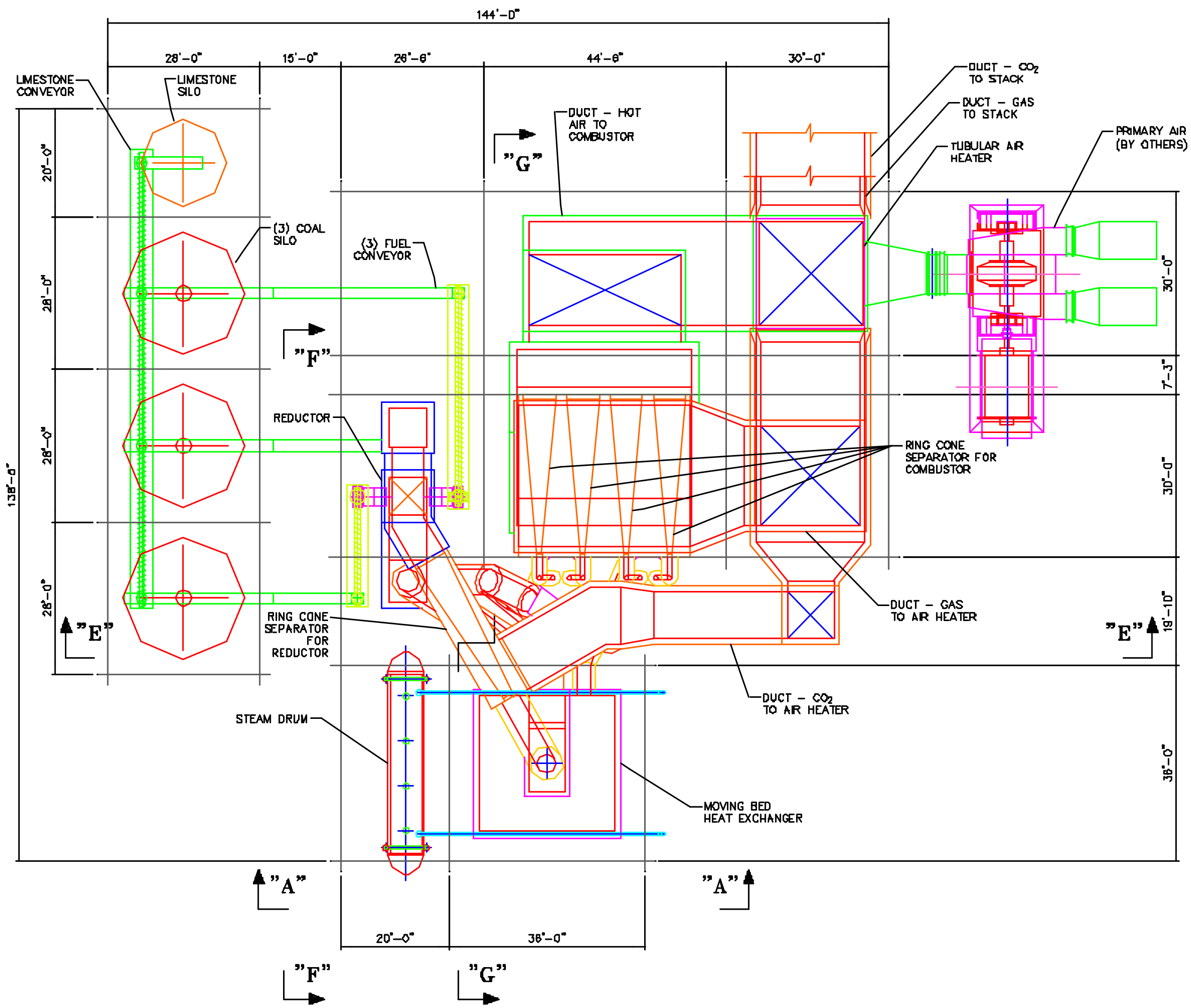


Figure 2.7. 3: Case-7 Boiler Island General Arrangement Drawing – Plan View

Reducer:

The reducer vessel is designed to react the oxygen contained in the oxygen carrying solids stream ($CaSO_4$) with the feed coal, thus producing a high CO_2 content flue gas stream that can be further processed for sequestration or usage. The reducer vessel for Case-7 is about 7 ft wide, 7 ft deep and 100 ft high. Crushed fuel, sorbent, and recycle solids are fed to the lower portion of the reducer.

The reducer vessel is constructed in the same fashion as the Case 4, 5 and 6 combustor. It can be described as a rectangular refractory lined vessel with vertical walls. The lower and upper regions are formed with a multilayer refractory liner without any waterwall panels. The lower reducer has penetrations for the admission of fuel, sorbent, and recycle bed material. These penetrations are similar to those used for other cases in this study.

Oxidizer:

The oxidizer is designed to absorb most of the oxygen contained in the incoming air stream with oxygen deficient recycle solids supplied from the MBHE thus producing an oxygen deficient nitrogen rich gas stream leaving the oxidizer vessel. The oxidizer vessel for Case-7 is about 32 ft wide, 7 ft deep and 100 ft high. Hot air from the air heater, and recycle solids from the MBHE and oxidizer ring cone separator are fed to the lower portion of the oxidizer vessel. The ring cone separator is designed to enhance fine particulate collection.

The oxidizer is constructed in the same fashion as the reducer. It can be described as a rectangular refractory lined vessel with vertical walls. The lower and upper regions are formed with a multilayer refractory liner without any waterwall panels. The lower oxidizer vessel includes several penetrations for the admission of hot air, recycle bed material and the removal of hot ash through the bed drain. These penetrations are similar to those used for other cases in this study.

Fuel Feed System:

The fuel feed system for Case-7 is very similar to the system used for the other cases. It is designed to transport prepared coal from the storage silos to the lower reducer. The system includes the storage silos and silo isolation valves, fuel feeders, feeder isolation valves, and fuel piping to the reducer.

Sorbent Feed System:

The limestone feed system for Case-7 is the same as for the other cases. The limestone feed system pneumatically transports prepared limestone from the storage silos to the lower reducer. The system includes the storage silos and silo isolation valves, rotary feeders, blower, and piping from the blower to the reducer injection ports.

Ash Coolers:

The ash cooler design for Case-7 is the same as for the other cases as the ash flow is nearly identical in all cases except for Case-6. Draining hot solids from the oxidizer vessel through two water-cooled ash coolers controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is provided by feedwater from the final extraction feedwater heater of the steam cycle. The heated water leaving the ash cooler is then combined with water from the economizer located in the convection pass to feed the steam drum.

Particulate Removal:

Flue gas and entrained solids exit the upper reducer vessel and oxidizer vessel and enter their respective ring cone separators. These are extremely high efficiency particle separation devices.

Seal Pots:

The seal pots for Case-7 are of the same design as in other cases. The seal pot is a device that provides a pressure seal between the reducer or oxidizer, which is at relatively high pressure, and the ring cone separator that is at near atmospheric pressure. The seal pot is a non-mechanical valve, which moves solids collected back to the reducer or oxidizer. The seal pot is constructed of steel plate with a multiple layer refractory lining with fluidizing nozzles located along the bottom to assist solids flow. Some of the solid flows directly from the seal pot back to the reducer or oxidizer while other solids are diverted through a plug valve. The diverted solids collected from the reducer flow through the external Moving Bed Heat Exchangers (MBHE), and then back to the oxidizer. The diverted solids collected from the oxidizer flow directly to the reducer.

Convection Pass:

There is no traditional convection pass containing pressure parts in Case-7and the gas streams leaving the ring cone separators located at the outlets of the reducer and oxidizer vessels are simply ducted directly to the High Temperature Air Heater for heat recovery.

Moving Bed Heat Exchanger:

The external heat exchanger for Case-7 is a single moving bed. The moving bed heat exchanger is not fluidized and contains several immersed tube bundles, which cool the hot solids leaving the reducer seal pot before the cooled solids return to the lower part of the oxidizer. The tube bundles in the MBHE utilize spiral-finned surface and include superheater, reheater, evaporator and economizer sections. Very high heat transfer rates are obtained in the MBHE due to the conduction heat transfer mechanism between the solids and tube. The MBHE is bottom supported and is constructed using steel plate refractory lined enclosure walls. It is rectangular in cross section with a hopper shaped bottom. The solids move through the bed by gravity at a design velocity of about 150 ft/hr. The cooled solids leaving the MBHE are feed to the oxidizer.

Superheater:

The superheater is divided into two major sections. Saturated steam leaving the steam drum supplies the horizontal low temperature superheater. Steam leaving the low temperature section flows through a de-superheating spray station and thenontothe finishing superheater section. Both sections are located in the external moving bed heat exchanger. There are no superheater banks located in the convection pass for Case-7. The steam leaving the finishing superheater is piped to the high-pressure turbine where it is expanded to reheat pressure and then returned to the low temperature reheat section of the MBHE.

Reheater:

The reheater, also located in the MBHE, is designed as a single section. The steam is supplied to the reheater inlet header from the de-superheating spray station, which is fed from the high-pressure turbine exhaust. The reheater is a horizontal section comprised of spiral-finned tubing and located between the superheat finishing section and the low temperature superheat section. There are no reheater banks located in the convection pass for Case-7. The steam leaving the reheater is returned to the intermediate pressure turbine where it continues its expansion through the intermediate and low-pressure turbines before being exhausted to the condenser.

Evaporator:

The evaporator section for Case-7 is also located in the lower MBHE. The evaporator is comprised of three banks of horizontal tubes, which evaporate high-pressure boiler feedwater. It is located just below the low temperature superheater section. The water/steam mixture exiting the evaporator tube banks is supplied to the steam drum through risers where the steam and water phases are separated. The feedwater supplying the evaporator is piped from the steam drum through circulating water pumps and is comprised of a combination of separated saturated water and subcooled water from the economizer.

Economizer:

The economizer section for Case-7 is also located in the lower MBHE. The economizer is comprised of two banks of horizontal tubes, which heat high-pressure boiler feedwater. The water exiting the economizer tube banks is supplied to the steam drum. The feedwater supplying the economizer is piped from the final extraction feedwater heater and the ash coolers.

Draft System:

The flue gas is moved through the Boiler Island equipment with the draft system. The draft system includes the primary air (PA) fan, the induced draft (ID) Fan, the associated ductwork, and expansion joints. The induced draft, and PA fan are driven with electric motors and are controlled to operate the unit in a balanced draft mode with the oxidizer vessel and reducer vessel outlet streams maintained at a slightly negative pressure (typically, -0.5 inwg).

High Temperature Air Heater:

A tubular regenerative air heater is used to cool the two gas streams (leaving the oxidizer and reducer) by heating the primary air stream prior to combustion in the system. This is a very high temperature air heater and is considered a development item.

2.7.2. Case-7 Gas Processing System Process Description and Equipment

This purpose of this system is to processes the CO_2 rich flue gas stream leaving the Case-7 Boiler Island to provide a liquid CO_2 product stream of suitable purity for an EOR application.

The Case-7 CO_2 capture system is designed for about 100 percent CO_2 capture. Cost and performance estimates were developed for all the systems and equipment required to cool, clean, compress and liquefy the CO_2, to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO_2 specification for EOR, given in Table 2.0.1, was used as the basis for the CO_2 capture system design.

A very low concentration of oxygen, in particular, is specified for meeting current pipeline operating practices, due to the corrosive nature of the oxygen. Hence, for Case-7, whereby the final CO_2 liquid product was found to contain about 1,800 ppmv of O_2, the design of the transport pipe to an EOR site for example would have to take this characteristic under consideration.

The nitrogen concentration specified in Table 2.0.1 is < 300 ppmv. It should be noted that according to Charles Fox of Kinder Morgan (Fox, 2002), this specification is very conservative as his company specifies a maximum nitrogen concentration of 4 percent (by volume) to control the minimum miscibility pressure. In Case 7 the nitrogen concentration in the liquid product was 9,700 ppmv. The exact reasoning behind the very low nitrogen specification listed in Table 2.0.1 is not clear.

2.7.2.1. Process Description

The following describes a CO_2 recovery system that compresses and then cools a CO_2 rich gas stream from an advanced air-fired Chemical Looping type boiler to a pressure high enough so CO_2 can be liquefied. The resulting liquid CO_2 is pumped to a high pressure so it can be economically transported for sequestration or usage. Pressure in the transport pipeline will be maintained above the critical pressure of CO_2 to avoid 2-phase flow.

The key process parameters (pressures, temperatures, duties etc.) are shown in the material and energy balance tables and will not be repeated here except in selected instances.

Figure 2.7.4 shows the Flue Gas Cooling process flow diagram and Figure 2.7.5 shows the Flue Gas Compression and Liquefaction process flow diagram.

Flue Gas Cooling:

Please refer to Figure 2.7.4 (drawing D 12173-07001-0).

The feed to the Gas Processing System is the flue gas stream that leaves the High Temperature Air Heater of the Boiler Island. At this point, the flue gas is near the dew point of H_2O. All of the flue gas leaving the boiler is cooled to 100 °F in Gas Cooler DA-101 which operates slightly below atmospheric pressure. A significant amount of water condenses out in this cooler. Excess condensate is blown down to the cooling water system. A single vessel has been provided for this cooler.

The Gas Cooler is configured in a packed tower arrangement where the flue gas is contacted with cold water in countercurrent fashion. Warm water from the bottom of the contactor is recycled back to the top of the contactor by Water Pump GA-101 after first cooling it in an external water cooled heat exchanger, Water Cooler EB-101 (plate and frame exchanger). The cooling water for this exchanger comes from the new cooling tower.

Because the flue gas may carry a small amount of fly ash, the circulating water is filtered in Water Filter FD-101A-F to prevent solids build-up in the circulating water. Condensate blowdown is filtered and is taken out downstream of the filter. However, the stream is not cooled and is split off before EB-101. Thus the heat load to the cooling tower is minimized.

From the Gas Cooler the gas stream proceeds to the gas compression area. The Gas Cooler minimizes the volumetric flow rate to, and the resulting power consumption of, the Flue Gas Compression equipment located downstream.

Three-Stage Gas Compression System:

Please refer to Figure 2.7.5 (drawing D 12173-07002-0).

The compression section, where the CO_2 rich stream is compressed to 311 psia by a three-stage centrifugal compressor, includes Gas Compressor GB-2301. This three-stage compressor set includes a series of gas coolers (aftercoolers) located after each compression stage. Following the third stage aftercoolers, the stream is then further cooled in a propane chiller to a temperature of –22 °F. Note that both the trim cooling water and water for the propane condenser comes from the cooling tower. Following the compression and liquefaction steps, the pressure of the liquid is boosted to 2,018 psia by CO_2 Pipeline Pump GA-2301. This stream is now available for sequestration or usage.

The volumetric flow to the compressor inlet is about 65,000 ACFM. The discharge pressures of the stages have been balanced to give reasonable power distribution and discharge temperatures across the various stages. They are:

- 1st Stage 40 psia
- 2nd Stage 110 psia
- 3rd Stage 311 psia

Power consumption for this large compressor has been estimated assuming adiabatic efficiency of 75 percent.

The hot gas stream from each compressor stage is first cooled in an air cooler to 120 ^{o}F in Flue Gas Compressor 1st/2nd/3rd Stage Aftercoolers (EC-2301A/B/C, EC-2302A/B, EC-2303). The gas is then further cooled by water-cooled heat exchangers to 95 ^{o}F in Flue Gas Compressor 1st/2nd Stage Trim Coolers (EA-2301A/B and EA-2302). The gas compressor's 3rd stage cooler (EA-2303) cools the gas to 90 ^{o}F to reduce the size of the dryers. Due to their large size, many of these heat exchangers consist of multiple shells. Because of highly corrosive conditions, the process side of the coolers must be stainless steel.

Because the flue gas stream leaving the boiler island is nearly saturated, some water condenses out in the three aftercoolers. The sour condensate is separated from the gas in knockout drums (FA-2300/1/2/4) equipped with mist eliminator pads. The condensate from these drums is drained to the cooling tower or to waste water treatment. To prevent corrosion, these drums have a stainless steel liner.

Flue gas leaving the 3rd stage discharge knockout drum (FA-2304) is fed to Flue Gas Drier PA-2351 where additional moisture is removed.

Gas Drying:

Please refer to Figure 2.7.5 (drawing D 12173-07002-0).

It is necessary to dry the CO_2 stream to meet the product specification. A molecular sieve drier has been selected.

The performance of a fixed-bed drier improves as pressure increases. This favors locating the drier at the discharge of the compressor. However, as the operating pressure of the drier increases, so does the design pressure of the equipment. This favors low-pressure operation. But, at low pressure the diameter or number of the drier vessels grows, increasing the cost of the vessel. For this design the drier has been optimally located downstream of the 3rd stage compressor. The CO_2 Drier system consists of four vessels filled with molecular sieve. One vessel is on line while the others are being regenerated. The flow direction is down during operation and up during regeneration.

The drier is regenerated with CO_2 exiting the online driers. After regeneration, heating is stopped while the gas flow continues. This cools the bed down to the normal operating range. The regeneration gas is cooled and recycled back to the drier inlet via a blower.

Regeneration of a molecular sieve bed requires relatively high temperature and, because HP steam pressure may fluctuate, a gas-fired heater has been specified for this service.

Flue Gas Filter FD-102 has been provided at the drier outlet to remove any fines that the gas stream may pick up from the desiccant bed.

CO_2 Condensation:

Please refer to Figure 2.7.5 (drawing D 12173-07002-0).
From the CO_2 Drier, the gas stream is cooled down further to -22 ^{o}F with propane refrigeration in CO_2 Condenser EA-2304A-F.

CO_2 Pumping and CO_2 Pipeline:

Please refer to Figure 2.7.5 (drawing D 12173-07002-0).
The CO_2 product must be increased in pressure to 2,000 psig. A multistage heavy-duty pump (GA-2301) is required for this service. This is a highly reliable derivative of an API-class boiler feed-water pump.

It is important that the pipeline pressure be always maintained above the critical pressure of CO_2 such that single-phase (dense-phase) flow is guaranteed. Therefore, the pressure in the line should be controlled with a pressure controller and the associated control valve located at the destination end of the line.

2.7.2.2. Process Flow Diagrams

Two process flow diagrams are shown below for these systems:

- Figure 2.7.4 (drawing D 12173-07001-0) Flue Gas Cooling PFD
- Figure 2.7.5 (drawing D 12173-07002-0) CO_2 Compression and Liquefaction PFD

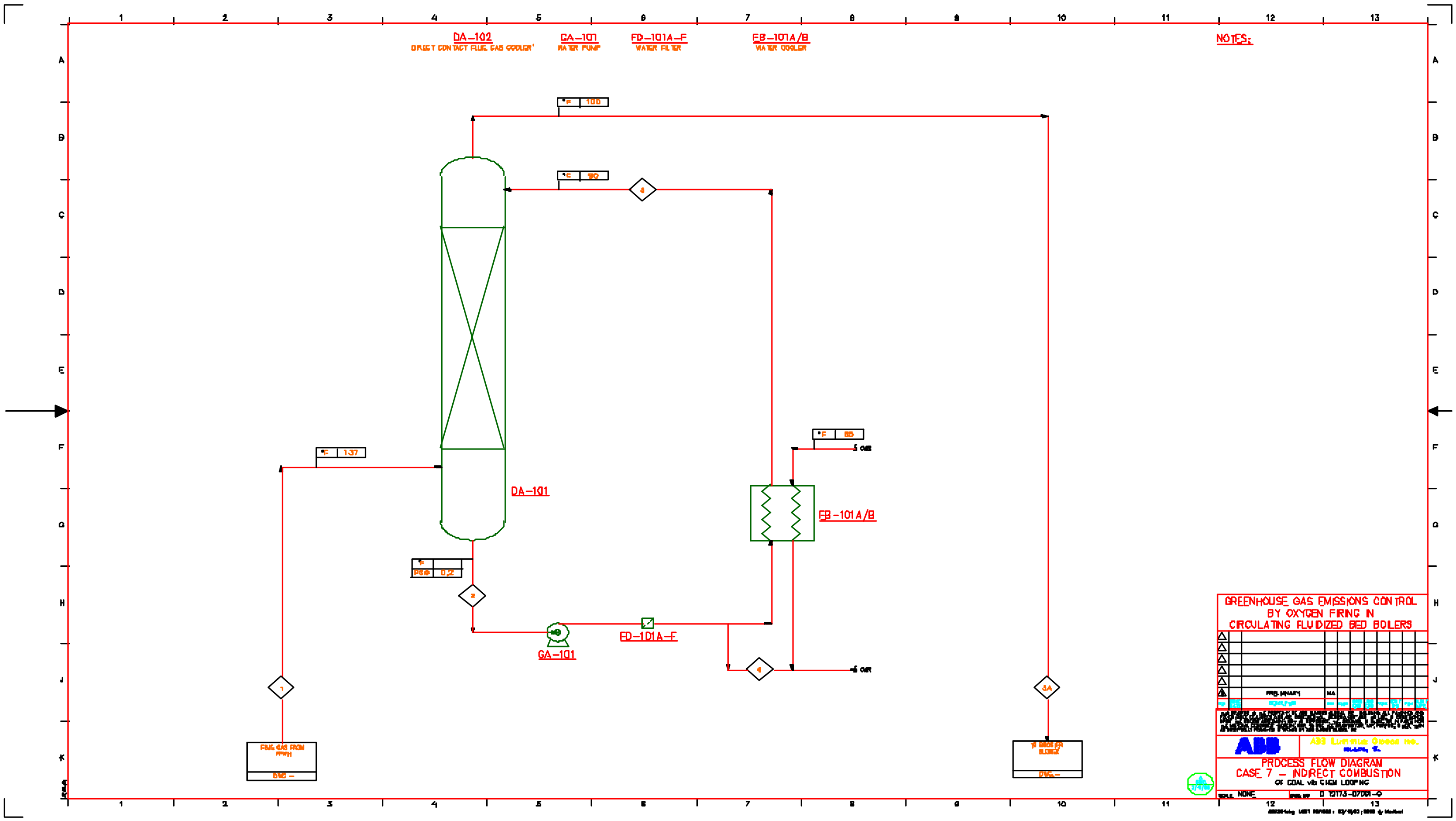


Figure 2.7. 4: Case-7 Flue Gas Cooling Process Flow Diagram

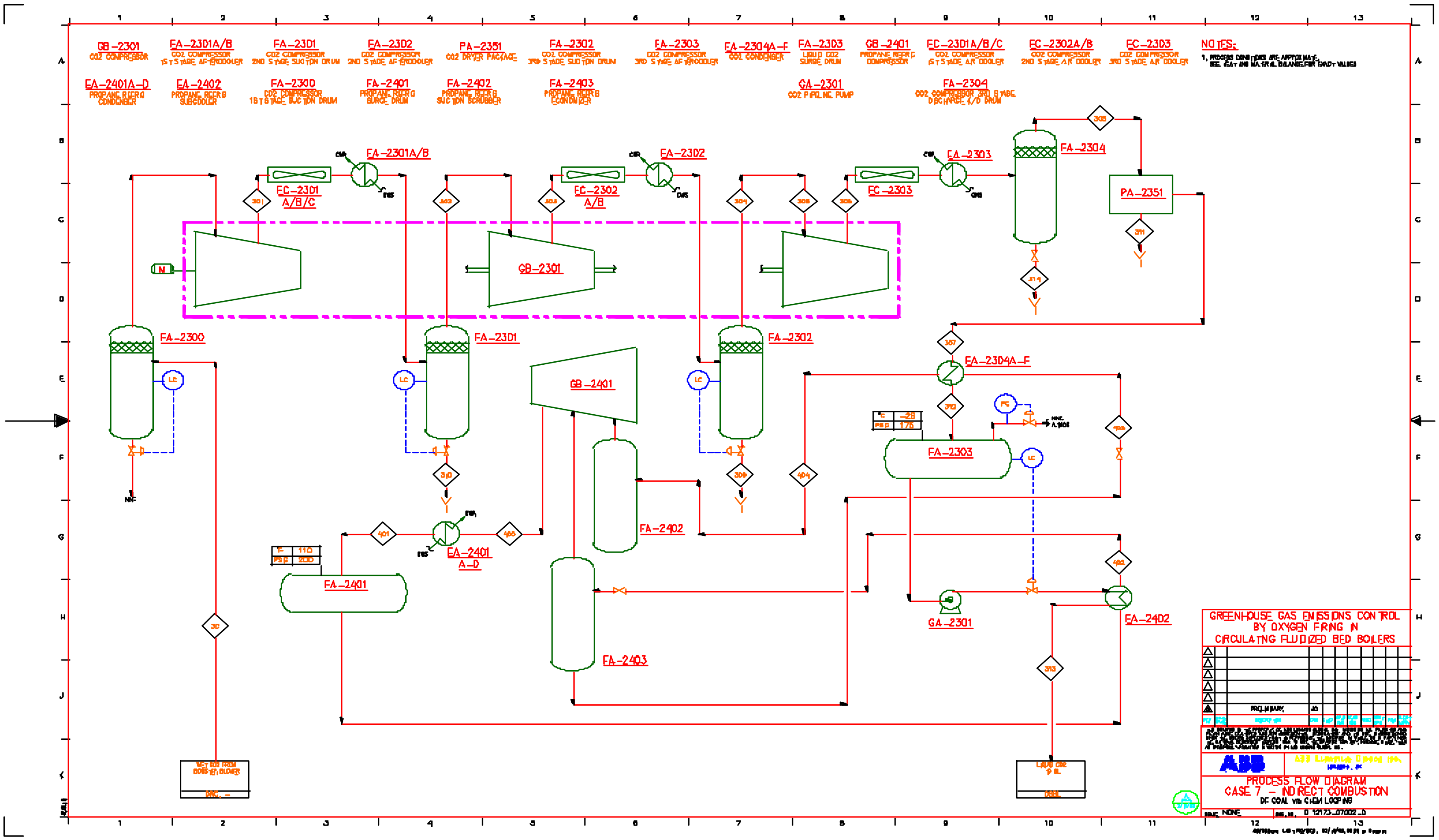


Figure 2.7. 5: Case-7 CO_2 Compression and Liquefaction Process Flow Diagram

2.7.2.3. Material and Energy Balance

Table 2.7.2 shows the material and energy balance for the Case-7 Gas Processing System.

Table 2.7.2: Case-7 Gas Processing System Material & Energy Balance

STREAM NAME		To quench columns	From Quench columns	Excess water	Quench water out	Quench water in	To liquefaction train	First stage discharge	To second stage	First stage water KO	2nd stage discharge	To 3rd stage	2nd stage water KO
PFD STREAM NO.		**1**	**3a**	**6**	**2**	**5**	**3c**	**301**	**302**	**310**	**303**	**304**	**309**
VAPOR FRACTION	Molar	1.000	1.000	0.000	0.000	0.000	1.000	1.000	1.000	0.000	1.000	1.000	0.000
TEMPERATURE	°F	170.0	100	130	130	90	100	279	95	95	301	95	95
PRESSURE	PSIA	14.7	14	55	14	45	14	40	34	34	110	104	104
MOLAR FLOW RATE	lbmol/hr	13,887	9,482.11	4,400.00	123,850	119,450	9,482.30	9,482.30	9,035.34	446.96	9,035.34	8,890.27	145.07
MASS FLOW RATE	lb/hr	477,790	398,400	79,291	2,232,000	2,152,600	398,400	398,400	390,330	8,062	390,330	387,710	2,624
ENERGY	Btu/hr	7.00E+07	4.19E+07	-6.09E+07	-1.71E+09	-1.74E+09	4.21E+07	5.78E+07	3.93E+07	-6.47E+06	5.68E+07	3.79E+07	-2.10E+06
COMPOSITON	Mol %												
CO2		62.76%	91.90%	0.02%	0.02%	0.02%	91.89%	91.89%	96.43%	0.09%	96.43%	98.00%	0.27%
H2O		36.51%	7.03%	99.98%	99.98%	99.98%	7.04%	7.04%	2.45%	99.91%	2.45%	0.86%	99.73%
Nitrogen		0.61%	0.90%	0.00%	0.00%	0.00%	0.90%	0.90%	0.94%	0.00%	0.94%	0.96%	0.00%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		0.12%	0.17%	0.00%	0.00%	0.00%	0.17%	0.17%	0.18%	0.00%	0.18%	0.18%	0.00%
SO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR													
MOLAR FLOW RATE	lbmol/hr	13,886.7	9,482.1	-	-	-	9,482.3	9,482.3	9,035.3	-	9,035.3	8,890.3	-
MASS FLOW RATE	lb/hr	477,790	398,400	-	-	-	398,400	398,400	390,330	-	390,330	387,710	-
STD VOL. FLOW	MMSCFD	126.47	86.36	-	-	-	86.36	86.36	82.29	-	82.29	80.97	-
ACTUAL VOL. FLOW	ACFM	105,790	69,281	-	-	-	64,570	31,134	26,048	-	11,028	8,175	-
MOLECULAR WEIGHT	MW	34.41	42.02	-	-	-	42.02	42.02	43.20	-	43.20	43.61	-
DENSITY	lb/ft³	0.08	0.10	-	-	-	0.10	0.21	0.25	-	0.59	0.79	-
VISCOSITY	cP	0.0131	0.0145	-	-	-	0.0146	0.0199	0.0150	-	0.0214	0.0153	-
LIGHT LIQUID													
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID													
MOLAR FLOW RATE	lbmol/hr	-	-	4,400	123,850	119,450	-	-	-	446.96	-	-	145.07
MASS FLOW RATE	lb/hr	-	-	79,291	2,232,000	2,152,600	-	-	-	8,062.06	-	-	2,623.51
STD VOL. FLOW	BPD	-	-	5,441	153,150	147,710	-	-	-	553	-	-	180
ACTUAL VOL. FLOW	GPM	-	-	160.79	4,526.42	4,290.25	-	-	-	16.10	-	-	5.23
DENSITY	lb/ft³	-	-	61.48	61.48	62.56	-	-	-	62.44	-	-	62.49
VISCOSITY	cP	-	-	0.5043	0.5046	0.7606	-	-	-	0.7185	-	-	0.7503
SURFACE TENSION	Dyne/Cm	-	-	66.91	66.92	70.83	-	-	-	70.30	-	-	70.18

STREAM NAME		From 3rd stage	To drier	3rd stage water KO	From drier / To condenser	Water from drier	From condenser	From product pump	To pipeline	Refrig compressor discharge	From refrig condenser	From subcooler	Refrig to CO2 condenser	Refrig from CO2 condenser
PFD STREAM NO.		306	305	314	307	311	312	308	313	400	401	402	403	404
VAPOR FRACTION	Molar	1.000	1.000	0.000	1.000	0.631	0.993	0.000	0.000	1.000	0.000	0.000	0.209	0.993
TEMPERATURE	°F	287	90	90	90	418	-33	-7	82	164	110	49	-33	-33
PRESSURE	PSIA	311	305	305	300	300	19	2,018	2,015	222	215	212	19	19
MOLAR FLOW RATE	lbmol/hr	8,890.27	8,841.34	48.93	8,813.58	27.76	9,632.74	8,813.58	8,813.58	10,250.00	10,250.00	10,250.00	9,632.74	9,632.74
MASS FLOW RATE	lb/hr	387,710	386,820	892	386,320	500	424,770	386,320	386,320	451,990	451,990	451,990	424,770	424,770
ENERGY	Btu/hr	5.34E+07	3.50E+07	-7.07E+05	3.49E+07	3.17E+04	4.52E+07	-2.25E+07	-3.67E+06	7.82E+07	7.62E+06	-1.13E+07	-1.47E+07	4.52E+07
COMPOSITON	Mol %													
CO2		98.00%	98.54%	0.80%	98.85%	0.00%	0.00%	98.85%	98.85%	0.00%	0.00%	0.00%	0.00%	0.00%
H2O		0.86%	0.31%	99.20%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.96%	0.97%	0.00%	0.97%	0.00%	0.00%	0.97%	0.97%	0.00%	0.00%	0.00%	0.00%	0.00%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%
Oxygen		0.18%	0.18%	0.00%	0.18%	0.00%	0.00%	0.18%	0.18%	0.00%	0.00%	0.00%	0.00%	0.00%
SO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR														
MOLAR FLOW RATE	lbmol/hr	8,890.3	8,841.3	-	8,813.6	17.5	9,569.8	-	-	10,250.0	-	-	2,012.3	9,569.8
MASS FLOW RATE	lb/hr	387,710	386,820	-	386,320	316	422,000	-	-	451,990	-	-	88,737	422,000
STD VOL. FLOW	MMSCFD	80.97	80.52	-	80.27	0.16	87.16	-	-	93.35	-	-	18.33	87.16
ACTUAL VOL. FLOW	ACFM	3,675.20	2,530.89	-	2,572.73	8.34	36,869.51	-	-	4,204.10	-	-	7,752.86	36,869.51
MOLECULAR WEIGHT	MW	43.61	43.75	-	43.83	18.02	44.10	-	-	44.10	-	-	44.10	44.10
DENSITY	lb/ft³	1.76	2.55	-	2.50	0.63	0.19	-	-	1.79	-	-	0.19	0.19
VISCOSITY	cP	0.0216	0.0158	-	0.0158	0.0162	0.0065	-	-	0.0103	-	-	0.0065	0.0065
LIGHT LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	62.90	8,813.58	8,813.58	-	10,250.00	10,250.00	7,620.43	62.90
MASS FLOW RATE	lb/hr	-	-	-	-	-	2,773.77	386,320	386,320	-	451,990	451,990	336,030	2,773.77
STD VOL. FLOW	BPD	-	-	-	-	-	375	32,044	32,044	-	61,083	61,083	45,412	375
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	9.64	712.69	1,045.08	-	1,955.29	1,741.36	1,168.24	9.64
DENSITY	lb/ft³	-	-	-	-	-	35.86	67.58	46.09	-	28.82	32.36	35.86	35.86
MOLECULAR WEIGHT	MW	-	-	-	-	-	44.10	43.83	43.83	-	44.10	44.10	44.10	44.10
VISCOSITY	cP	-	-	-	-	-	0.1849	0.1520	0.0555	-	0.0835	0.1165	0.1849	0.1849
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	14.66	13.32	0.85	-	4.81	8.78	14.66	14.66
HEAVY LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	-	48.93	-	10.24	0.00	-	-	-	-	-	(0.00)	0.00
MASS FLOW RATE	lb/hr	-	-	891.76	-	184.55	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	61	-	13	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	1.77	-	0.44	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	62.78	62.78	-	52.31	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	0.7775	0.7775	-	0.1246	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	70.28	70.28	-	34.48	-	-	-	-	-	-	-	-

2.7.2.4. Gas Processing System Utilities

The following tables define the cooling water, natural gas, and electrical requirements for the Gas Processing System.

Table 2.7.3: Case-7 Gas Processing System Cooling Water and Fuel Gas Requirements

COOLING WATER

REV	Equipment TAG NO	SERVICE	No. Installed	DUTY MMBTU/HR	INLET TEMP, F	OUTLET TEMP, F	FLOWRATE LB/HR
	EA-2301	FG Comp 1 stg trim cooler	1	6.27	85	103	348,485
	EA-2302	FG Comp 1 stg trim cooler	1	3.73	85	103	207,071
	EA-2303	FG Comp 1 stg trim cooler	1	3.36	85	103	186,869
	EA-2402	Refrig Condenser	1	70.64	85	100	4,709,091
	EB-101	Water cooler	1	90.00	85	105	4,500,000
		TOTAL COOLING WATER		174.00			9,951,515

FUEL GAS FUEL GAS VALUE BASIS: 930 BTU/SCF (LHV)

REV	Equipment TAG NO	SERVICE	ONLINE FACTOR	COMPR HP	HEAT RATE BTU/HP-HR	DUTY MMBTU/HR	EFFICIENCY %	FLOWRATE (Peak) MMSCFD	FLOWRATE (Peak) SCFH	FLOW (Avg) MMSCFD
	FH-101	Mole sieve regeneration	72%			7.90	80%	0.255	10,618	0.183
		TOTAL FUEL GAS				7.90		0.255	10,618	0.183

Table 2.7.4: Case-7 Gas Processing System Electrical Requirements

Number of Trains	Item Number	Service	Number Operating per train	Power (ea) including 0.95 motor eff	Total all trains
				(kW)	(kW)
1	EC-101	Flue Gas Compressor 1st Stage Aftercooler	1	60	60
1	EC-102	Flue Gas Compressor 2nd Stage Aftercooler	1	50	50
1	EC-103	Flue Gas Compressor 3rd Stage Aftercooler	1	46	46
1	PA-2352	Drier Package	1	347	347
1	GB-101	1st Stage	1	4847	4847
1		2nd Stage	1	5400	5400
1		3rd Stage	1	4791	4791
1	GB-102	1st Stage	1	4698	4698
1		2nd Stage	1	4451	4451
1	GA-103	CO2 Pipeline pump	1	763	763
		Total			**25453**

2.7.2.5. Gas Processing System Equipment

The equipment list for the Gas Processing System is provided in Appendix I, Section 9.1.7.2.

2.7.3. Case-7 Balance of Plant Equipment and Performance

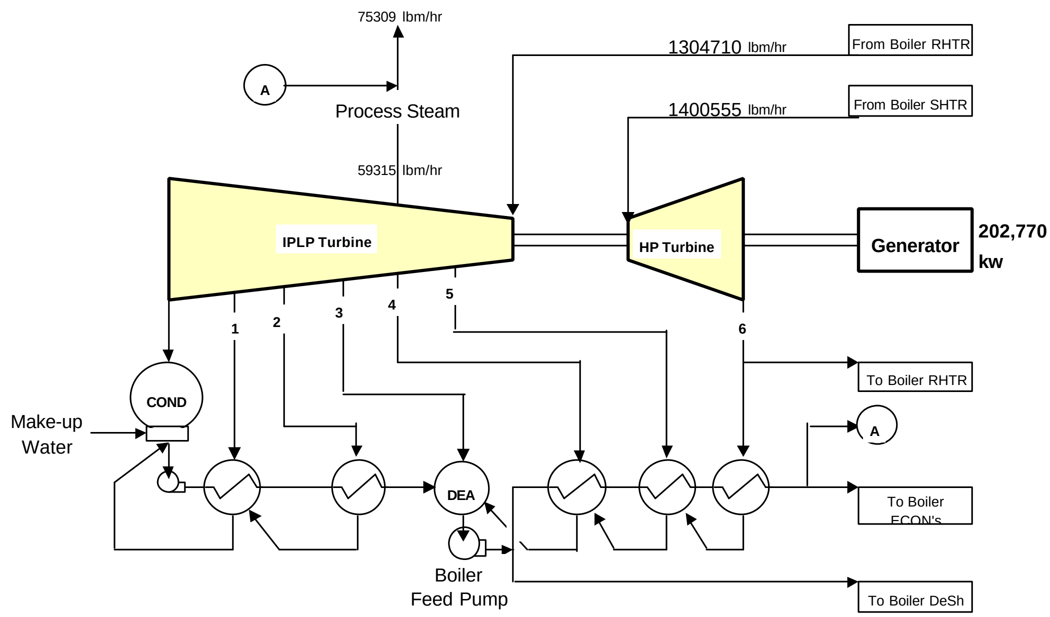
The balance of plant equipment described in this section includes the steam cycle performance and equipment, the draft system equipment, the cooling system equipment, and the material handling equipment (coal, limestone, and ash). Refer to Appendix I for equipment lists and Appendix II for drawings.

2.7.3.1. Steam Cycle Performance

The steam cycle for Case-7 is shown schematically in Figure 2.7.6. The Mollier diagram which illustrates the process on enthalpy - entropy coordinates is the same as for Case-1 and is not repeated here. The steam cycle arrangement and performance is slightly different than Cases 2, 3, and 4 and very similar to that of Case-5. In this case a small amount of low-pressure process steam, which is required for the boiler, is extracted from the low-pressure turbine and de-superheated. It should be borne in mind that no process steam was used in Cases 1, 2, 3 and 4.

The high-pressure turbine expands 1,400,555 lbm/hr of steam at 1,800 psia and 1,005 ^{o}F, the same as for all cases. Reheat steam (1,304,710 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,005 ^{o}F. The condenser pressure used for Case-7 and all other cases in this study was 3.0 in. Hga. The steam turbine performance analysis results show the generator produces 202,770 kW output and the steam turbine heat rate is about 8,404 Btu/kWh. The generator output and condenser losses are slightly lower than for the other cases due to the process steam requirement.

Turbine heat rate is somewhat higher for Case-7 than other cases also as a result of the process steam.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Steam Turbine Power Output	704	Boiler Heat Input	1673
Process Steam Heat Loss	84	BFP & CP Input	12
Condenser Loss	898	**Total Energy Input**	1685
Total Energy Output	1685	**In - Out**	0

Turbine Heat Rate 8404.3 (Btu/kwhr)

Figure 2.7. 6: Case-7 Steam Cycle Schematic and Performance

2.7.3.2. Steam Cycle Equipment

This section provides a brief description of the major equipment (steam turbine, condensate and feedwater systems) utilized for the steam cycle of this case.

Steam Turbine:

The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the boiler passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheated steam flows through the reheat stop valves and intercept valves and enters the IP section at 465 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. A small amount of steam is extracted for process steam as required in the Boiler Island for this case. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser.

The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for economic unit cycling.

Condensate and Feedwater Systems:

The function of the condensate system is to pump condensate from the condenser hot well to the deaerator, through the gland steam condenser and the LP feedwater heaters. The system consists of one main condenser; two 50 percent capacity, motor-driven vertical condensate pumps; one gland steam condenser; two LP heaters, and one deaerator with a storage tank.

Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump feedwater from the deaerator storage tank to the boiler economizer. Two motor-driven boiler feed pumps are provided to pump feedwater through the three stages of HP feedwater heaters. Pneumatic flow control valves control the recirculation flow. In addition, the suctions of the boiler feed pumps are equipped with startup strainers, which are utilized during initial startup and following major outages or system maintenance.

2.7.3.3. Other Balance of Plant Equipment

The systems for draft, solids handling (coal, limestone, and ash), cooling, electrical, and other BOP systems are described in this section for Case-7.

Draft System:

The flue gas is moved through the reducer, air heater and other Boiler Island equipment with the draft system. The draft system includes the primary air fans, the induced draft (ID) Fan, the transport air blowers, the associated ductwork and expansion joints and the Stack, which disperses the flue gas leaving the system to the atmosphere. The induced draft, primary air fans, and transport air blowers are driven with electric motors and controlled to operate the unit in a balanced draft mode with the ring cone separator inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

A forced draft primary air fan provides air to the oxidizer, which is split into several flow paths, as follows:

- An air stream flows to the fuel feeders and flows with the fuel into the furnace via a complement of fuel / air downcomers and feed spouts. This air stream provides initial fluidization of the coal mixture.
- A second air stream flows to, and cools, the complement of two ash coolers.
- A third air stream flows through a steam coil air heater followed by a regenerative air heater; this preheated air then flows to the coal feed spouts. This air stream acts to sweep the fuel / air mixture into the furnace and to support the initial stages of combustion. This air stream is also used for pre-mixing and firing of natural gas or No. 2 oil used for startup and warm-up.

A forced draft secondary air fan provides an air stream that is preheated in a steam coil air heater and a regenerative air preheater, and is then introduced into the furnace as secondary air.

On the oxidizer side, an oxygen depleted air stream exits the oxidizer and flows through ring cone separators, which separate out solids. These solids are recycled back to the oxidizer and reducer, passing through J-valves, or seal pots, located below the separators.

The gas exiting the oxidizer ring cone separators passes directly to the tubular air preheater and then exits the Chemical Looping steam generator. The gases are drawn through the system with the Induced Draft Fan and then are discharged to atmosphere through the Stack.

On the reducer side, a CO_2 rich stream exits the reducer and flows through ring cone separators, which separate out solids. These solids, after passing through J-valves or seal pots located below the separators, are recycled back to the reducer and oxidizer after being cooled in the MBHE.

The CO_2 rich gas exiting the reducer ring cone separators passes directly to the tubular air preheater and then exits the CFB steam generator and flows to the Gas Processing System (GPS). The gases are drawn through the system with the GPS compressor system.

The following fans and blowers are provided with the scope of supply of the Chemical Looping steam generator:

- Primary air fan, which provides forced draft primary airflow. This fan is a centrifugal type unit, supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.7.5). The electric power required for the electric motor drive is 2,559 kW.

Table 2.7. 5: Primary Air Fan Specification

Gas Analysis			
Oxygen	(wt percent)	22.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	1592703	1911244
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	16.40	
Pressure Rise	(in wg)	47.0	61.1

- The Transport Air Blower provides transport air for solids transport. This fan is a centrifugal type unit supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.7.6). The electric power required for the electric motor drive is 46 kW.

Table 2.7. 6: Transport Air Blower Specification

Gas Analysis			
Oxygen	(wt percent)	22.89	
Nitrogen	"	75.83	
Water Vapor	"	1.28	
Carbon Dioxide	"	0.00	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	16088	19305
Gas Inlet Temperature	(Deg F)	80.0	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	17.84	
Pressure Rise	(in wg)	87.0	113.1

- Induced draft fan, a centrifugal unit supplied with electric motor drive and inlet damper (see Table 2.7.7). The electric power required for the electric motor drive is 1,117 kW.

Table 2.7. 7: Induced Draft Fan Specification

Gas Analysis			
Oxygen	(wt percent)	4.14	
Nitrogen	"	90.34	
Water Vapor	"	4.65	
Carbon Dioxide	"	0.87	
Sulfur Dioxide	"	0.00	
Total	"	100.00	
Operating Conditions			Design Spec
Mass Flow Rate	(lbm/hr)	1363911	1636694
Gas Inlet Temperature	(Deg F)	152.2	
Inlet Pressure	(psia)	13.87	
Outlet Pressure	(psia)	14.70	
Pressure Rise	(in wg)	23.0	29.9

Ducting and Stack:

One stack is provided with a single 19.5-foot-diameter FRP liner. The stack is constructed of reinforced concrete, with an outside diameter at the base of 70 feet. The stack is 480 feet high for adequate dispersion of criteria pollutants, to assure that ground level concentrations are within regulatory limits. Table 2.7.8 shows the stack design parameters.

Table 2.7.8: Stack Design Summary

Design Parameter	Value
Flue Gas Temperature, ^oF	150
Flue Gas Flow Rate, lbm/h	1,363,911
Flue Gas Flow Rate, acfm	555,505
Particulate Loading, grains/acfm	nil

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1/4" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the three silos.

Technical Requirements and Design Basis

- Coal burn rate:
  - Maximum coal burn rate = 163,425 lbm/h = 81.7 tph plus 10 percent margin = 90 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
  - Average coal burn rate = 142,000 lbm/h = 71 tph (based on MCR rate multiplied by an 85 percent capacity factor)
  - Coal delivered to the plant by unit trains:
  - One and one-half unit trains per week at maximum burn rate
  - One unit train per week at average burn rate
  - Each unit train shall have 10,000 tons (100-ton cars) capacity
  - Unloading rate = 9 cars/hour (maximum)
  - Total unloading time per unit train = 11 hours (minimum)
  - Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
  - Reclaim rate = 300 tph
  - Storage piles with liners, run-off collection, and treatment systems:

- Active storage = 6,600 tons (72 hours at maximum burn rate)
- Dead storage = 50,000 tons (30 days at average burn rate)

Table 2.7. 9: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	92
Active Storage, tons	6,600
Dead Storage, tons	50,000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,000 ton silo to accommodate 3 days operation.

Bottom Ash Removal:

Bottom ash, or bed drain material, constitutes all of the solid waste material discharged by the Chemical Looping steam generator. This bottom ash is discharged through a complement of two bed coolers (any one of which must be able to operate at 100 percent load on the design coal). The stripper/coolers cool the bed material to a temperature in the range of 300 ^{o}F (design coal) to a maximum of 500 ^{o}F (worst fuel) prior to discharge via rotary valves to the bed material conveying system. The steam generator scope terminates at the outlets of the rotary valves.

Fly Ash Removal:

There is no significant amount of fly ash leaving the Boiler Island in this case. All ash collected in the ring cone separators (which are extremely high efficiency particulate collection devices) is recycled within the boiler island such that all ash leaves the system as bottom ash.

Ash Handling:

The function of the ash handling system is to convey, prepare, store, and dispose of the bottom ash produced on a daily basis by the boiler. The scope of the system is from the bottom ash hoppers to the truck filling stations.

The bottom ash from the boiler is drained from the ash coolers, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. Ash from the fluidized-bed ash coolers is drained to a complement of screw coolers, which discharge the cooled ash to a drag chain conveyor for transport to a surge bin. The ash is pneumatically conveyed to the bottom ash silo from the surge bin.
The silos are sized for a nominal holdup capacity of 36 hours of full-load operation (1,140 tons capacity) per each. At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately 30 truck loads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.7.10: Ash Handling System Design Summary

Design Parameter	Value
Ash from Boiler, lbm/h	63,386
Ash temperature, ^{o}F	520

Circulating Water System:

The function of the circulating water system is to supply cooling water to condense the main turbine exhaust steam. The system consists of two 50 percent capacity vertical circulating water pumps, a multi-cell mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Condenser Analysis:

The condenser system analysis is detailed in Table 2.7.11.

Table 2.7.11: Condenser Analysis

Item	Value	Units
Pressure	3.0	in. Hga
M stm-in	969,726	lbm/h
T stm-in	115.1	°F
P stm-in	1.474	psia
H stm-in	1051.7	Btu/lbm
M drain-in	110,141	lbm/h
H drain-in	89.7	Btu/lbm
M make-up	75,309	lbm/h
H make-up	83.0	Btu/lbm
H condensate	83.0	Btu/lbm
M condensate	1,094,702	lbm/h
Q condenser	940.5	10^6 Btu/h

Waste Treatment System:

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment is housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge de-watering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water are provided. A 200,000-gallon storage tank provides a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide distributed control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Plant Layout and Plot Plan:

The Case-7 plant is arranged functionally to address the flow of material and utilities through the plant site. A plan view of the boiler, power-generating components, and overall site plan for the entire plant is shown in Appendix II.

2.7.4. Case-7 Overall Plant Performance and CO_2 Emissions

Overall plant performance and emissions for Case-7 are summarized in Table 2.7.12. The Case-1 (Base Case) values are also listed along side for comparison purposes.

Boiler efficiency for Case-7 is calculated to be 94.42 percent (HHV basis) as compared to 89.46 percent for the Base Case. The improvement is primarily due to the reduced dry gas loss resulting from additional low-level heat recovery.

The steam cycle thermal efficiency including the boiler feed pump debit is about 40.61 percent as compared to 41.89 percent for Case-1. The slight reduction is due to a small amount of process steam, which is required for the Case-7 Boiler Island.

The net plant heat rate and thermal efficiency for Case-7 are calculated to be 11,051 Btu/kWh and 30.88 percent respectively (HHV basis).

Auxiliary power for Case-7 is 38,286 kW (about 18.9 percent of generator output). The large auxiliary power increase, as compared to the Base Case, is due primarily to the large power requirement of the gas compression equipment in the Gas Processing System for Case-7.

The resulting net plant output for Case-7 is 164,484 kW or about 85 percent of the Base Case net output.

Carbon dioxide emissions for Case-7 are 1,034 lbm/hr or about 0.01 lbm/kWh on a normalized basis. This represents less than 1 percent of the Case-1 normalized CO_2 emissions and a CO_2 avoided value of 1.99 lbm/kWh.

Table 2.7. 12: Case-7 Overall Plant Performance and Emissions

		CFB Air Fired (Case 1)	2 x CFB Chem Loop (Case 7)
Auxiliary Power Listing	(Units)		
Induced Draft Fan	(kW)	2285	1117
Primary Air Fan	(kW)	2427	2559
Secondary Air Fan	(kW)	1142	n/a
Fluidizing Air Blower	(kW)	920	n/a
Transport Air Fan	(kW)	n/a	46
Gas Recirculation Fan	(kW)	n/a	n/a
Coal Handling, Preperation, and Feed	(kW)	300	293
Limestone Handling and Feed	(kW)	200	173
Limestone Blower	(kW)	150	130
Ash Handling	(kW)	200	189
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	n/a
Boiler Feed Pump	(kW)	3715	3757
Condensate Pump	(kW)	79	79
Circulating Water Pump	(kW)	1400	1563
Cooling Tower Fans	(kW)	1400	1563
Steam Turbine Auxilliaries	(kW)	200	187
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719
Transformer Loss	(kW)	470	456
Subtotal	(kW)	16007	12832
	(frac. of Gen. Output)	0.077	0.063
Air Separation Unit	(kW)	n/a	n/a
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a
CO2 Removal System Auxiliary Power	(kW)	n/a	25453
Total Auxilary Power	(kW)	16007	38286
	(frac. of Gen. Output)	0.077	0.189
Output and Efficiency			
Main Steam Flow	(lbm/hr)	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8404
OTM System Expander Generator Output	(kW)	n/a	n/a
Steam Turbine Generator Output	(kW)	209041	202770
Net Plant Output	(kW)	193034	164484
	(frac. of Case-1 Net Output)	1.00	0.85
Simplified Boiler Efficiency (HHV)[1]	(fraction)	0.8946	0.9242
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855	1810
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	7.9
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855	1818

[1] Boiler Heat Output / Qcoal (HHV)
[2] Required for GPS Desicant Regen in Cases 2-7 and ASU in Cases 2-4

		CFB Air Fired (Case 1)	2 x CFB Chem Loop (Case 7)
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	11051
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.3088
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.87
CO_2 Emissions			
CO_2 Produced	(lbm/hr)	385427	384454
CO_2 Captured	(lbm/hr)	0	383420
Fraction of CO2 Captured	(fraction)	0.00	1.00
CO_2 Emitted	(lbm/hr)	385427	1034
Specific CO_2 Emissions	(lbm/kwhr)	2.00	0.01
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.00
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.99

2.8. Case-8: Built and Operating IGCC plant without CO_2 Capture (Base Case for Comparison with Case-9)

Case-8 presents an IGCC design that is based on a built and commercially operating plant without CO_2 capture and provides the basis for comparison with Case-9. This plant is an IGCC without CO_2 capture utilizing a Texaco gasifier with a radiant syngas cooler. The overall plant design and cost basis is from a plant being operated at Tampa Electric Company (TECO) ICGG Demonstration Plant.

A brief performance summary for this plant reveals the following information. The Case-8 plant produces a net plant output of about 263 MW. The net plant heat rate and thermal efficiency are calculated to be 9,069 Btu/kWh and 37.6 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 1.81 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.8.2.

The ground rules for selection of plant configurations for Case-8 include:

- The IGCC plant is to be based on an operating plant, which has been demonstrated to be connected to the grid on a commercial scale.
- The IGCC plant will be sized with a single train GE-7FA gas turbine, producing in the range of net 260 MWe.
- The steam cycle will operate at 1,800 psig, 1,000 ^oF/ 1,000 ^oF, 3.0 in. Hga
- The coal (medium volatile bituminous) will be the same one used on all other cases in this study.
- SO_2 emissions will be equivalent to 99 percent overall removal.
- NOx emissions will be based on current GE estimates for syngas combustion.

This gasifier was selected for both Case-8 and Case-9 plant configurations. It is well suited for either producing syngas for combustion or integration with a shift reactor, as was used in Case-9, for CO_2 and hydrogen production (i.e., $CO + H_2O <==> H_2 + CO_2$).

The main criteria for this plant is to use commercially demonstrated technology for process selection. The following have been selected:

- <u>Gasifier</u>: Texaco gasifier with radiant syngas cooler, based on experience from Tampa Electric IGCC plant. Oxygen blown gasifier utilizing medium volatile bituminous coal and operating at 450 psig. Fuel gas exiting the gasifier is cooled to 1,300°F in the radiant syngas cooler.
- <u>Gas Cooling</u>: Sequential syngas coolers followed by wet scrubber.
- <u>Air Separation Plant</u>: Single train, 95 percent pure oxygen.
- <u>Gas Cleanup</u>: COS hydrolysis reactor.
- <u>Sulfur removal</u>: Proprietary amine gas removal process. H_2S sent to Claus Plant for elemental sulfur recovery.
- <u>Power Generation</u>: Single Train GE Model 7FA Gas Turbine. Heat recovery Steam Generator (HRSG) for cooling of high temperature gas turbine exhaust, 1,800 psig 1,000 ^oF/ 1,000 ^oF, 3.0 in. Hga steam cycle. Steam turbine for combined cycle power generation.

All these main components of the Demonstration Plant are depicted schematically in Figure 2.8.1, from the coal slurry feed plant to the buss bar.

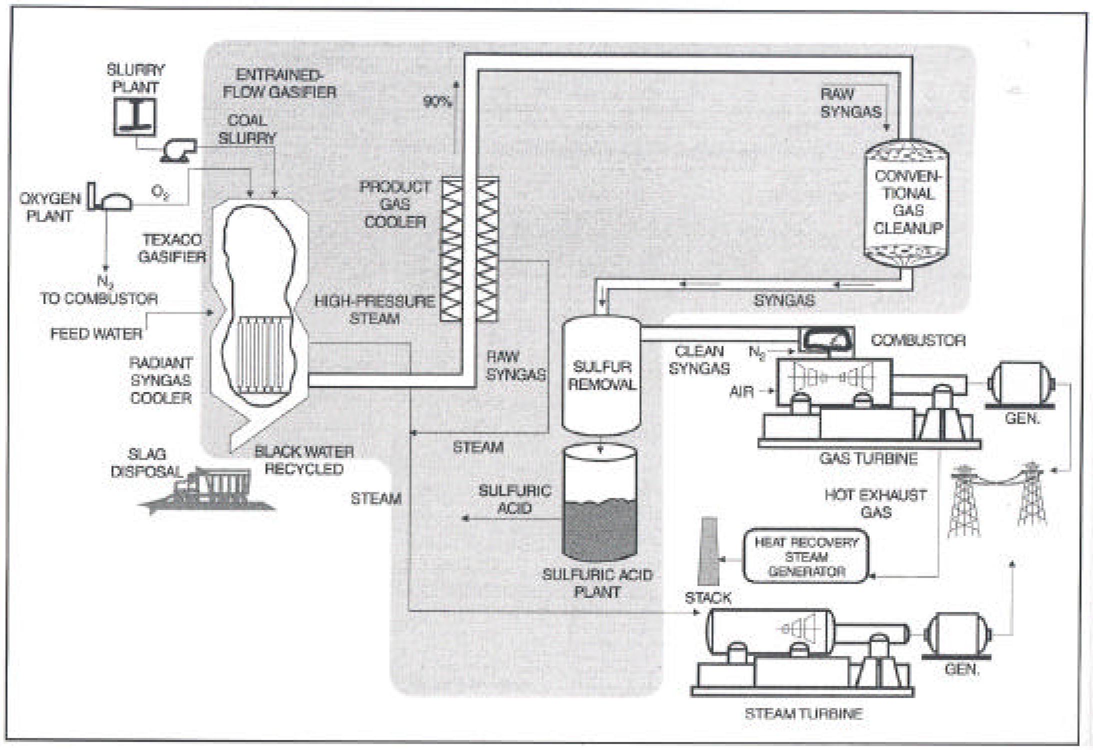


Figure 2.8. 1: Schematic of Tampa Electric Integrated Gasification Combined Cycle Demonstration Project (CCT Demonstration Program: Update 2001, USDOE, 2002)

2.8.1. Case-8 IGCC Plant Process Description

This IGCC plant design is based on the Tampa Electric IGCC Demonstration Project, which utilizes an entrained-flow, oxygen-blown Texaco gasification process. The plant configuration is based on the radiant syngas cooler gasifier mode.

The power generation technology is based on selection of a gas turbine derived from the General Electric 7FA machine. The plant is configured with two gasifier trains including processes to progressively cool and clean the gas, making it suitable for combustion in the gas turbines. The resulting plant produces a net output of 263 MWe at a net efficiency of 37.6 percent (HHV basis). Performance is based on the properties of the coal described in the plant design basis (Table 2.0.2).

The operation of the combined cycle unit in conjunction with oxygen-blown IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate (slag). A saleable by-product is produced in the form of elemental sulfur although no credit was taken for this in the economic analysis (Section 4). The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the amine-based acid gas removal (AGR) process. The AGR process removes approximately 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus unit with tail gas cleanup.

NOx emissions are limited to approximately 30 ppm by the use of steam injection in the gas turbine. The ammonia is removed from the fuel gas with process condensate prior to the low-temperature AGR process. This helps lower NOx levels as well. Particulate discharge to the atmosphere is limited to low values by the gas-washing effect of the syngas scrubber and the AGR absorber.

2.8.1.1. Block Flow Diagram

The overall plant block flow diagram for Case-8 is shown in Figure 2.8.2

The pressurized Texaco entrained-flow gasifier operating at a nominal 450 psia uses a slurry feed of water and coal, combined with oxygen, to produce a medium-Btu hot fuel gas. The fuel gas produced in the gasifier leaves at 2,450°F and enters the radiant syngas cooler, where the gas is cooled to 1,300°F. Cooling the gas to 1,300°F retains a significant fraction of the sensible heat in the gas. High-pressure saturated steam is generated in the Radiant Syngas Cooler (RSC) and is joined with the main steam supply. The fuel gas proceeds through a series of gas cleanup processes including a syngas scrubber, COS hydrolysis reactor, and an amine-based acid gas removal plant.

Particulate captured by the scrubber is routed to the “black” water system, where the solids are separated. The solids are sent off site. Regeneration gas from the AGR plant is fed to a Claus plant to produce elemental sulfur.

The clean gas exiting the AGR system is conveyed to the combustion turbines where it serves as fuel for the combustion turbine/HRSG/steam turbine power conversion system. The exhaust gas from the combustion turbine and HRSG is released to the atmosphere via a conventional stack.

This plant utilizes a combined cycle for combustion of the medium-Btu gas from the gasifier to generate electric power. A Brayton cycle using air and combustion products as working fluid is used in conjunction with a conventional subcritical steam Rankine cycle. The two cycles are coupled by generation of steam in the HRSG, by feedwater heating in the HRSG, and by heat recovery from the IGCC process (gas cooling modules).

The hot combustion gases are conveyed to the inlet of the turbine section, where they enter and expand through the turbine to produce power to drive the compressor and electric generator. The turbine exhaust gases are conveyed through a HRSG to recover the large quantities of thermal energy that remain. The HRSG exhausts to a separate stack.

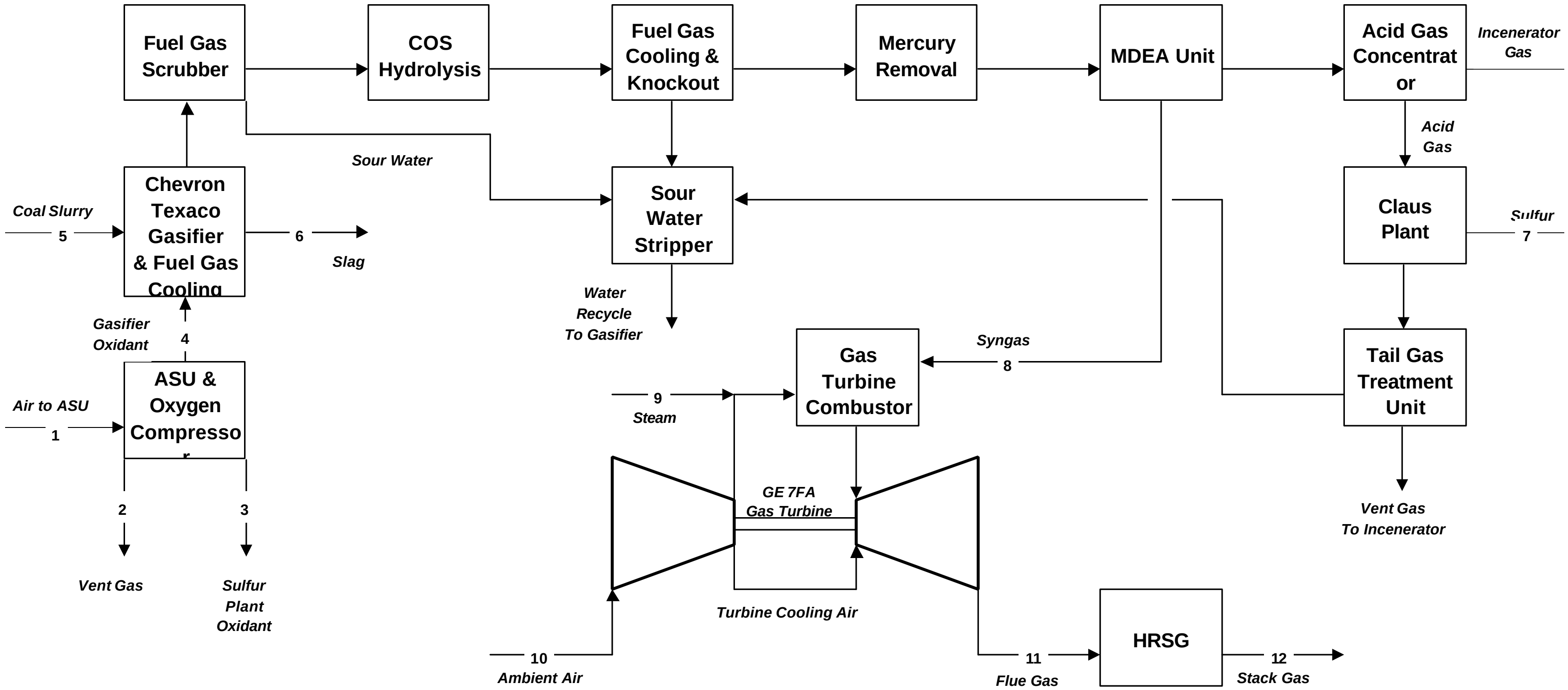


Figure 2.8. 2: Case-8 - Oxygen Blown Integrated Gasification Combined Cycle Block Flow Diagram

The steam cycle design is based on maximizing heat recovery from the combustion turbine exhaust gas, as well as efficiently utilizing steam generation opportunities in the gasifier island processes. As the turbine exhaust gas passes through the HRSG, it progressively transfers heat for superheating main steam, reheating steam, and evaporating high-pressure main steam, which is separated in a HP drum. The HRSG also evaporates and superheats intermediate-pressure steam. This steam supplements the reheat steam flow, and is also used for the integral deaerator.

The steam turbine selected to match this cycle is a two-casing, reheat, double-flow (exhaust) machine, exhausting downward to the condenser. The HP and IP turbine sections are contained in one casing, with the LP section in a second casing.

2.8.1.2. Material and Energy Balance

Table 2.8.1 provides the material and energy balance summary for the IGCC plant. It is based on the syngas fuel requirements for one General Electric 7001FA gas turbine. Ambient operating conditions for the site are indicated in the plant design basis (Section 2). The pressurized entrained-flow gasifier uses a coal/water slurry and oxygen to produce a medium heating value fuel gas. The stream numbers shown in Table 2.8.1 refer to Figure 2.8.2, a modified block flow diagram for the overall plant showing state points for all numbered streams.

Table 2.8.1: Case-8 Overall Material and Energy Balance

Mole Fraction	1	2	3	4	5	6	7	8	9	10	11	12
Ar	0.0094	0.0029	0.0348	0.0360	0.0000	0.0000	0.0000	0.0116	0.0000	0.0094	0.0094	0.0094
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4389	0.0000	0.0000	0.0000	0.0000
CO_2	0.0003	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.1367	0.0000	0.0003	0.0749	0.0749
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3998	0.0000	0.0000	0.0000	0.0000
H_2O	0.0104	0.0131	0.0000	0.0000	1.0000	0.0000	0.0000	0.0029	1.0000	0.0104	0.1484	0.1484
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.7722	0.9645	0.0152	0.0140	0.0000	0.0000	0.0000	0.0096	0.0000	0.7722	0.6479	0.6479
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2077	0.0191	0.9500	0.9500	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.1195	0.1195
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flow (lb_{mol}/hr)	27,803	22,170	82	5,549	7,470	---	---	17,123	11,556	110,743	132,240	132,240
V-L Flow (lb/hr)	802,244	620,780	2,637	178,827	134,471	---	---	340,894	208,180	3,195,400	3,744,470	3,744,470
V-L Flow (acfm)	10,953,800	7,684,020	28,371	62,870	---	---	---	390,136	340,023	43,629,700	147,617,000	71,426,300
Solids Flow (lb/hr)	0	0	0	0	215,454	52,615	5,018	0	0	0	0	0
Temperature (°F)	80	70	70	227	300	300	347	310	500	80	1,084	280
Pressure (psia)	14.7	16.4	16.4	650.0	750.0	14.7	23.6	362.5	350.0	14.7	14.8	14.7
Density (lb/ft^3)	0.073	0.081	0.093	2.844	---	---	---	0.874	0.612	0.073	0.025	0.052
Average Molecular Weight	28.85	28.00	32.21	32.23	---	---	---	19.91	18.02	28.85	28.32	28.32

The following paragraphs describe the process sections in more detail. The equipment required is described in Appendix I.

2.8.1.3. Gasifier Island

Coal Grinding and Slurry Preparation:

Coal is fed onto a conveyor by vibratory feeders located below each coal silo. The conveyor feeds the coal to an inclined conveyor that delivers the coal to the rod mill feed hopper. The feed hopper provides a surge capacity of about two hours and contains two hopper outlets. A vibrating feeder on each hopper outlet supplies the weigh feeder, which in turn feeds a rod mill. Each rod mill is sized to process 60 percent of the coal feed requirements of the gasifier. The rod mill grinds the coal and wets it with treated slurry water transferred from the slurry water tank by the slurry water pumps. The coal slurry is then discharged into the rod mill product tank. The slurry is then pumped from the rod mill product tank to the slurry storage and slurry blending tanks.

The coal grinding system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. The degree of dust suppression required will depend on local environmental regulations. All of the tanks are equipped with vertical agitators to keep the coal slurry solids suspended.

The equipment in the coal grinding and slurry preparation system is fabricated of materials appropriate for the abrasive environment present in the system. The tanks and agitators are rubber lined. The pumps are constructed with either rubber lined or hardened metals to minimize erosion. Piping is fabricated of high-density polyethylene (HDPE).

Gasifier:

This plant utilizes two gasifier trains to process a total of 2,585 tons of coal per day. The gasifiers operate at about 60 percent capacity and achieve a plant availability of 85 percent. The gasifier vessel is a refractory-lined, high-pressure combustion chamber. Coal slurry is transferred from the slurry storage tank to the gasifier with a high-pressure pump. A combination fuel injector is located at the top of the gasifier vessel through which coal slurry feedstock and oxidant (oxygen) are fed. These materials flow co-currently downward through the gasifier, where they are partially combusted to form syngas.

The coal, oxygen, and water react in the gasifier at a very high temperature to produce a syngas at 2,450ºF consisting of hydrogen, carbon monoxide, water vapor, and carbon dioxide. It also contains small amounts of hydrogen sulfide, carbonyl sulfide, methane, and nitrogen. Particles of soot and slag are also entrained in the syngas.

Radiant Syngas Cooler (RSC):

The hot gas flows downward into the RSC, where high-pressure saturated steam is produced. The syngas exits the RSC at 1,300°F and changes direction while passing over the surface of a pool of water at the bottom of the vessel. The slag drops from the gas stream into the water pool and flows from the RSC sump into a lockhopper.
The RSC in the TECO plant is about 17 feet in diameter, 100 feet long, and weighs about 900 tons. This plant utilizes an improved RSC design from MAN GHH (a large German industrial equipment supplier) that significantly reduces the size, complexity, and cost of the RSC. The following points describe the improvements that have led to the third generation of the MAN GHH RSC:

- Reduced heat exchanger surface.
- Reduced number of soot blowers.
- Reduced weight and overall dimensions.
- Simplified suspension of internals on the upper vessel head.
- Optimization of upper head area.
- Shop assembly of shell and internals.
- Easier transport and plant erection.
- Reduction to one steam circuit.

In addition to improvements in the RSC configuration, the syngas route has been modified by changing the direction of gas exiting the RSC upward at high velocity, then vertically downward through the convective syngas coolers, followed by the gas/gas heat exchanger. The gas is conducted at the bottom of the vertical run to the syngas scrubber where solid particles are separated.

Other positive effects of the revised RSC design include easily accepted thermal expansion of plant components, more compact design of the Gasification Island, and cost savings potential in equipment and steel structures. Height of the overall structure is reduced from 254 feet in the TECO plant to 213 feet in this plant.

Convective Syngas Coolers:

Two convective syngas coolers are provided to cool the syngas before it enters the gas-to-gas cooler and the scrubber. The first cooler transfers the heat to the high-pressure evaporator, which is, fed from the HRSG economizer and the second cooler transfers the heat to the high-pressure feedwater going to the RSC. The heat exchangers are both shell and tube designs.

Gas-to-Gas Heat Exchanger:

The gas-to-gas heat exchanger cools the fuel gas prior to entering the scrubber and transfers the heat to the fuel gas exiting the scrubber. This is done to provide an efficient means to heat the scrubber exit gas before it enters the COS hydrolysis unit. The gas has to be heated to approximately 410°F before entering the COS hydrolysis unit.

The gas-to-gas heat exchanger is a shell and tube design with soot blowers and cleaning devices to keep the exchanger clean and reliability operating. Utilizing the MAN GHH design, the heat exchanger is designed vertically and is constructed of corrosion-resistant materials. This is to avoid problems associated with the current TECO design.

Quench/Scrubbing:

The raw synthesis gas exiting the RSC is cooled in the series of heat exchangers before entering the scrubber. The cooled syngas at 450°F then enters the scrubber for particulate removal. The quench scrubber washes the syngas in a counter-current flow in two packed beds. After leaving the scrubber at a temperature of about 290ºF, the gas has a residual soot content of less than 1 mg/m^3 and is suitable for feeding to the COS hydrolysis reactor. The quench scrubber removes essentially all traces of entrained particles, principally unconverted carbon, slag, and metals. The bottoms from the scrubber are sent to the slag removal and handling system for further processing.

COS Hydrolysis:

Following the syngas scrubber, the fuel gas is reheated to 410°F and fed to the COS hydrolysis reactor. The COS is hydrolized with steam in the fuel gas, over a catalyst bed to H_2S, which is more easily removed by the AGR solvent. Before the raw fuel gas can be treated in the sulfur removal process, it must be cooled to 104°F. During this cooling, part of the water vapor condenses. This water, which contains some NH_3, is sent to the wastewater treatment section. No separate hydrogen cyanide (HCN) removal unit is needed due to the very low HCN concentration in the fuel gas.

Acid Gas Removal:

The promoted monodiethanolamine (MDEA) process was chosen because of its high selectivity toward H_2S and because of the low partial pressure of H_2S in the fuel gas resulting from low gas pressure, necessitating a chemical absorption process rather than a physical absorption process such as the Selexol. The AGR process utilizes an MDEA sorbent and several design features to effectively remove and recover H_2S from the fuel gas stream. The MDEA solution is relatively expensive, and measures are taken to conserve the solution during operations. As the presence of CO causes amine degradation in the form of heat stable salts, an amine reclaimer is included in the process. Also, additional water wash trays are included in the absorber tower to prevent excessive solvent loss due to vaporization.

Fuel gas enters the absorber tower at 104°F and 378 psia. Approximately 99 percent of the H_2S is removed from the fuel gas stream. The resulting clean fuel gas stream exits the absorber and is heated in a regenerative heater to 310°F.

The rich MDEA solution is pumped to a regeneration stripping tower in which the H_2S and CO_2 are stripped from the MDEA by counter-current contact with CO_2 vapors generated in a steam-heated reboiler. The regenerated H_2S stream contains 79 percent CO_2, which can affect the size and efficiency of the Claus reactor. The H_2S stream flows to an H_2S concentration absorber that separates the H_2S from the CO_2. The remaining CO_2-rich stream is incinerated with the vent gas from the tail gas treatment unit. Although not considered in this design, these concentrated streams offer an excellent opportunity for CO_2 capture and sequestration. H_2S is regenerated and sent in a concentrated stream to the Claus plant.

Sulfur Recovery System:

The sulfur recovery unit is a Claus bypass type sulfur recovery unit utilizing oxygen instead of air and with a Beavon Sulfur Removal (BSR)/Flexsorb tail gas unit. The Claus plant produces molten sulfur by reacting approximately a third of the H_2S in the feed to SO_2, then reacting the H_2S and SO_2 to sulfur and water. The combination of Claus technology and BSR/Flexsorb tail gas technology will result in an overall sulfur recovery exceeding 99 percent and a vent gas of less than 50 ppmv of SO_2. Utilizing oxygen instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The sulfur plant will produce approximately 54 long tons of sulfur per day. Feed for this case consists of acid gas from both acid gas cleanup units and a vent stream from the sour water stream in the gasifier section. Vent gas from the tail gas unit will be sent to the incinerator, and the resulting vent will meet the air quality standards of 50 ppmv of SO_2.

Sour Gas Stripper:

The sour gas stripper removes NH_3, SO_2, and other impurities from the waste stream of the scrubber. The sour gas stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from syngas coolers. Sour water from the drum flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the sulfur recovery unit. Remaining water is sent to wastewater treatment.

Air Separation Plant:

The air separation plant is designed to produce a nominal output of 2,200 tons/day of 95 percent pure O_2. The plant is designed with one production train. The dual air compressors are powered by electric motors.

In this air separation process, air is compressed to 70 psig and then cooled in a water-scrubbing spray tower. The cooled air enters a reversing heat exchanger, where it is cooled to the liquefaction point prior to entering a double column (high/low pressure) separator. Refrigeration for cooling is provided by expansion of high-pressure gas from the lower part of the high-pressure column. Approximately 50 tons/day of oxygen are fed to the Claus plant.

Flare Stack:

A self-supporting, refractory-lined, carbon steel flare stack is provided to combust and dispose of product gas during start-up, shutdown, and upset conditions. The flare stack is provided with multiple pilot burners, fueled by natural gas or propane, with pilot home monitoring instrumentation.

2.8.1.4. Power Generation System

Gas Turbine Generator:

The gas turbine generator selected for this application is the same General Electric MS 7001FA model turbine chosen for the Tampa Electric IGCC Demonstration Project. This machine is an axial flow, single spool, constant speed unit, with variable inlet guide vanes. The machine is designed for maximum reliability and efficiency with low maintenance. The turbine includes advanced bucket cooling techniques, compressor aerodynamic design and advanced alloys, enabling a higher firing temperature than the previous generation machines.

The standard production version of this machine, fired with natural gas, will develop a compressor pressure ratio of 15.2:1 and a rotor inlet temperature of almost 2,350°F. Due to the ambient site conditions, the power output from the gas turbine is 187,150 kWe.

Steam Generation:

The Radiant Syngas Cooler (RSC) is a wing wall design, which produces steam at main steam pressure, saturated conditions. This saturated steam is conveyed to the HRSG, where it is superheated. The Heat Recovery Steam Generator (HRSG) is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the gas turbine exhaust gas when firing medium-Btu gas. The HP drum produces steam at main steam pressure, while the IP drum produces steam for export to supplement the cold reheat flow. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater heating, and also provides deaeration of the condensate.

Natural circulation of steam is accomplished in the HRSG by utilizing differences in densities due to temperature differences of the steam. The natural circulation HRSG provides the most cost-effective and reliable design.

Steam Turbine Generator and Auxiliaries:

The steam turbine consists of a HP section, an IP section, and one double-flow LP section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing.

Main steam from the HRSG and Gasifier Island is combined in a header, and then passes through the stop valves and control valves and enters the turbine at 1,800 psig/1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the HRSG for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at 400 psig/1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser.

Condensate System:

The condensate system transfers condensate from the condenser hotwell to the deaerator, through the gland steam condenser, gasifier, and the low-temperature economizer section in the HRSG. The system consists of one main condenser; two 50 percent capacity, motor-driven, vertical condensate pumps; one gland steam condenser; and a low-temperature tube bundle in the HRSG. The condensate is delivered to a common discharge header through separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

Feedwater:

The function of the feedwater system is to pump the various feedwater streams from the deaerator storage tank in the HRSG to the respective steam drums. Two 50 percent capacity boiler feed pumps are provided. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

The feedwater pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing temperature. Feedwater pump suction pressure and temperature are also monitored. In addition, the suction of each boiler feed pump is equipped with a startup strainer.

Main and Reheat Steam:

The function of the main steam system is to convey main steam generated in the RSC and HRSG from the HRSG superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey steam from the HP turbine exhaust to the HRSG reheater, and to the turbine reheat stop valves.

Main steam at approximately 1,900 psig/1,000°F exits the HRSG superheater through a motor-operated stop/check valve and a motor-operated gate valve, and is routed to the HP turbine. Cold reheat steam at approximately 450 psig/645°F exits the HP turbine, flows through a motor-operated isolation gate valve, to the HRSG reheater. Hot reheat steam at approximately 420 psig/1,000°F exits the HRSG reheater through a motor-operated gate valve and is routed to the IP turbine.

Circulating Water System:

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the condenser to condense the main turbine exhaust steam. The system also supplies cooling water to the auxiliary cooling system. A mechanical-draft cooling tower removes the heat transferred from the steam to the circulating water in the condenser.

The system consists of two 50 percent capacity vertical circulating water pumps, a mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counterflow mechanical-draft cooling tower.

The condenser is a single-pass, horizontal type design with divided water boxes. There are two separate circulating water circuits in each box. One-half of the condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

2.8.1.5. Other Balance of Plant Equipment

Coal Handling System:

The function of the coal handling system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to and including the slide gate valves on the outlet of the coal storage silos.

The medium volatile bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading will be done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor. The coal is then transferred to a conveyor that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

A rubber-tired front-end loader loads the coal into two vibratory feeders located in the reclaim hopper located under the pile. The feeders transfer the coal onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0 by the first of two crushers. The coal then enters the second crusher, which reduces the coal size to 1-¼" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper, which loads the coal into one of the three silos.

Technical Requirements and Design Basis:

- Coal burn rate:

  -Maximum coal burn rate = 215,454 lbm/h = 108 tph plus 10 percent margin = 119 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)

  - Average coal burn rate = 184,000 lbm/h = 92 tph (based on MCR rate multiplied by 85 percent capacity factor)
- Coal delivered to the plant by unit trains:

  - Two unit trains per week at maximum burn rate. One and one-half unit trains per week at average burn rate

  - Each unit train shall have 10,000 tons (100-ton cars) capacity

  - Unloading rate = 9 cars/hour (maximum)

  - Total unloading time per unit train = 11 hours (minimum)

  - Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)

  - Reclaim rate = 600 tph
- Storage piles with liners, run-off collection, and treatment systems:

  - Active storage = 8,000 tons (72 hours at maximum burn rate)

  - Dead storage = 67,000 tons (30 days at average burn rate)

Table 2.8. 2: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	94
Active Storage, tons	8,000
Dead Storage, tons	80,000

Slag Handling:

The plant includes two slag-handling systems. One system handles the slag generated at the base of the RSC, and the second handles the slag removed from the syngas in the syngas scrubber.

The RSC coarse slag handling system conveys, stores, and disposes of slag removed from the gasification process. Slag exits through the slag tap into a water bath in the bottom of the RSC vessel. A slag crusher receives slag from the water bath and grinds the material into pea-sized fragments. A slag/water slurry that contains between 5 and 10 percent solids flows out of the bottom of the RSC through a pressure letdown valve into a lined ground level tank. The components listed above, up to the pressure letdown valve, are within the gasifier pressure boundary and at high pressure.

The cooled, dewatered slag is removed from the bottom of the slag-handling tank by drag chain conveyors. The slag mixture is discharged to a vibrating screen where the fine slag is removed. The larger screened slag is stored in 3 storage bins. The bins are sized for a nominal hold-up capacity of approximately 72 hours of full-load operation. At periodic intervals, a convoy of slag hauling trucks will transit the unloading station underneath the hopper and remove a quantity of slag for disposal. Approximately 24 truckloads per day are required to remove the total quantity of slag produced by the plant operating at nominal rated power. (The selected coal produces a relatively large amount of slag). The fine slag separated by the vibrating screen is transported to the coal slurry storage tank for reburn in the gasifier.

The fine slag handling system removes the slag removed by the scrubber. The system consists of a clarifier and rotary drum filter. The slag/water mixture flows by gravity to the clarifier where the solids settle to the bottom. The solids are removed by pumps and transported to the drum filter. The thickened slag/water mixture is further dewatered, and the solids are discharged to a belt conveyor. The conveyor transports the slag to an awaiting truck or dumpster for transport to the waste disposal area.

Raw Water, Fire Protection, and Cycle Makeup Water Systems:

The raw water system supplies 1,300 gpm of cooling tower makeup, 600 gpm for the cycle makeup, and 15 gpm for service water use and potable water requirements. The pumps will be installed on an intake structure located on the river in close proximity to the plant.

The fire protection system provides water under pressure to the fire hydrants, hose stations, and fixed water suppression system within the buildings and structures. The system consists of pumps, underground and aboveground supply piping, distribution piping, hydrants, hose stations, spray systems, and deluge spray systems. One motor-operated booster pump is supplied on the intake structure of the cooling tower with a diesel engine pump installed on the intake structure located on the river.
The cycle makeup water system provides high quality demineralized water for makeup to the HRSG cycle and for injection steam to the combustion turbine for control of NOx emissions and auxiliary boiler.

The cycle makeup system will consist of two 100 percent trains, each with a full-capacity activated carbon filter, primary cation exchanger, primary anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. The equipment will be skid-mounted and include a control panel, and associated piping, valves, and instrumentation.

Accessory Electric Plant:

The accessory electric plant consists of switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, and wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed control system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are manually implemented, with operator selection of modular automation routines available. The exception to this, and an important facet of the control system for gasification, is the critical controller system, which is a part of the license package from Texaco and is a dedicated and distinct hardware segment of the DCS.

This critical controller system is used to control the gasification process. The partial oxidation of the fuel feed and oxygen feed streams to form a syngas product is a stoichiometric, temperature- and pressure-dependent reaction. Texaco pilot plant experience has yielded dynamic compensations and ratio controls for the feed components that are continuously updated with combustion product quality feedback data. The critical controller utilizes a redundant microprocessor executing calculations and dynamic controls at 100- to 200-millisecond intervals. The enhanced execution speeds as well as evolved predictive controls allow the critical controller to mitigate process upsets and maintain the reactor operation within a stable set of operating parameters.

Layout Arrangement:

The development of the plant site to incorporate structures required for this technology is based on the assumption of a flat site. The IGCC gasifiers and related structures are arranged in a cluster, with the coal and slurry preparation facilities adjacent to the southeast, as shown in the conceptual general arrangement shown in Appendix II.

The gasifier and its associated process blocks are located west of the coal storage pile. The gas turbine and its ancillary equipment are sited northwest of the gasifier island, in a turbine building. The HRSG and stack are east of the gas turbine, with the steam turbine and its generator in a separate building to the north. Service and administration buildings are located at the west side of the steam turbine building.

The cooling tower heat sink for the steam turbine is located to the east of the steam turbine building. The air separation plant is further to the southwest, with storage tanks for liquid O_2 located near the gasifier and its related process blocks. Sulfur recovery, slag recovery, and wastewater treatment areas are located east and north of the gasifier.

The arrangement described above provides good alignment and positioning for major interfaces; relatively short steam, feedwater, and fuel gas pipelines; and allows good access for vehicular traffic. Transmission line access from the gas turbine and steam turbine step-up transformer to the switchyard is also maintained at short distances.

The air and gas path is developed in a short and direct manner, with ambient air entering an inlet filter/silencer located south of the gas turbine. The clean, hot, medium-Btu gas is conveyed to the turbine combustor for mixing with the air that remains on-board the machine. Turbine exhaust is ducted directly through the HRSG and then the 213-foot (65-meter) stack. The height of the stack is established by application of a good engineering practice rule from 40 CFR 51.00.

Access and construction lay-down spaces are freely available on the periphery of the plant.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building

- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

2.8.2. Case-8 Overall Plant Performance and Emissions

The Case-8 IGCC produces a net output of 263 MWe at a net efficiency of 37.6 percent on HHV basis. Overall performance for the entire plant is summarized in Table 2.8.3, which includes auxiliary power requirements.

Table 2.8.3: Case-8 IGCC Plant Performance

POWER SUMMARY	(Gross Power at Generator Terminals, kWe)
Gas Turbine Power	187,150
Steam Turbine Power	113,717
Total	300,867
Coal Handling	280
Slurry Pumps	200
Slag Handling	100
Air Separation Unit Auxiliaries	20,680
Oxygen Compressor	9,150
Scrubber Pumps	50
Incinerator Blower	60
Wastewater Treatment Auxiliaries	20
LP Oxygen Blower	30
HP Boiler Feedwater Pump	2,040
IP Boiler Feedwater Pump	90
Condensate Pump	120
Circulating Water Pump	980
Cooling Tower Fans	580
Gas Turbine Auxiliaries	400
Steam Turbine Auxiliaries	200
Miscellaneous Balance-of-Plant	1,000
Claus Plant Auxiliaries	100
MDEA Unit Auxiliaries	1,010
Transformer Losses	690
Total	37,780
Net Auxiliary Load	37,780
Net Plant Power	263,087
Net Plant Efficiency (HHV)	37.6%
Net Plant Heat Rate (HHV)	9,069
Coal Feed Flowrate	215,454
Thermal Input (HHV)	699,073

The operation of the combined cycle unit in conjunction with oxygen-blown Texaco IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate. A saleable by-product is produced in the form of elemental sulfur although no credit was taken for this product in the economic analysis (Section 4). A summary of the plant emissions is presented in Table 2.8.4.

The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the amine-based AGR process. The AGR process removes approximately 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur.

Since the selected site has no restrictions on water consumption, NOx emissions in the flue gas are limited by the use of steam injection to approximately 30 ppm based on 15 percent oxygen, which is equivalent to 90 ppm @ 3 percent O_2. Steam injection was selected instead of nitrogen injection, which is used at the TECO plant. Adequate water is available at the reference site, and costs associated with compression of nitrogen from the air separation unit (ASU) are avoided. Selective catalytic reduction (SCR) or selective non-catalytic reduction (SNCR) can reduce emissions further, but are not applied to the subject plant.

Particulate discharge to the atmosphere is limited to extremely low values by the use of the syngas scrubber and the gas washing effect of the AGR absorber. CO_2 emissions are equal to those of other coal-burning facilities on an intensive basis (lb/10^6 Btu), since a similar fuel is used. However, total CO_2 emissions are lower for a plant with this capacity due to the relatively high thermal efficiency.

Table 2.8.4 Overall Plant Emissions

	lb/10^6 Btu	lbm/hour	lbm/MWh
SO_2	0.042	97.6	0.38
NO_x	0.023	53.4	0.21
Particulate	< 0.002	< 4.5	< 0.018
CO_2	200.0	477,093	1,810

2.9. Case-9: Built and Operating IGCC with Shift Reactor and CO_2 Capture

The Case-9 IGCC plant design is based on the Tampa Electric IGCC Demonstration Project, which utilizes an entrained-flow, oxygen-blown Texaco gasification process. The plant configuration is based on the radiant syngas cooler gasifier design. The plant utilizes two gasifier trains including processes to progressively cool and clean the gas, making it suitable for combustion in the GE 7FA gas turbine.

The CO_2 removal design is based on parameters associated with other studies, which have been conducted by Parsons to denote the effect of design change on efficiency and equipment requirements. The reference report for this study is "Evaluation of Innovative Fossil Fuel Power Plants with CO_2 Removal," EPRI Interim Report (Holt, 2000). Cases 3A and 3B of the EPRI report are indicative of the changes, which would occur as the design is modified for CO_2 removal.

This plant produces a net output of 231 MWe at a net efficiency of 29.8 percent on an HHV basis. Performance is based on the properties of the selected coal, described in the plant design basis (Table 2.0.1). The plant captures and recovers 90 percent of the CO_2, which would be produced from the coal feed. Carbon dioxide emissions are about 0.23 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.9.2.

The operation of the combined cycle unit in conjunction with oxygen-blown IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate (slag). A saleable by-product is produced in the form of elemental sulfur although no credit was taken for this product in the economic analysis (Section 4). The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the amine-based acid gas removal (AGR) process. The AGR process removes approximately 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus unit with tail gas cleanup.

NOx emissions are limited to approximately 30 ppm at 15 percent O_2, which is equivalent to 90 ppm @3 percent O_2, by the use of steam injection. The ammonia is removed with process condensate prior to the low-temperature AGR process. This helps lower NOx levels as well. Particulate discharge to the atmosphere is limited to low values by the gas-washing effect of the syngas scrubber and the AGR absorber.

2.9.1. Case-9 IGCC Plant Process Description

The Case-9 IGCC plant design is based on the Tampa Electric IGCC Demonstration Project, which utilizes an entrained-flow, oxygen-blown Texaco gasification process. CO_2 capture is added to this process. The plant configuration is based on the radiant cooler gasifier mode.

The power generation technology is based on selection of a gas turbine derived from the General Electric 7FA machine operating on hydrogen-rich syngas with much of the CO_2 removed. The plant is configured with the gasifier including processes to progressively cool and clean the gas, shift the CO to hydrogen and CO_2, making it suitable for CO_2 capture and combustion in the gas turbines. The resulting plant produces a net output of 231 MWe at a net efficiency of 29.8 percent on an HHV basis. Performance is based on the properties of the coal, described in the plant design basis.

The operation of the combined cycle unit in conjunction with oxygen-blown IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate (slag). A saleable by-product is produced in the form of elemental sulfur although no credit was taken for this product in the economic analysis (Section 4). The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the amine-based acid gas removal (AGR) process. The AGR process removes approximately 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus unit with tail gas cleanup.

NOx emissions are limited to approximately 30 ppm by the use of steam injection. The ammonia is removed with process condensate prior to the low-temperature AGR process. This helps lower NOx levels as well. Particulate discharge to the atmosphere is limited to low values by the gas-washing effect of the syngas scrubber and the AGR absorber.

2.9.1.1. Block Flow Diagram:

The overall plant block flow diagram for Case-9 is shown in Figure 2.9.1. The pressurized Texaco entrained-flow gasifier is identical to that of Case 8. The gas goes through a series of gas cleanup processes including a syngas scrubber, but instead of a COS hydrolysis reactor, the syngas goes through a sour CO-shift reactor. A two-stage amine-based acid gas removal (AGR) plant separately removes H_2S and CO_2.

The clean hydrogen-rich gas exiting the AGR system is conveyed to the combustion turbines where it serves as fuel for the combustion turbine/HRSG/steam turbine power conversion system. The exhaust gas from the combustion turbine and HRSG is released to the atmosphere via a conventional stack.

This plant also utilizes a combined cycle for combustion of the medium-Btu gas from the gasifier to generate electric power. A Brayton cycle using air and combustion products as working fluid is used in conjunction with a conventional subcritical steam Rankine cycle. The two cycles are coupled by generation of steam in the HRSG, by feedwater heating in the HRSG, and by heat recovery from the IGCC process (gas cooling modules).

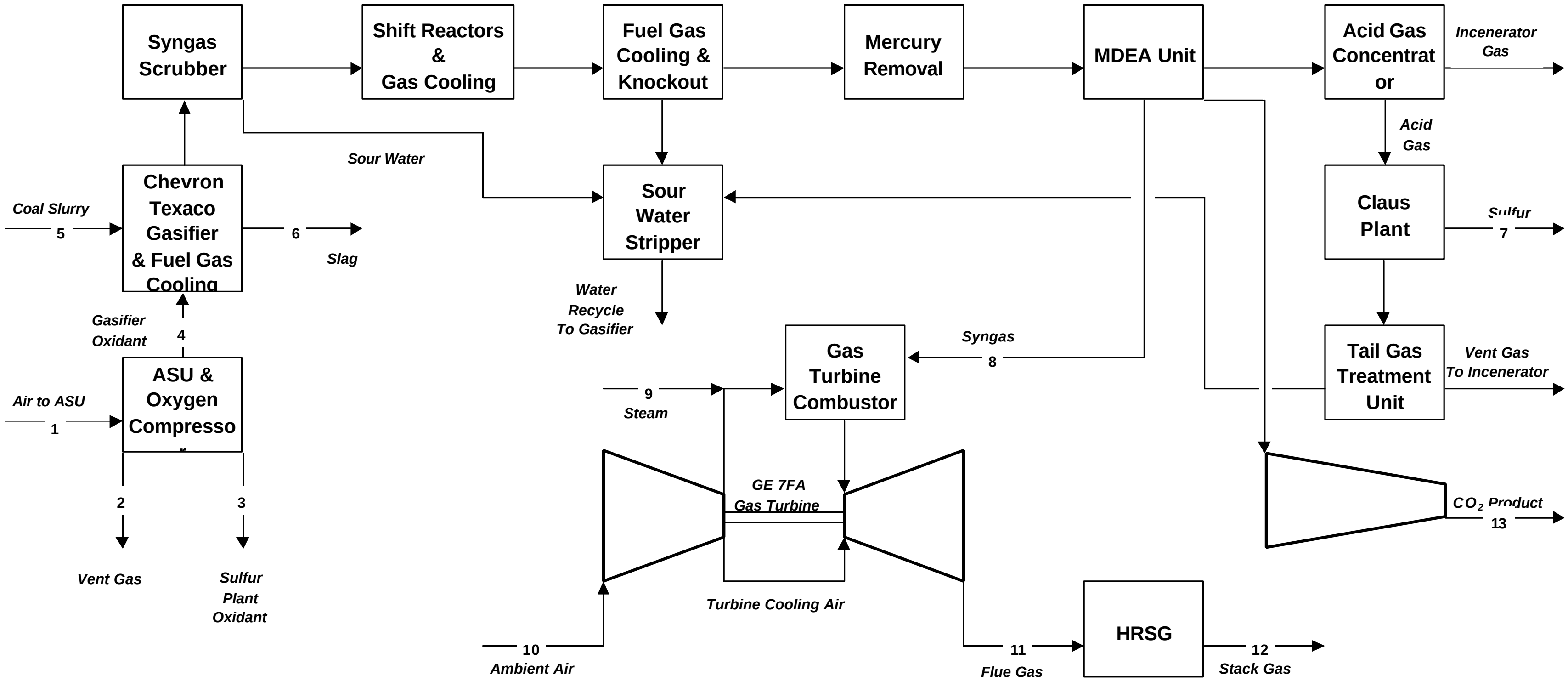


Figure 2.9. 1: Case-9 – IGCC with CO_2 Capture Block Flow Diagram

The pressurized Texaco entrained-flow gasifier operating at a nominal 450 psia uses a slurry feed of water and coal, combined with oxygen, to produce a medium-Btu hot fuel gas. The fuel gas produced in the gasifier leaves at 2450°F and enters the radiant syngas cooler, where the gas is cooled to 1300°F.

Cooling the gas to 1300°F retains a significant fraction of the sensible heat in the gas. High-pressure saturated steam is generated in the RSC and is joined with the main steam supply.

The gas goes through a series of gas cleanup processes including a syngas scrubber, water gas shift reactor, and an amine-based acid gas removal (AGR) plant. Particulate captured by the scrubber is routed to the "black" water system, where the solids are separated. The solids are sent off site. Regeneration gas from the AGR plant is fed to a Claus plant to produce elemental sulfur.

The hot combustion gases are conveyed to the inlet of the turbine section, where they enter and expand through the turbine to produce power to drive the compressor and electric generator. The turbine exhaust gases are conveyed through a HRSG to recover the large quantities of thermal energy that remain. The HRSG exhausts to a separate stack.

The steam cycle is based on maximizing heat recovery from the gas turbine exhaust gas, as well as utilizing steam generation opportunities in the gasifier process. As the turbine exhaust gas passes through the HRSG, it progressively transfers heat for reheating steam (cold reheat to hot reheat), superheating main steam, and generating main steam in an HP drum. The HRSG also generates and superheats steam from an IP drum (as reheat, and for use in the integral deaerator), and heats feedwater.

The steam turbine selected to match this cycle is a two-casing, reheat, double-flow (exhaust) machine, exhausting downward to the condenser. The HP and IP turbine sections are contained in one casing, with the LP section in a second casing.

2.9.1.2. Material and Energy Balance

Table 2.9.1 provides the material and energy balance for the IGCC plant is based on the syngas fuel requirements for one General Electric 7001FA gas turbine. Ambient operating conditions for the site are indicated in the plant design basis.

The pressurized entrained-flow gasifier uses a coal/water slurry and oxygen to produce a medium heating value fuel gas. The stream numbers shown in Table 2.9.1 refer to Figure 2.9.1, a modified block diagram for the overall plant showing state point for all numbered streams.

Table 2.9. 1: Case-9 Overall Material and Energy Balance

Mole Fraction	1	2	3	4	5	6	7	8	9	10	11	12
Ar	0.0094	0.0029	0.0348	0.0360	0.0000	0.0000	0.0000	0.0116	0.0000	0.0094	0.0094	0.0094
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4389	0.0000	0.0000	0.0000	0.0000
CO_2	0.0003	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.1367	0.0000	0.0003	0.0749	0.0749
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3998	0.0000	0.0000	0.0000	0.0000
H_2O	0.0104	0.0131	0.0000	0.0000	1.0000	0.0000	0.0000	0.0029	1.0000	0.0104	0.1484	0.1484
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.7722	0.9645	0.0152	0.0140	0.0000	0.0000	0.0000	0.0096	0.0000	0.7722	0.6479	0.6479
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2077	0.0191	0.9500	0.9500	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.1195	0.1195
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flow (lb_{mol}/hr)	27,803	22,170	82	5,549	7,470	---	---	17,123	11,556	110,743	132,240	132,240
V-L Flow (lb/hr)	802,244	620,780	2,637	178,827	134,471	---	---	340,894	208,180	3,195,400	3,744,470	3,744,470
V-L Flow (acfm)	10,953,800	7,684,020	28,371	62,870	---	---	---	390,136	340,023	43,629,700	147,617,000	71,426,300
Solids Flow (lb/hr)	0	0	0	0	215,454	52,615	5,018	0	0	0	0	0
Temperature (°F)	80	70	70	227	300	300	347	310	500	80	1,084	280
Pressure (psia)	14.7	16.4	16.4	650.0	750.0	14.7	23.6	362.5	350.0	14.7	14.8	14.7
Density (lb/ft^3)	0.073	0.081	0.093	2.844	---	---	---	0.874	0.612	0.073	0.025	0.052
Average Molecular Weight	28.85	28.00	32.21	32.23	---	---	---	19.91	18.02	28.85	28.32	28.32

The following paragraphs describe the process sections in more detail. The equipment required is described in Appendix I.

2.9.1.3. Gasifier Island

Coal Grinding and Slurry Preparation:

Coal is fed onto a conveyor by vibratory feeders located below each coal silo. The conveyor feeds the coal to an inclined conveyor that delivers the coal to the rod mill feed hopper. The feed hopper provides a surge capacity of about two hours and contains two hopper outlets. A vibrating feeder on each hopper outlet supplies the weigh feeder, which in turn feeds a rod mill. Each rod mill is sized to process 60 percent of the coal feed requirements of the gasifier. The rod mill grinds the coal and wets it with treated slurry water transferred from the slurry water tank by the slurry water pumps. The coal slurry is then discharged into the rod mill product tank. The slurry is then pumped from the rod mill product tank to the slurry storage and slurry blending tanks.

The coal grinding system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. The degree of dust suppression required will depend on local environmental regulations. All of the tanks are equipped with vertical agitators to keep the coal slurry solids suspended.

The equipment in the coal grinding and slurry preparation system is fabricated of materials appropriate for the abrasive environment present in the system. The tanks and agitators are rubber lined. The pumps are either rubber lined or hardened metal to minimize erosion. Piping is fabricated of high-density polyethylene (HDPE).

Gasifier:

This plant utilizes two gasifier trains to process a total of 2,585 tons of coal per day. The gasifiers operate at about 60 percent capacity and achieve a plant availability of 85 percent. The gasifier vessel is a refractory-lined, high-pressure combustion chamber. Coal slurry is transferred from the slurry storage tank to the gasifier with a high-pressure pump. At the top of the gasifier vessel is located a combination fuel injector through which coal slurry feedstock and oxidant (oxygen) are fed. These materials flow co-currently downward through the gasifier, where they are partially combusted to form syngas.

The coal, oxygen, and water react in the gasifier at a very high temperature to produce a syngas at 2450ºF consisting of hydrogen, carbon monoxide, water vapor, and carbon dioxide. It also contains small amounts of hydrogen sulfide, carbonyl sulfide, methane, and nitrogen. Particles of soot and slag are also entrained in the syngas.

Radiant Syngas Cooler (RSC):

The hot gas flows downward into the RSC, where high-pressure saturated steam is produced. The syngas exits the RSC at 1300°F and changes direction while passing over the surface of a pool of water at the bottom of the vessel. The slag drops from the gas stream into the water pool and flows from the RSC sump into a lock-hopper.
The RSC in the TECO plant is about 17 feet in diameter, 100 feet long, and weighs about 900 tons. This plant utilizes an improved RSC design from MAN GHH that significantly reduces the size, complexity, and cost of the RSC. The following points describe the improvements that have led to the third generation of the MAN GHH RSC:

- Reduced heat exchanger surface.
- Reduced number of soot blowers.

- Reduced weight and overall dimensions.
- Simplified suspension of internals on the upper vessel head.
- Optimization of upper head area.
- Shop assembly of shell and internals.
- Easier transport and plant erection.
- Reduction to one steam circuit.

In addition to improvements in the RSC configuration, the syngas route has been modified by changing the direction of gas exiting the RSC upward at high velocity, then vertically downward through the convective syngas coolers, followed by the gas/gas heat exchanger. The gas is conducted at the bottom of the vertical run to the syngas scrubber where solid particles are separated.

Other positive effects of the revised RSC design include easily accepted thermal expansion of plant components, more compact design of the Gasification Island, and savings potential in equipment and steel structures. Height of the overall structure is reduced from 254 feet in the TECO plant to 213 feet in this plant.

Convective Syngas Coolers:

Two convective syngas coolers are provided to cool the syngas before it enters the gas-to-gas cooler and the scrubber. The first cooler transfers the heat to the high-pressure steam from the HRSG economizer, and the second cooler transfers the heat to the high-pressure feedwater going to the RSC. The heat exchangers are tube and shell construction.

Gas-to-Gas Heat Exchanger:

The gas-to-gas heat exchanger cools the gas prior to entering the scrubber and transfers the heat to the gas exiting the scrubber. This is done to provide an efficient means to heat the scrubber exit gas before it enters the water gas shift unit.

The gas-to-gas heat exchanger is a shell and tube construction with soot blowers and cleaning devices to keep the exchanger clean and reliability operating. Utilizing the MAN GHH design, the heat exchanger is designed vertically and is constructed of corrosion-resistant materials. This is to avoid problems associated with the current TECO design.

Quench/Scrubbing:

The raw synthesis gas exiting the RSC is cooled in the series of heat exchangers before entering the scrubber. The cooled syngas at 450°F then enters the scrubber for particulate removal. The quench scrubber washes the syngas in a counter-current flow in two packed beds. After leaving the scrubber at a temperature of about 290ºF, the gas has a residual soot content of less than 1 mg/m^3 and is suitable for feeding to the CO-shift reactor. The quench scrubber removes essentially all traces of entrained particles, principally unconverted carbon, slag, and metals.

Water Gas Shift:

Hot, particulate-free syngas from the scrubber and gas to gas heat exchanger is fed to the CO-shift reactor. A set of high-temperature shift reactors is used to shift the bulk of the CO in the fuel gas to CO_2. Heat exchange between reaction stages helps maintain a moderate reaction temperature. A two-stage shift was utilized in order to maximize CO conversion while maintaining reasonable reactor volumes.

The shifted raw gas temperature exiting the second shift converter is approximately 553°F. This stream is cooled to 370°F in a low-temperature economizer. The fuel gas stream is cooled to 103°F in a series of low-temperature economizers and then routed to the Acid Gas Removal unit. Fuel gas condensate is recovered and routed to a sour water drum.

Acid Gas Removal:

The promoted monodiethanolamine (MDEA) process was chosen because of its high selectivity toward H_2S and because of the relatively low partial pressure of H_2S in the fuel gas favoring a chemical absorption process rather than a physical absorption process such as the Selexol. The AGR process utilizes an MDEA sorbent and several design features to effectively remove and recover H_2S from the fuel gas stream. The MDEA solution is relatively expensive, and measures are taken to conserve the solution during operations. As the presence of CO causes amine degradation in the form of heat stable salts, an amine reclaimer is included in the process. Also, additional water wash trays are included in the absorber tower to prevent excessive solvent loss due to vaporization.

Cool, dry, and particulate-free synthesis gas enters the first absorber unit at approximately 378 psia and 105°F. In this absorber, H_2S is preferentially removed from the fuel gas stream. This is achieved by "loading" the lean amine solvent with CO_2. The solvent, saturated with CO_2, preferentially removes H_2S. The rich solution leaving the bottom of the absorber is regenerated in a stripper through the indirect application of thermal energy via condensing low-pressure steam in a reboiler. Sweet fuel gas flowing from the first absorber is cooled and routed to the second absorber unit. In this absorber, the fuel gas is contacted with "unloaded" lean solvent. The solvent removes approximately 97 percent of the CO_2 in the fuel gas stream. A CO_2 balance is maintained by hydraulically expanding the CO_2-saturated rich solution and then flashing CO_2 vapor off the liquid at reduced pressure. Sweet fuel gas off the second absorber is warmed and humidified in the fuel gas saturator, reheated and sent to the burner of the combustion turbine.

The rich MDEA solution is pumped to a regeneration-stripping tower in which the H_2S and CO_2 are stripped from the MDEA by counter-current contact with CO_2 vapors generated in a steam-heated reboiler. The regenerated H_2S stream contains 79 percent CO_2, which can affect the size and efficiency of the Claus reactor. The H_2S stream flows to an H_2S concentration absorber that separates the H_2S from the CO_2. The remaining CO_2-rich stream is incinerated with the vent gas from the tail gas treatment unit.

Sulfur Recovery System:

The sulfur recovery unit is a Claus bypass type sulfur recovery unit utilizing oxygen instead of air and with a Beavon Sulfur Removal (BSR)/Flexsorb tail gas unit. The Claus plant produces molten sulfur by reacting approximately a third of the H_2S in the feed to SO_2, then reacting the H_2S and SO_2 to sulfur and water. The combination of Claus technology and BSR/Flexsorb tail gas technology will result in an overall sulfur recovery exceeding 99 percent and a vent gas of less than 50 ppmv of SO_2. Utilizing oxygen instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The sulfur plant will produce approximately 60 long tons of sulfur per day. Feed for this case consists of acid gas from both acid gas cleanup units and a vent stream from the sour water stream in the gasifier section. Vent gas from the tail gas unit will be sent to the incinerator, and the resulting vent will meet the air quality standards of 50 ppmv of SO_2.

CO_2 Compression and Drying:

CO_2 is flashed from the rich solution at two pressures. The bulk of the CO_2 is flashed off at approximately 50 psia, while the remainder is flashed off at atmospheric pressure. The second low-pressure CO_2 stream is "boosted" to 50 psia and then combined with the first

CO_2 stream. The combined flow is then compressed in a multiple-stage, intercooled compressor to supercritical conditions at 2,000 psia. During compression, the CO_2 stream is dehydrated with triethylene glycol. The virtually moisture-free supercritical CO_2 steam is then ready for pipeline transportation.

Sour Gas Stripper:

The sour gas stripper removes NH_3, SO_2, and other impurities from the waste stream of the scrubber. The sour gas stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from syngas coolers. Sour water from the drum flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the sulfur recovery unit. Remaining water is sent to wastewater treatment.

Air Separation Plant:

The air separation plant is designed to produce a nominal output of 2,500 tons/day of 95 percent pure O_2. The plant is designed with one production train. The dual air compressors are powered by electric motors.

In this air separation process, air is compressed to 70 psig and then cooled in a water-scrubbing spray tower. The cooled air enters a reversing heat exchanger, where it is cooled to the liquefaction point prior to entering a double column (high/low pressure) separator. Refrigeration for cooling is provided by expansion of high-pressure gas from the lower part of the high-pressure column. Approximately 50 tons/day of oxygen are fed to the Claus plant.

Flare Stack:

A self-supporting, refractory-lined, carbon steel flare stack is provided to combust and dispose of product gas during start-up, shutdown, and upset conditions. The flare stack is provided with multiple pilot burners, fueled by natural gas or propane, with pilot home monitoring instrumentation.

2.9.1.4. Power Generation System

Gas Turbine Generator:

The gas turbine generator selected for this application is the same General Electric MS 7001FA model turbine chosen for the Tampa Electric IGCC Demonstration Project. This machine is an axial flow, single spool, constant speed unit, with variable inlet guide vanes. The machine is designed for maximum reliability and efficiency with low maintenance. The turbine includes advanced bucket cooling techniques, compressor aerodynamic design and advanced alloys, enabling a higher firing temperature than the previous generation machines. The standard production version of this machine, fired with natural gas, will develop a compressor pressure ratio of 15.2:1 and a rotor inlet temperature of almost 2350°F. Due to the ambient site conditions, the power output from the gas turbine is 187,150 kWe.

Steam Generation:

The Radiant Syngas Cooler (RSC) is a wing wall design, which produces steam at main steam pressure, saturated conditions. This steam is conveyed to the HRSG, where it is superheated. The Heat Recovery Steam Generator (HRSG) is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the gas turbine exhaust gas when firing medium-Btu gas. The HP drum produces steam at main steam pressure, while the IP drum produces steam for export to the cold reheat. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater heating, and also provides deaeration of the condensate.

Natural circulation of steam is accomplished in the HRSG by utilizing differences in densities due to temperature differences of the steam. The natural circulation HRSG provides the most cost-effective and reliable design.

Steam Turbine Generator and Auxiliaries:

The steam turbine consists of an HP section, an IP section, and one double-flow LP section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing.

Main steam from the HRSG and Gasifier Island is combined in a header, and then passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the HRSG for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at 400 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser.

Condensate System:

The condensate system transfers condensate from the condenser hotwell to the deaerator, through the gland steam condenser, gasifier, and the low-temperature economizer section in the HRSG. The system consists of one main condenser; two 50 percent capacity, motor-driven, vertical condensate pumps; one gland steam condenser; and a low-temperature tube bundle in the HRSG. The Condensate is delivered to a common discharge header through separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

Feedwater:

The function of the feedwater system is to pump the various feedwater streams from the deaerator storage tank in the HRSG to the respective steam drums. Two 50 percent capacity boiler feed pumps are provided. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

The feedwater pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing temperature. Feedwater pump suction pressure and temperature are also monitored. In addition, the suction of each boiler feed pump is equipped with a startup strainer.

Main and Reheat Steam:

The function of the main steam system is to convey main steam generated in the RSC and HRSG from the HRSG superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey steam from the HP turbine exhaust to the HRSG reheater, and to the turbine reheat stop valves.

Main steam at approximately 1,900 psig / 1,000°F exits the HRSG superheater through a motor-operated stop/check valve and a motor-operated gate valve, and is routed to the HP turbine. Cold reheat steam at approximately 450 psig / 645°F exits the HP turbine, flows through a motor-operated isolation gate valve, to the HRSG reheater. Hot reheat steam at approximately 420 psig / 1,000°F exits the HRSG reheater through a motor-operated gate valve and is routed to the IP turbines.

Circulating Water System:

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the condenser to condense the main turbine exhaust steam. The system also supplies cooling water to the auxiliary cooling system. The heat transferred from the steam to the circulating water in the condenser is removed by a mechanical draft cooling tower.

The system consists of two 50 percent capacity vertical circulating water pumps, a mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counterflow mechanical draft cooling tower.

The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of the condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

2.9.1.5. Other Balance of Plant Equipment

Coal Handling System :

The function of the coal handling system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to and including the slide gate valves on the outlet of the coal storage silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading will be done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor. The coal is then transferred to a conveyor that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

A rubber-tired front-end loader loads the coal into two vibratory feeders located in the reclaim hopper located under the pile. The feeders transfer the coal onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0 by the first of two crushers. The coal then enters the second crusher, which reduces the coal size to 1-¼" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper, which loads the coal into one of the three silos.

Technical Requirements and Design Basis:

Coal burn rate:

- Maximum coal burn rate = 238,694 lbm/h = 120 tph plus 10 percent margin = 132 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
- Average coal burn rate = 204,000 lbm/h = 102 tph (based on MCR rate multiplied by 85 percent capacity factor)

Coal delivered to the plant by unit trains:

- Two unit trains per week at maximum burn rate. One and one-half unit trains per week at average burn rate
- Each unit train shall have 10,000 tons (100-ton cars) capacity
- Unloading rate = 9 cars/hour (maximum)
- Total unloading time per unit train = 11 hours (minimum)
- Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
- Reclaim rate = 600 tph

Storage piles with liners, run-off collection, and treatment systems:

- Active storage = 8,800 tons (72 hours at maximum burn rate)
- Dead storage = 75,000 tons (30 days at average burn rate)

Table 2.9. 2: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	120
Active Storage, tons	8,800
Dead Storage, tons	75,000

Slag Handling:

The plant includes two slag-handling systems: one system handles the slag generated at the base of the RSC, and the second handles the slag removed from the syngas in the syngas scrubber.

The RSC coarse slag handling system conveys, stores, and disposes of slag removed from the gasification process. Slag exits through the slag tap into a water bath in the bottom of the RSC vessel. A slag crusher receives slag from the water bath and grinds the material into pea-sized fragments. A slag/water slurry that is between 5 and 10 percent solids flows out of the bottom of the RSC through a pressure letdown valve into a lined ground level tank. The components listed above, up to the pressure letdown valve, are within the gasifier pressure boundary and at high pressure.

The cooled, dewatered slag is removed by drag chain conveyors from the bottom of the slag handling tank. The slag mixture is discharged to a vibrating screen where the fine slag is removed. The larger screened slag is stored in 3 storage bins. The bins are sized for a nominal hold-up capacity of approximately 72 hours of full-load operation. At periodic intervals, a convoy of slag hauling trucks will transit the unloading station underneath the hopper and remove a quantity of slag for disposal. Approximately 24 truckloads per day are required to remove the total quantity of slag produced by the plant operating at nominal rated power. (The selected coal produces a relatively large amount of slag). The fine slag separated by the vibrating screen is transported to the coal slurry storage tank for reburn in the gasifier.

The fine slag handling system removes the slag removed by the scrubber. The system consists of a clarifier and rotary drum filter. The slag/water mixture flows by gravity to the clarifier where the solids settle to the bottom. The solids are removed by pumps and transported to the drum filter. The thickened slag/water mixture is further dewatered, and the solids are discharged to a belt conveyor. The conveyor transports the slag to an awaiting truck or dumpster for transport to the waste disposal area.

Raw Water, Fire Protection, and Cycle Makeup Water Systems:

The raw water system supplies 1,300 gpm of cooling tower makeup, 600 gpm for the cycle makeup, and 15 gpm for service water use and potable water requirements. The pumps will be installed on an intake structure located on the river in close proximity to the plant.

The fire protection system provides water under pressure to the fire hydrants, hose stations, and fixed water suppression system within the buildings and structures. The system consists of pumps, underground and aboveground supply piping, distribution piping, hydrants, hose stations, spray systems, and deluge spray systems. One motor-operated booster pump is supplied on the intake structure of the cooling tower with a diesel engine pump installed on the intake structure located on the river.

The cycle makeup water system provides high quality demineralized water for makeup to the HRSG cycle and for injection steam to the combustion turbine for control of NOx emissions and auxiliary boiler.

The cycle makeup system will consist of two 100 percent trains, each with a full-capacity activated carbon filter, primary cation exchanger, primary anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. The equipment will be skid-mounted and include a control panel, and associated piping, valves, and instrumentation.

Accessory Electric Plant:

The accessory electric plant consists of switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, and wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed control system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are manually implemented, with operator selection of modular automation routines available. The exception to this, and an important facet of the control system for gasification, is the critical controller system, which is a part of the license package from Texaco and is a dedicated and distinct hardware segment of the DCS.

This critical controller system is used to control the gasification process. The partial oxidation of the fuel feed and oxygen feed streams to form a syngas product is a stoichiometric, temperature- and pressure-dependent reaction. Texaco pilot plant experience has yielded dynamic compensations and ratio controls for the feed components that are continuously updated with combustion product quality feedback data. The critical controller utilizes a redundant microprocessor executing calculations and dynamic controls at 100- to 200-millisecond intervals. The enhanced execution speeds as well as evolved predictive controls allow the critical controller to mitigate process upsets and maintain the reactor operation within a stable set of operating parameters.

Layout Arrangement:

The development of the reference plant site to incorporate structures required for this technology is based on the assumption of a flat site. The IGCC gasifiers and related structures are arranged in a cluster, with the coal and slurry preparation facilities adjacent to the southeast, as shown in the conceptual general arrangement in Appendix II.

The gasifier and its associated process blocks are located west of the coal storage pile. The gas turbine and its ancillary equipment are sited northwest of the gasifier island, in a turbine building. The HRSG and stack are east of the gas turbine, with the steam turbine and its generator in a separate building to the north. Service and administration buildings are located at the west side of the steam turbine building.

The cooling tower heat sink for the steam turbine is located to the east of the steam turbine building. The air separation plant is further to the southwest, with storage tanks for liquid O_2 located near the gasifier and its related process blocks. Sulfur recovery, slag recovery, and wastewater treatment areas are located east and north of the gasifier.

The arrangement described above provides good alignment and positioning for major interfaces; relatively short steam, feedwater, and fuel gas pipelines; and allows good access for vehicular traffic. Transmission line access from the gas turbine and steam turbine step-up transformer to the switchyard is also maintained at short distances.

The air and gas path is developed in a short and direct manner, with ambient air entering an inlet filter/silencer located south of the gas turbine. The clean, hot, medium-Btu gas is conveyed to the turbine combustor for mixing with the air that remains on-board the machine. Turbine exhaust is ducted directly through the HRSG and then the 213-foot (65-meter) stack. The height of the stack is established by application of a good engineering practice rule from 40 CFR 51.00.

Access and construction laydown spaces are freely available on the periphery of the plant.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building

- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

2.9.2. Case-9 Overall Plant Performance and CO_2 Emissions

The reference report for this study is the Evaluation of Innovative Fossil Fuel Power Plants with CO_2 Removal, EPRI Interim Report, (Holt, 2000). Referring to the report, Cases 3A and 3B are indicative of the changes, which would occur as the design is modified for CO_2 removal. Using the relative changes from the report, Table 2.9.3 indicates the changes in performance, which are also applied to Cases 8 and 9.

The Case-9 IGCC produces a net output of 231 MWe at a net efficiency of 29.8 percent on HHV basis. Overall performance for the entire plant is summarized in Table 2.9.3, which includes auxiliary power requirements.

Table 2.9.3: Case-9 Overall Plant Performance: Impact of CO_2 Recovery

	Case 8 IGCC without CO_2 Removal	Case 9 IGCC with CO_2 Removal
Gas Turbine Power	187,150 kW	187,150 kW
Steam Turbine Power	113,717 kW	112,318 kW
Gross Plant Power	300,867 kW	299,468 kW
Key Auxiliary Power Items		
ASU Auxiliaries	20,680 kW	22,911 kW
Oxygen Compressor	9,150 kW	10,137 kW
CO_2 Compression	N/A	27,105 kW
Balance of Auxiliaries	7,950 kW	8,800 kW
Total Auxiliary Power	(37,780 kW)	(68,953 kW)
Net Plant Power	**263,087 kW**	**230,515 kW**
Coal Feed	215,454 lbm/hr	238,694 lbm/hr
Thermal Input	699,073 kWe	774,479 kWe
Plant Efficiency, HHV	**37.6%**	**29.8%**

The operation of the combined cycle unit in conjunction with oxygen-blown Texaco IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate. A saleable by-product is produced in the form of elemental sulfur although no credit was taken for this product in the economic analysis (Section 4). A summary of the plant emissions is presented in Table 2.9.4.

The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the amine-based AGR process. The AGR process removes approximately 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur.

Since the selected site has no restrictions on water consumption, NOx emissions are limited by the use of steam injection to approximately 30 ppm based on 15 percent oxygen in the flue gas. Steam injection was selected instead of nitrogen injection, which is used at the TECO plant. Adequate water is available at the reference site, and costs associated with compression of nitrogen from the air separation unit (ASU) are avoided. Selective catalytic reduction (SCR) or selective non-catalytic reduction (SNCR) can reduce emissions further, but are not applied to the subject plant.

Particulate discharge to the atmosphere is limited to extremely low values by the use of the syngas scrubber and the gas washing effect of the AGR absorber.

The low CO_2 emissions are indicative of the removal of 90 percent of the CO_2 produced from the coal feed.

The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the amine-based AGR process. The AGR process removes approximately 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur.

Since the selected site has no restrictions on water consumption, NOx emissions in the flue gas are limited by the use of steam injection to approximately 30 ppm based on 15 percent oxygen, which is equivalent to 90 ppm @ 3 percent O_2. Steam injection was selected instead of nitrogen injection, which is used at the TECO plant. Adequate water is available at the reference site, and costs associated with compression of nitrogen from the air separation unit (ASU) are avoided. Selective catalytic reduction (SCR) or selective non-catalytic reduction (SNCR) can reduce emissions further, but are not applied to the subject plant.

Particulate discharge to the atmosphere is limited to extremely low values by the use of the syngas scrubber and the gas washing effect of the AGR absorber.

CO_2 emissions are equal to those of other coal-burning facilities on an intensive basis (lb/10^6 Btu), since a similar fuel is used. However, total CO_2 emissions are lower for a plant with this capacity due to the relatively high thermal efficiency.

Table 2.9. 4: Overall Plant Emissions

	lb/10^6 Btu	lbm/hour	lbm/MWh
SO_2	0.042	108.1	0.47
NO_x	0.023	59.2	0.26
Particulate	< 0.002	< 5.1	< 0.022
CO_2	20.5	52,749	227

2.10. Case-10: Commercially Offered Future IGCC without CO_2 Capture (Base Case for Comparison with Case-11)

Case-10 represents a commercially offered but not yet built and operated future IGCC plant without CO_2 capture and provides the basis for comparison with Case-11. This plant is an IGCC without CO_2 capture utilizing Texaco high pressure, oxygen-blown, entrained flow, quench gasification technology with syngas expander. The gasification system design is based on the Eastman Chemical Company Acetic Anhydride plant.

A brief performance summary for this plant reveals the following information. The Case-10 plant produces a net plant output of about 235 MW. The net plant heat rate and thermal efficiency are calculated to be 9,884 Btu/kWh and 34.5 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 1.98 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.10.2.

The ground rules for selection of plant configurations for Case- 10 include:

- The IGCC plant is based on technology that has been offered on a commercial scale.
- The IGCC plant is sized for a single train GE 7FA gas turbine, producing in the range of 250 MWe net.
- The steam cycle operating parameters are 1,800 psig / 1,000 ^{o}F / 1,000 ^{o}F, 3.0 in. Hga.
- The coal (medium volatile bituminous) is the same one used in all other cases in this study.
- SO_2 emissions are equivalent to 99 percent overall removal.
- NOx emissions are based on current GE estimates for syngas combustion.

The high pressure Texaco quench gasifier was selected as the gasifier for both Case-10 and Case-11 plant configurations. It is currently being operated at the Eastman Chemical Company Acetic Anhydride Plant in Kingsport, Tennessee. The gasifier is well suited for either producing syngas for combustion or integration with a shift reactor for CO_2 and hydrogen production ($CO + H_2O \Rightarrow H_2 + CO_2$) as was done in Case-11.

The following plant process selections have the points of reference for commercial scale demonstration shown in parenthesis. The main criterion for this plant is to use commercially offered technology for process selection. The following have been selected:

- Gasifier: Texaco Quench, based on experience from Eastman Chemical Acetic Anhydride Plant. Oxygen Blown, Bituminous Coal, Operates at 950 psig. Gas exiting quench zone at saturation temperature.
- Gas Cooling: Sequential syngas coolers followed by wet scrubber. (Eastman Chemical Plant)
- Air Separation Plant: Single Train, 95 percent pure Oxygen (Wabash and TECO IGCC Plant) High pressure oxygen compressor (Eastman Plant).
- Gas Cleanup: COS hydrolysis reactor. (Wabash and TECO IGCC Plant)
- Sulfur removal: Selexol acid gas removal process (Refinery Syngas). Regenerated H_2S sent to sulfuric acid plant. (TECO IGCC Plant)
- Power Generation: Sweet syngas expander. (GE Rotoflow) Single Train GE Model 7FA Gas Turbine. Heat recovery Steam Generator (HRSG) on high temperature turbine exhaust, 1,800 psig steam cycle. Steam turbine for combined cycle generation. (Wabash and TECO IGCC Plant)

2.10.1. Case-10 IGCC Plant Process Description

This IGCC plant design is based on the Chevron-Texaco Power and Gasification Corporation technology, which utilizes a pressurized entrained-flow, oxygen-blown gasification process. The plant configuration is based on the quench gasifier design option operating at approximately 950 psig.

The power generation technology is based on selection of a gas turbine derived from the General Electric 7FA machine. The plant is configured with one gasifier including processes to progressively cool and clean the gas, making it suitable for combustion in the gas turbine. The resulting plant produces a net output of 235 MWe at a net efficiency of 34.5 percent on an HHV basis

The operation of the combined cycle unit in conjunction with oxygen-blown IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate (slag). A salable by-product is produced in the form of elemental sulfur although no credit was taken for this product in the economic analysis (Section 4). The low level of SO_2 in the plant emissions is achieved by the capture of the sulfur in the gas by the Selexol acid gas removal (AGR) process. The AGR process is designed to remove 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus unit with tail gas being recycled to the AGR system.

NOx emissions are limited to 9 ppm in the flue gas (normalized to 15 percent O_2) – or 27 ppm @ 3 percent O_2 -- by the use of syngas humidification and nitrogen dilution. The ammonia is removed with process condensate prior to the low-temperature AGR process. This helps lower NOx levels as well. Selective catalytic reduction (SCR) can reduce emissions further to 5 ppmv level (at 15 percent O_2) or 15 ppmv @ 3 percent O_2 or lower if required.

Particulate discharge to the atmosphere is limited to low values by the gas-washing effect of the syngas scrubber and the AGR absorber.

2.10.1.1. Block Flow Diagram

The overall plant block flow diagram for Case-10 is shown in Figure 2.10.1. This diagram shows only the major process units and streams.

The pressurized Chevron-Texaco entrained-flow gasifier uses a slurry feed of water and coal, combined with oxygen, to produce a medium-Btu hot fuel gas (about 203 Btu/scf). The gas goes through a series of gas coolers and cleanup processes including a COS hydrolysis reactor, a carbon bed mercury removal system and an acid gas removal plant. The particulate captured by the scrubber is routed to the "black" water system, where the solids are separated. The solids are sent off site. Regeneration gas from the AGR plant is fed to a Claus plant to produce elemental sulfur.

The clean gas exiting the AGR system is conveyed to the combustion turbines where it serves as fuel for the combustion turbine/HRSG/steam turbine power conversion system. The exhaust gas from the combustion turbine and HRSG is released to the atmosphere via a conventional stack.

This plant utilizes a combined cycle for combustion of the medium-Btu gas from the gasifier to generate electric power. A Brayton cycle using air and combustion products as working fluid is used in conjunction with a conventional subcritical steam Rankine cycle. The two cycles are coupled by generation of steam in the HRSG, by feedwater heating in the HRSG, and by heat recovery from the IGCC process (gas cooling modules).

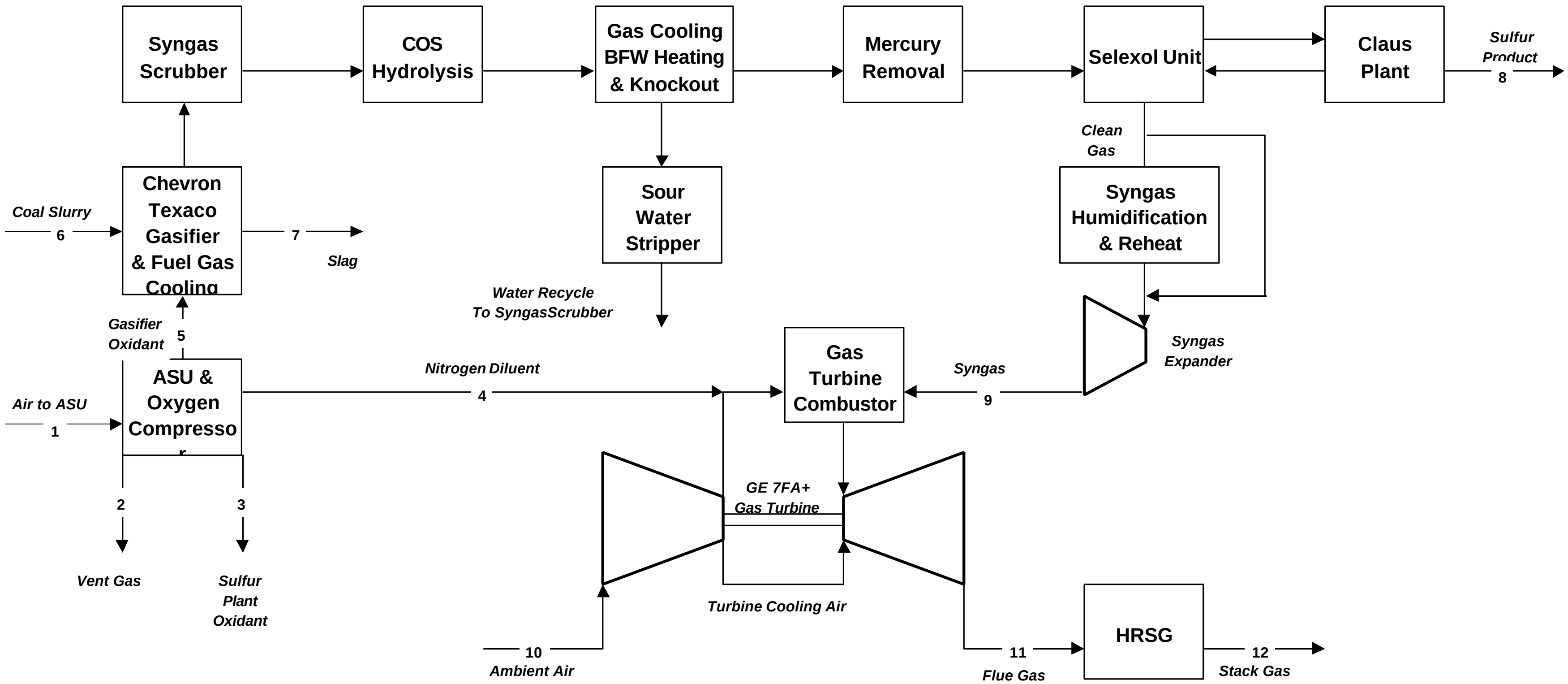


Figure 2.10.1: Case-10 Overall Plant Block Flow Diagram

2.10.1.2. Material and Energy Balance

Table 2.10.1 provides the material and energy balance for the IGCC plant. It is based on General Electric's estimate for the syngas fuel requirements for one 7FA gas turbine. Ambient operating conditions are indicated in the plant design basis. The pressurized entrained-flow gasifier uses a coal/water slurry and oxygen to produce a medium heating value fuel gas. The stream numbers shown in Table 2.10.1 refer to Figure 2.10.1, a modified block flow diagram for the overall plant showing state points for all numbered streams.

Table 2.10. 1: Case-10 Overall Plant Material and Energy Balance

Mole Fraction	1	2	3	4	5	6	7	8	9	10	11	12
Ar	0.0094	0.0045	0.0322	0.0018	0.0360	0.0000	0.0000	0.0000	0.0095	0.0094	0.0091	0.0091
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0013	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3491	0.0000	0.0000	0.0000
CO_2	0.0003	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1561	0.0003	0.0785	0.0785
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3152	0.0000	0.0000	0.0000
H_2O	0.0104	0.0300	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.1336	0.0104	0.0783	0.0783
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.7722	0.9645	0.0178	0.9982	0.0140	0.0000	0.0000	0.0000	0.0351	0.7722	0.7164	0.7164
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2077	0.0000	0.9500	0.0000	0.9500	0.0000	0.0000	0.0000	0.0000	0.2077	0.1176	0.1176
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000
V-L Flow (lb_{mol}/hr)	25,371	8,798	130	11,028	5,408	11,778	0	19	20,987	110,746	135,798	135,798
V-L Flow (lb/hr)	732,071	244,425	4,170	309,167	174,309	156,653	0	4,904	442,348	3,195,490	3,947,010	3,947,010
V-L Flow (acfm)	166,594	49,659	425	4,971	706	397	0	0	10,822	727,182	2,488,150	1,222,468
Solids Flow (lb/hr)	0	0	0	0	0	210,011	51,084	0	0	0	0	0
Temperature (°F)	80	64	90	284	354	73	300	350	535	80	1,061	280
Pressure (psia)	14.7	16.4	30.0	295.0	1,115.0	1,050.0	14.7	23.6	345.0	14.7	14.8	14.7
Density (lb/ft^3)	0.073	0.082	0.164	1.037	4.117	---	---	329.774	0.681	0.073	0.026	0.054
Average Molecular Weight	28.85	27.78	32.18	28.03	32.23	---	---	256.53	21.08	28.85	29.07	29.07

The following paragraphs describe the process sections in more detail. The equipment required is described in Appendix I.

2.10.1.3. Gasifier Island

Coal Grinding and Slurry Preparation:

Coal is fed onto a conveyor by vibratory feeders located below each coal silo. The conveyor feeds the coal to an inclined conveyor that delivers the coal to the rod mill feed hopper. The feed hopper provides a surge capacity of about two hours and contains two hopper outlets. A vibrating feeder on each hopper outlet supplies the weigh feeder, which in turn feeds a rod mill. Each rod mill is sized to process 60 percent of the coal feed requirements of the gasifier. The rod mill grinds the coal and wets it with treated slurry water transferred from the slurry water tank by the slurry water pumps. The coal slurry is then discharged into the rod mill product tank. The slurry is then pumped from the rod mill product tank to the slurry storage and slurry blending tanks.

The coal grinding system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. The degree of dust suppression required will depend on local environmental regulations. All of the tanks are equipped with vertical agitators to keep the coal slurry solids suspended.

The equipment in the coal grinding and slurry preparation system is fabricated of materials appropriate for the abrasive environment present in the system. The tanks and agitators are rubber lined. The pumps are either rubber lined or hardened metal to minimize erosion. Piping is fabricated of high-density polyethylene (HDPE).

Gasifier:

This plant utilizes one gasification train to process a total of 2,520 tons of coal per day. Due to the properties of the coal such as high ash and low as-received heating value, additional coal mass flow is needed to reach the syngas requirements. This results in a need for additional oxygen and resultant auxiliary power requirements. This is reflected in the higher than expected heat rate. The gasifier operates at maximum capacity. The slurry feed pump takes suction from the slurry run tank, and the discharge is sent to the feed injector of the Chevron-Texaco gasifier. Oxygen from the ASU is vented during preparation for startup and is sent to the feed injector during normal operation. The air separation plant supplies 2,092 tons of 95 percent pure oxygen per day to the gasifier.

The gasifier vessel is a refractory-lined, high-pressure combustion chamber. The coal slurry is transferred from the slurry storage tank to the gasifier with a high-pressure pump. At the top of the gasifier vessel is located a combination fuel injector through which coal slurry feedstock and oxidant (oxygen) are fed. The coal slurry and the oxygen feeds react in the gasifier at about 965 psia at a high temperature to produce syngas.

The syngas consists primarily of hydrogen and carbon monoxide, with lesser amounts of water vapor and carbon dioxide, and small amounts of hydrogen sulfide, carbonyl sulfide, methane, argon, and nitrogen. The heat in the gasifier liquefies the coal ash. Hot syngas and molten solids from the reactor flow downward into a water-filled quench chamber where the syngas is cooled and the slag solidifies. Raw syngas then flows to the syngas scrubber for removal of any remaining entrained solids. The solids collect in the water sump at the bottom of the gasifier and are removed periodically, using a lock hopper system. Fine material, which does not settle as easily, is removed in the gasification blowdown and goes to the vacuum flash drum by way of the syngas scrubber.

Syngas Scrubbing:

The recycled condensate is sprayed into the raw syngas, and the water/syngas mixture enters the syngas scrubber and is directed downward by a dip tube into a water sump at the bottom of the syngas scrubber. Most of the solids are separated from the syngas at the bottom of the dip tube as the syngas goes upwards through the water. From the overhead of the syngas scrubber, the syngas enters the low-temperature gas cooling section for further cooling.

The water removed from the syngas scrubber contains all the solids that were not removed in the quench gasifier water sump. In order to limit the amount of solids recycled to the quench chamber, a continuous blowdown stream is removed from the bottom of the syngas scrubber.

Slag Handling System:

The slag handling system removes solids from the gasification process equipment. These solids consist of a small amount of unconverted carbon and essentially all of the ash contained in the feed coal. These solids are in the form of glass, which fully encapsulates any metals.

Low-Temperature Gas Cooling:

Hot, particulate-free syngas from the scrubber is partially cooled in the low-pressure (LP) steam generator by producing LP steam and then is reheated to 400°F and fed to the COS hydrolysis reactor. After hydrolysis the raw syngas is further cooled in the LP steam generator to 103°F. The condensate in the syngas is removed in the steam generator knockout (KO) drum. A pump transfers the process condensate to the syngas scrubber and the quench gasifier exit.

During this cooling, part of the water vapor condenses. This water, which contains some NH_3, is sent to the wastewater treatment section. No separate hydrogen cyanide (HCN) removal unit is needed due to the very low HCN concentration in the fuel gas.

COS Hydrolysis:

Cooled syngas is fed to the COS hydrolysis reactor at 400°F. The COS is hydrolized with steam, over a catalyst bed to H_2S, which is more easily removed by the AGR solvent. The hydrolysis reactor converts essentially all of the COS to H_2S. After hydrolysis the raw syngas is further cooled in the LP steam generator to 103°F and is sent to the Mercury Removal section.

Mercury Removal:

Mercury removal is based on packed beds of sulfur-impregnated carbon similar to what has been used at the Eastman Chemical gasification plant. A single bed of sulfur-impregnated carbon with approximately a 20-second superficial gas residence time achieves 95 percent reduction of mercury in addition to removal of other volatile heavy metals such as arsenic.

Acid Gas Removal:

The design basis utilizes a single train Selexol process to remove sulfur with minimal CO_2 capture. The Selexol process treats the stream of synthesis gas to reduce the level of total sulfur (H_2S and COS) to no more than 15 ppm prior to it being sent to the combustion turbine, while maximizing the CO_2 slip. A recycle stream of acid gas from the Sulfur Recovery Unit (SRU) is also treated.

Untreated gas is sent to the Selexol absorber, where it contacts cooled regenerated solvent, which enters at the top of the tower. In the absorber, H_2S, COS, CO_2 and other gases such as hydrogen, are transferred from the gas phase to the liquid phase. The treated gas exits the absorber and is sent to fuel gas saturation and the expander.

The solvent streams from the absorber and reabsorber are termed rich solvent, and are combined and sent to the Lean/Rich Exchanger. In the Lean/Rich Exchanger the temperature of the rich solvent is increased by heat exchange with the lean solvent. The rich solvent is then sent to the H_2S concentrator, where a portion of the CO_2, CO, H_2 and other gases are stripped from the solvent. Nitrogen from the ASU is used as the stripping medium. The temperature of the overhead stream from the H_2S concentrator is reduced in the Stripped Gas Cooler, and it is then sent to the reabsorber, where H_2S, COS and a portion of the other gases are transferred to the liquid phase. The stream from the reabsorber is sent to the gas turbine.

The partially regenerated solvent exits the H_2S Concentrator and is sent to the stripper, where the solvent is regenerated. Tail gas from the SRU is recycled back to the Acid Gas Removal Unit and enters with the feed to the reabsorber.

Syngas Expander:

After sulfur removal, the sweet fuel gas is saturated with condensate, reheated and depressurized through an expander from 825 psia to 345 psia, which is near the pressure required by the gas turbine. The expander generates 6,650 kW_e.

Sulfur Recovery System:

The sulfur recovery unit is a Claus bypass type sulfur recovery unit utilizing oxygen instead of air. The Claus plant produces molten sulfur by converting approximately a third of the H_2S in the feed to SO_2, then converting the remaining H_2S and SO_2 to elemental sulfur and water. The combination of Claus technology and tail gas recycle to the Selexol results in an overall sulfur recovery of 99 percent. Utilizing oxygen instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The sulfur plant will produce approximately 53 tons of elemental sulfur per day.

Air Separation Plant:

The air separation plant is designed to produce a nominal output of 2,150 tons/day of 95 percent pure O_2. Most of the oxygen is used in the gasifier. A small portion, approximately 50 tons/day, is used in the Claus plant. The plant is designed with one production train. The air compressor is powered by an electric motor. Approximately 3,500 tons/day of nitrogen are also recovered, compressed and used as dilution in the gas turbine combustor.

In this air separation process, air is compressed to 85 psia and then cooled in a water-scrubbing spray tower. The cooled air enters a reversing heat exchanger, where it is cooled to the liquefaction point prior to entering a double column (high/low pressure) separator.

Flare Stack:

A self-supporting, refractory-lined, carbon steel flare stack is provided to combust and dispose of product gas during startup, shutdown, and upset conditions. The flare stack is provided with multiple pilot burners, fueled by natural gas or propane, with pilot home monitoring instrumentation.

2.10.1.4. Power Generation System

Gas Turbine Generator:

The gas turbine generator selected for this application is the same General Electric MS 7001FA model turbine chosen for the Tampa Electric IGCC Demonstration Project. There are over 140 GE 7FA and GE 9FA units ordered or in operation. This machine is an axial flow, single spool, constant speed unit, with variable inlet guide vanes. The machine is designed for maximum reliability and efficiency with low maintenance. The turbine includes advanced bucket cooling techniques, compressor aerodynamic design and advanced alloys, enabling a higher firing temperature than the previous generation machines. The standard production version of this machine, fired with natural gas, will develop a compressor pressure ratio of 15.2:1 and a rotor inlet temperature of almost 2350°F.

In this service, with syngas from an IGCC plant and the 80°F site temperature, the machine requires some modifications to the burner and turbine nozzles in order to properly combust the medium-Btu gas and expand the combustion products in the turbine section of the machine. A reduction in rotor inlet temperature of about 50°F results, relative to a production model 7FA machine firing natural gas. This temperature reduction is necessary to not exceed design basis gas path temperatures throughout the expander. If the first-stage rotor inlet temperature were maintained at the design value, the gas path temperatures downstream of the inlet to the first (HP) turbine stage may increase, relative to natural gas-fired temperatures, due to gas property changes. The gas turbine net power output amounts to 187,150 kWe.

The modifications to the machine include some redesign of the original can-annular combustors. A second modification involves increasing the nozzle areas of the turbine to accommodate the mass and volume flow of medium-Btu fuel gas combustion products, which are increased relative to those produced when firing natural gas.

Other modifications include rearranging the various auxiliary skids that support the machine to accommodate the spatial requirements of the plant general arrangement. The generator is a standard hydrogen-cooled machine with static exciter.

Steam Generation:

The heat recovery steam generator (HRSG) is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the gas turbine exhaust gas when firing medium-Btu gas. The HP drum produces steam at main steam pressure, while the IP drum produces steam for export to the cold reheat.

The HRSG drum pressures are nominally 1, 800 and 420 psia for the HP and IP turbine sections, respectively. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater heating, and also provides deaeration of the condensate.

Steam Turbine Generator and Auxiliaries:

The steam turbine consists of an HP section, an IP section, and one double-flow LP section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing. The LP turbine has a last-stage bucket length of 30 inches.

Main steam from the HRSG and gasifier island is combined in a header, and then passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the HRSG for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at 400 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser.

The generator is a hydrogen-cooled synchronous type, generating power at 23 kV. A static, transformer type exciter is provided. The generator is cooled with a hydrogen gas recirculation system using fans mounted on the generator rotor shaft. The heat absorbed by the gas is removed as it passes over finned tube gas coolers mounted in the stator frame. Gas is prevented from escaping at the rotor shafts by a closed-loop oil seal system. The oil seal system consists of a storage tank, pumps, filters, and pressure controls, all skid-mounted.

The steam turbine generator is controlled by a triple-redundant, microprocessor-based electro-hydraulic control system. The system provides digital control of the unit in accordance with programmed control algorithms, color CRT operator interfacing, and datalink interfaces to the balance-of-plant distributed control system (DCS), and incorporates on-line repair capability.

Condensate System:

The condensate system transfers condensate from the condenser hotwell to the deaerator, through the gland steam condenser, gasifier, and the low-temperature economizer section in the HRSG. The system consists of one main condenser; two 50 percent capacity, motor-driven, vertical condensate pumps; one gland steam condenser; and a low-temperature tube bundle in the HRSG. Condensate is delivered to a common discharge header through separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

Feedwater System:

The function of the feedwater system is to pump the various feedwater streams from the deaerator storage tank in the HRSG to the respective steam drums. Two 50 percent capacity boiler feed pumps are provided. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

Main and Reheat Steam:

The function of the main steam system is to convey main steam generated HRSG from the HRSG superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey steam from the HP turbine exhaust to the HRSG reheater, and to the turbine reheat stop valves.

Main steam at approximately 1,900 psig / 1,000°F exits the HRSG superheater through a motor-operated stop/check valve and a motor-operated gate valve, and is routed to the HP turbine. Cold reheat steam at approximately 450 psig / 645°F exits the HP turbine, flows through a motor-operated isolation gate valve, to the HRSG reheater. Hot reheat steam at approximately 420 psig / 1,000°F exits the HRSG reheater through a motor-operated gate valve and is routed to the IP turbines.

Circulating Water System:

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the condenser to condense the main turbine exhaust steam. The system also supplies cooling water to the auxiliary cooling system. The heat transferred from the steam to the circulating water in the condenser is removed by a mechanical draft cooling tower.

The system consists of two 50 percent capacity vertical circulating water pumps, a mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counterflow mechanical draft cooling tower.

The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of the condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

In addition to the condenser, additional cooling is required for the air separation unit. This amounts to an additional 175 MM-Btu/hr.

2.10.1.5. Other Balance of Plant Equipment

Coal Handling System:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading will be done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron, and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the two silos.

Technical Requirements and Design Basis:

- Coal burn rate:
  -Maximum coal burn rate = 210,011 lbm/h = 110 tph plus 10 percent margin = 120 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)

-Average coal burn rate = 186,000 lbm/h = 93 tph (based on MCR rate multiplied by an 85 percent capacity factor)

- Coal delivered to the plant by unit trains:
  -One and one-half unit trains per week at maximum burn rate
  -One unit train per week at average burn rate
  -Each unit train shall have 10,000 tons (100-ton cars) capacity
  -Unloading rate = 9 cars/hour (maximum)
  -Total unloading time per unit train = 11 hours (minimum)
  -Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
  -Reclaim rate = 300 tph
- Storage piles with liners, run-off collection, and treatment systems:
  -Active storage = 8,000 tons (72 hours at maximum burn rate)
  -Dead storage = 80,000 tons (30 days at average burn rate)

Table 2.10. 2: Case-10 Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	89
Active Storage, tons	8,000
Dead Storage, tons	80,000

Slag Handling:

The plant includes two slag handling systems: one system handles the slag generated at the base of the quench gasifier, and the second handles the slag removed from the syngas in the scrubber.

The coarse slag handling system conveys, stores, and disposes of slag removed from the gasification process. Slag exits through the slag tap into a water bath in the bottom of the quench vessel. A slag crusher receives slag from the water bath and grinds the material into pea-sized fragments. A slag/water slurry that is between 5 and 10 percent solids flows out of the bottom of the quench vessel through a pressure letdown valve into a lined ground level tank. The components listed above, up to the pressure letdown valve, are within the gasifier pressure boundary and at high pressure.

Drag chain conveyors from the bottom of the slag-handling tank remove the cooled, dewatered slag. The slag mixture is discharged to a vibrating screen where the fine slag is removed. The larger screened slag is stored in a storage bin. The bin is sized for a nominal holdup capacity of approximately 72 hours of full-load operation. At periodic intervals, a convoy of slag hauling trucks will transit the unloading station underneath the hopper and remove a quantity of slag for disposal. Approximately 16 truckloads per day are required to remove the total quantity of slag produced by the plant operating at nominal rated power.

The fine slag handling system removes the slag removed by the scrubber. The system consists of a clarifier and rotary drum filter. The slag/water mixture flows by gravity to the clarifier where the solids settle to the bottom. The solids are removed by pumps and transported to the drum filter. The thickened slag/water mixture is further dewatered, and the solids are discharged to a belt conveyor. The conveyor transports the slag to an awaiting truck or dumpster for transport to the coal slurry storage tank for reburn in the gasifier.

Raw Water, Fire Protection, and Cycle Makeup Water Systems:

The raw water system supplies 1,700 gpm of cooling tower makeup, 200 gpm for the cycle makeup, and 15 gpm for service water use and potable water requirements. The pumps will be installed on an intake structure located on the river in close proximity to the plant.

The fire protection system provides water under pressure to the fire hydrants, hose stations, and fixed water suppression system within the buildings and structures. The system consists of pumps, underground and aboveground supply piping, distribution piping, hydrants, hose stations, spray systems, and deluge spray systems. One motor-operated booster pump is supplied on the intake structure of the cooling tower with a diesel engine pump installed on the intake structure located on the river.
The cycle makeup water system provides high quality demineralized water for makeup to the HRSG cycle, and for injection steam to the combustion turbine for control of NOx emissions and auxiliary boiler.

The cycle makeup system will consist of two 100 percent trains, each with a full-capacity activated carbon filter, primary cation exchanger, primary anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. The equipment will be skid-mounted and include a control panel, and associated piping, valves, and instrumentation.

Waste Treatment:

An onsite water treatment facility will treat all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment will be housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge dewatering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water will be provided. A 200,000 gallon storage tank will provide a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, and wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Layout Arrangement:

The development of the reference plant site to incorporate structures required for this technology is based on the assumption of a flat site. The IGCC gasifiers and related structures are arranged in a cluster, with the coal and slurry preparation facilities adjacent to the southeast, as shown in the conceptual general arrangement shown in Appendix II.

The gasifier and its associated process blocks are located west of the coal storage pile. The gas turbine and its ancillary equipment are sited northwest of the gasifier island, in a turbine building. The HRSG and stack are east of the gas turbine, with the steam turbine and its generator in a separate building to the north. Service and administration buildings are located at the west side of the steam turbine building.
The cooling tower heat sink for the steam turbine is located to the east of the steam turbine building. The air separation plant is further to the southwest, with storage tanks for liquid O_2 located near the gasifier and its related process blocks. Sulfur recovery, slag recovery, and wastewater treatment areas are located east and north of the gasifier.

The arrangement described above provides good alignment and positioning for major interfaces; relatively short steam, feedwater, and fuel gas pipelines; and allows good access for vehicular traffic. Transmission line access from the gas turbine and steam turbine step-up transformer to the switchyard is also maintained at short distances.

The air and gas path is developed in a short and direct manner, with ambient air entering an inlet filter/silencer located south of the gas turbine. The clean, hot, medium-Btu gas is conveyed to the turbine combustor for mixing with the air that remains on-board the machine. Turbine exhaust is ducted directly through the HRSG and then the 213-foot (65-meter) stack. The height of the stack is established by application of a good engineering practice rule from 40 CFR 51.00.

Access and construction laydown spaces are freely available on the periphery of the plant.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

2.10.2. Case-10 Overall Plant Performance and Emissions

The Case-10 IGCC plant produces a net output of 235 MWe at a net efficiency of 34.5 percent on an HHV basis. Overall performance for the entire plant is summarized in Table 2.10.3, which includes auxiliary power requirements.

Table 2.10. 3: Case-10 Overall Plant Performance

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine Power	187,150
Sweet Gas Expander Power	6,650
Steam Turbine	97,924
Total	291,724
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	270
Coal Milling	550
Coal Slurry Pumps	190
Slag Handling and Dewatering	100
Air Separation Unit Auxiliaries	20,080
Oxygen Compressor	10,570
Main Nitrogen Compressor	15,720
Claus Oxygen Compressor	30
Claus Tail Gas Recycle Compressor	1,080
HP Boiler Feedwater Pumps	1,160
IP Boiler Feedwater Pumps	80
LP Boiler Feedwater Pumps	340
Scrubber Pumps	50
Circulating Water Pumps	1,550
Cooling Tower Fans	920
Condensate Pump	150
Selexol Unit Auxiliaries	1,220
Gas Turbine Auxiliaries	400
Steam Turbine Auxiliaries	200
Claus Plant Auxiliaries	100
Miscellaneous Balance of Plant	1,000
Transformer Loss	670
TOTAL AUXILIARIES, kWe	56,430
Net Power, kWe	235,294
Net Plant Efficiency, % HHV	34.5
Net Heat Rate, Btu/kWh (HHV)	9,884
CONSUMABLES	
As-Received Coal Feed, lbm/h	210,010
Thermal Input, kWt	681,410
Gasifier Oxygen (95% pure), lbm/h	174,309
Water (for slurry), lbm/h	156,653

The operation of the combined cycle unit in conjunction with oxygen-blown Chevron-Texaco IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate. A salable by-product is produced in the form of elemental sulfur although no credit was taken for this product in the economic analysis (Section 4). A summary of the plant emissions is presented in Table 2.10.4.

The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the Selexol AGR process. The AGR process removes 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The tail gas treatment unit removes most of the sulfur from the Claus tail gas, which is recycled to the Claus unit. Tail gas from the tail gas treatment unit is recycled to the AGR system.

NOx emissions are limited by the use of humidification and nitrogen dilution to at least 9 ppm based on 15 percent oxygen in the flue gas. This is equivalent to 16 ppm at the 10.3 percent oxygen in the flue gas of the design. The ammonia is removed with process condensate prior to the low-temperature AGR process. This helps lower NOx levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of the syngas scrubber and the gas washing effect of the AGR absorber.

CO_2 emissions are equal to those of other coal-burning facilities on an intensive basis ($lb/10^6$ Btu), since a similar fuel is used. However, total CO_2 emissions are lower for a plant with this capacity due to the relatively high thermal efficiency.

Table 2.10. 4: Case-10 Overall Plant Emissions

	$lb/10^6$ Btu	Lbm/hr	lbm/MWh
SO_2	0.042	95.1	0.42
NO_x	0.023	52.1	0.23
Particulate	< 0.002	< 4.5	< 0.020
CO_2	200.3	464,940	1,980

2.11. Case-11: Commercially Offered Future IGCC with Shift Reactor and CO_2 Capture

Case-11 represents a commercially offered but not yet built and operated future IGCC plant with CO_2 Capture. This case is directly comparable with Case-10. This plant is an IGCC with CO_2 capture utilizing the Texaco high pressure, oxygen-blown, entrained flow, quench gasification technology, with shift reactor and syngas expander. The gasification system design is based on the Eastman Chemical Company Acetic Anhydride design. A single train GE 7FA gas turbine with a HRSG and a 1,800 psig / 1,000 ^oF / 1,000 ^oF steam cycle is used for power production the same as for all other IGCC cases in this study.

A brief performance summary for this plant reveals the following information. The Case-11 plant produces a net plant output of about 201 MW. The net plant heat rate and thermal efficiency are calculated to be 12,441 Btu/kWh and 27.4 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 0.25 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.11.2.

The ground rules for selection of plant configurations for Case-11 include:

- The IGCC plant is based on technology that has been offered on a commercial scale.
- The IGCC plant is sized for a single train GE 7FA gas turbine, producing in the range of 250 MWe net.
- The steam cycle operating parameters are 1,800 psig, 1,000 ^oF / 1,000 ^oF, 3.0 in. Hga.
- The coal (medium volatile bituminous) is the same one used in all other cases in this study.
- SO_2 emissions are equivalent to 99 percent overall removal.
- NOx emissions are based on current GE estimates for syngas combustion.
- Captured CO_2 is compressed to 2,000 psia and liquefied.

The high pressure Texaco quench gasifier was selected as the gasifier for both Case-10 and Case-11 plant configurations. It is currently being operated at the Eastman Chemical Company Acetic Anhydride Plant in Kingsport, Tennessee. The gasifier is well suited for either producing syngas for combustion or integration with a shift reactor for CO_2 and hydrogen production ($CO + H_2O => H_2 + CO_2$).

The following plant process selections have the points of reference for commercial scale demonstrations shown in parenthesis. The main criterion for this plant is to use commercially offered technology for process selection. The following have been selected:

- Gasifier: Texaco Quench, based on experience from Eastman Chemical Acetic Anhydride Plant. Oxygen Blown, Bituminous Coal, Operates at 950 psig. Gas exiting quench zone at saturation temperature.
- Gas Cooling: Wet scrubber followed by sequential syngas coolers. (Eastman Chemical Plant)
- Air Separation Plant: Single Train, 95 percent pure Oxygen (Wabash and TECO IGCC Plant) High-pressure oxygen compressor (Eastman Plant).
- Shift Reactor: Sour gas shift reactor to convert CO and H_2O to CO_2 and Hydrogen (Eastman Chemical Plant).

- Sulfur removal: Two-Stage Selexol acid gas removal process. Regenerated H_2S sent to sulfuric acid plant (TECO IGCC Plant). CO_2 compressed and liquefied (ABB Lummus/AES Warrior Run Plant)
- Power Generation: Sweet syngas expander. (GE Rotoflow) Single Train GE Model 7FA Gas Turbine (GE Test Burners on Hydrogen). Heat recovery Steam Generator (HRSG) on high temperature turbine exhaust, 1,800 psig steam cycle. Steam turbine for combined cycle generation.

2.11.1. Case-11 IGCC Plant Process Description

This IGCC plant design is based on the Chevron-Texaco Power and Gasification Corporation technology, which utilizes a pressurized entrained-flow, oxygen-blown gasification process. The plant configuration is based on the quench gasifier option operating at approximately 950 psig.

The power generation technology is based on selection of a gas turbine derived from the General Electric 7FA machine operating on hydrogen-rich syngas with much of the CO_2 removed. The plant is configured with the gasifier including processes to progressively cool and clean the gas, shift the CO to hydrogen and CO_2, making it suitable for CO_2 capture and combustion in the gas turbines. The resulting plant produces a net output of 201 MW_e at a net efficiency of 27.4 percent on an HHV basis.

The operation of the combined cycle unit in conjunction with oxygen-blown IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate (slag). A salable by-product is produced in the form of elemental sulfur although no credit was taken for this product in the economic analysis (Section 4). The low level of SO_2 in the plant emissions is achieved by the capture of the sulfur in the gas by the Selexol acid gas removal (AGR) process. The AGR process is designed to remove 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus unit with tail gas being recycled to the AGR system.

NOx emissions are limited to 9 ppm in the flue gas (normalized to 15 percent O_2) – or 27 ppm @ 3 percent O_2 --by the use of syngas humidification and nitrogen dilution. The ammonia is removed with process condensate prior to the low-temperature AGR process. This helps lower NOx levels as well. Selective catalytic reduction (SCR) can reduce emissions further to 5 ppmv level (at 15 percent O_2) or 15 ppmv @ 3 percent O_2 or lower if required.

Captured CO_2 will be compressed to 2,000 psia and liquefied.

Particulate discharge to the atmosphere is limited to low values by the gas-washing effect of the syngas scrubber and the AGR absorber.

2.11.1.1. Block Flow Diagram

The overall plant block flow diagram for Case-11 is shown in Figure 2.11.1. This diagram shows only the major process units and streams.

The pressurized Chevron-Texaco entrained-flow gasifier uses a slurry feed of water and coal, combined with oxygen, to produce a medium-Btu hot fuel gas (about 235 Btu/scf). Two gasifier trains are necessary to accommodate the increased coal throughput for this plant design. The gas goes through a series of gas coolers and cleanup processes including a CO-shift reactor, a carbon bed mercury removal system and an acid gas removal plant. The particulate captured by the scrubber are routed to the “black” water system, where the solids are separated and recycled to the gasifier. Regeneration H_2S gas from the AGR plant is fed to a Claus plant to produce elemental sulfur, and CO_2 is dried and compressed and for pipeline shipment as a liquid.

The hydrogen-rich clean gas exiting the AGR system is conveyed to the combustion turbine where it serves as fuel for the combustion turbine/HRSG/steam turbine power conversion system. The exhaust gas from the combustion turbine and HRSG is released to the atmosphere via a conventional stack.

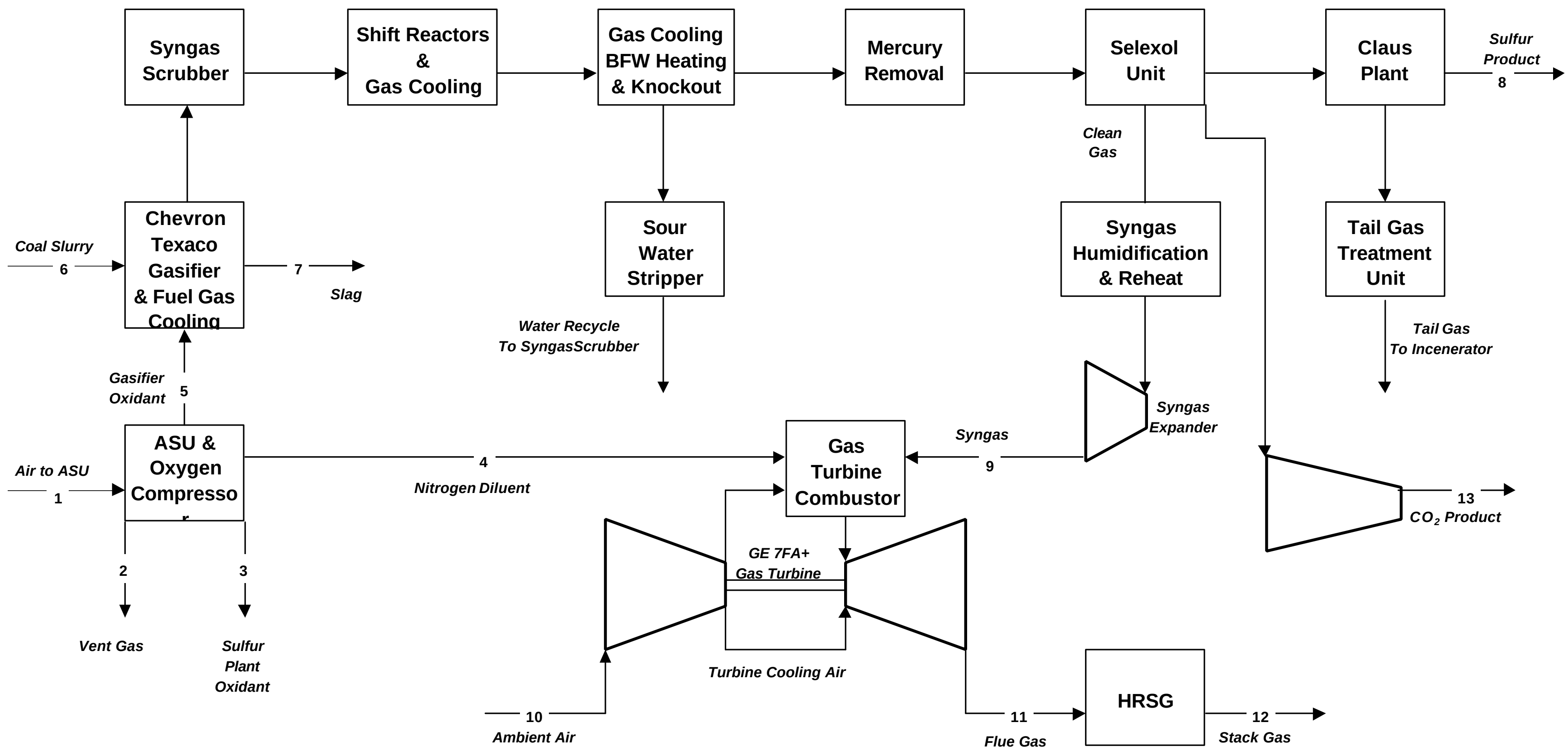


Figure 2.11.1: Case-11 Overall Plant Block Flow Diagram

This plant utilizes a combined cycle for combustion of the medium-Btu gas from the gasifier to generate electric power. A Brayton cycle using air and combustion products as working fluid is used in conjunction with a conventional subcritical steam Rankine cycle. The two cycles are coupled by generation of steam in the HRSG, by feedwater heating in the HRSG, and by heat recovery from the IGCC process (gas cooling modules).

2.11.1.2. Material and Energy Balance

Table 2.11.1 provides the material and energy balance for the IGCC plant. It is based on General Electric's estimate for the syngas fuel requirements for one 7FA gas turbine. Ambient operating conditions are indicated in the plant design basis. The pressurized entrained-flow gasifier uses a coal/water slurry and oxygen to produce a medium heating value fuel gas, which is shifted to maximize hydrogen and CO_2 content. The stream numbers shown in Table 2.11.1refer to Figure 2.11.1, a modified block flow diagram for the overall plant showing state points for all numbered streams.

Table 2.11. 1: Case-11 Overall Plant Material and Energy Balance

Mole Fraction	1	2	3	4	5	6	7	8	9	10	11	12	13
Ar	0.0094	0.0043	0.0322	0.0018	0.0360	0.0000	0.0000	0.0000	0.0101	0.0094	0.0092	0.0092	0.0002
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0014	0.0000	0.0000	0.0000	0.0001
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0162	0.0000	0.0000	0.0000	0.0006
CO_2	0.0003	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0131	0.0003	0.0051	0.0051	0.9881
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7034	0.0000	0.0000	0.0000	0.0110
H_2O	0.0104	0.0272	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.2485	0.0104	0.1605	0.1605	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000
N_2	0.7722	0.9677	0.0178	0.9982	0.0140	0.0000	0.0000	0.0000	0.0073	0.7722	0.7137	0.7137	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2077	0.0000	0.9500	0.0000	0.9500	0.0000	0.0000	0.0000	0.0000	0.2077	0.1114	0.1114	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flow (lb_{mol}/hr)	27,144	10,397	109	10,814	5,816	12,665	0	20	20,583	109,638	133,635	133,635	10,628
V-L Flow (lb/hr)	783,228	289,096	3,521	303,182	187,431	168,445	0	5,234	155,432	3,163,540	3,622,150	3,622,150	462,688
V-L Flow (acfm)	178,235	58,623	359	4,875	759	419	0	0	9,765	719,910	2,433,733	1,202,998	337
Solids Flow (lb/hr)	0	0	0	0	0	225,822	58,472	0	0	0	0	0	0
Temperature (°F)	80	63	90	284	354	62	300	341	535	80	1,051	280	169
Pressure (psia)	14.7	16.4	30.0	295.0	1,115.0	1,050.0	14.4	23.5	375.0	14.7	14.8	14.7	2,000.0
Density (lb/ft^3)	0.073	0.082	0.164	1.037	4.117	---	---	330.439	0.265	0.073	0.025	0.050	22.908
Average Molecular Weight	28.85	27.81	32.18	28.03	32.23	---	---	256.53	7.55	28.85	27.10	27.10	43.53

The following paragraphs describe the process sections in more detail. The equipment required is described in Appendix I.

2.11.1.3. Gasifier Island

Coal Grinding and Slurry Preparation:

Coal is fed onto a conveyor by vibratory feeders located below each coal silo. The conveyor feeds the coal to an inclined conveyor that delivers the coal to the rod mill feed hopper. The feed hopper provides a surge capacity of about two hours and contains two hopper outlets. A vibrating feeder on each hopper outlet supplies the weigh feeder, which in turn feeds a rod mill. Each rod mill is sized to process 60 percent of the coal feed requirements of the gasifier. The rod mill grinds the coal and wets it with treated slurry water transferred from the slurry water tank by the slurry water pumps. The coal slurry is then discharged into the rod mill product tank. The slurry is then pumped from the rod mill product tank to the slurry storage and slurry blending tanks.

The coal grinding system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. The degree of dust suppression required will depend on local environmental regulations. All of the tanks are equipped with vertical agitators to keep the coal slurry solids suspended.

The equipment in the coal grinding and slurry preparation system is fabricated of materials appropriate for the abrasive environment present in the system. The tanks and agitators are rubber lined. The pumps are either rubber lined or hardened metal to minimize erosion. Piping is fabricated of high-density polyethylene (HDPE).

Gasifier:

This plant utilizes one gasification train to process a total of 2,710 tons of coal per day. The gasifier operates at approximately 60 percent maximum capacity. Due to the properties of the coal such as high ash and low as-received heating value, additional coal mass flow is needed to reach the syngas requirements. This results in a need for additional oxygen and resultant auxiliary power requirements. This is reflected in the higher than expected heat rate. Also, additional coal is required due to the decreased efficiency associated with the CO-shift and CO_2 capture. The slurry feed pump takes suction from the slurry run tank, and the discharge is sent to the feed injector of the Chevron-Texaco gasifier. Oxygen from the ASU is vented during preparation for startup and is sent to the feed injector during normal operation. The air separation plant supplies 2,249 tons of 95 percent pure oxygen per day to the gasifier.

The gasifier vessel is a refractory-lined, high-pressure combustion chamber. The coal slurry is transferred from the slurry storage tank to the gasifier with a high-pressure pump. At the top of the gasifier vessel is located a combination fuel injector through which coal slurry feedstock and oxidant (oxygen) are fed. The coal slurry and the oxygen feeds react in the gasifier at about 965 psia at a high temperature to produce syngas.

The syngas consists primarily of hydrogen and carbon monoxide, with lesser amounts of water vapor and carbon dioxide, and small amounts of hydrogen sulfide, carbonyl sulfide, methane, argon, and nitrogen. The heat in the gasifier liquefies the coal ash. Hot syngas and molten solids from the reactor flow downward into a water-filled quench chamber where the syngas is cooled and the slag solidifies. Raw syngas then flows to the syngas scrubber for removal of any entrained solids. The slag collects in the water sump at the bottom of the gasifier and is removed periodically, using a lock hopper system.

Solids collected in the quench gasifier water sump are removed by gravity and forced circulation of water from the lock hopper circulating pump. Fine material, which does not settle as easily, is removed in the gasification blowdown and goes to the vacuum flash drum by way of the syngas scrubber.

Syngas Scrubbing:

Process condensate from the sour stripper is mixed with the raw syngas. The water/syngas mixture enters the syngas scrubber and is directed downward by a dip tube into a water sump at the bottom of the syngas scrubber. Most of the solids are separated from the syngas at the bottom of the dip tube as the syngas goes upwards through the water. From the overhead of the syngas scrubber, the syngas enters the low-temperature gas cooling section for further cooling.

The water removed from the syngas scrubber contains all the solids that were not removed in the quench gasifier water sump. In order to limit the amount of solids recycled to the quench chamber, a continuous blowdown stream is removed from the bottom of the syngas scrubber.

Slag Handling System:

The slag handling system removes solids from the gasification process equipment. These solids consist of a small amount of unconverted carbon and essentially all of the ash contained in the feed coal. These solids are in the form of glass, which fully encapsulates any metals.

Water Gas Shift:

Hot, particulate-free syngas from the scrubber is fed to the CO-shift reactor. A set of high-temperature shift reactors is used to shift the bulk of the CO in the fuel gas to CO_2. Heat exchange between reaction stages helps maintain a moderate reaction temperature. The shift catalyst also promotes COS hydrolysis. A two-stage shift was utilized in order to maximize CO conversion while maintaining reasonable reactor volumes.

The shifted raw gas temperature exiting the second shift converter is approximately 553°F. This stream is cooled to 370°F in a low-temperature economizer. The fuel gas stream is cooled to 103°F in a series of low-temperature economizers and then routed to the Mercury Removal section and the Selexol unit. Fuel gas condensate is recovered and routed to a sour water drum.

Mercury Removal:

Mercury removal is based on packed beds of sulfur-impregnated carbon similar to what has been used at the Eastman Chemical gasification plant. A single bed of sulfur-impregnated carbon with approximately a 20-second superficial gas residence time achieves 95 percent reduction of mercury in addition to removal of other volatile heavy metals such as arsenic.

Sulfur Removal and Recovery / Carbon Dioxide Removal and Compression:

A unique feature of this power plant configuration is that H_2S and CO_2 are removed within the same process system, the double-stage Selexol unit. The purpose of the Selexol unit is to preferentially remove H_2S as a product stream and then to preferentially remove CO_2 as a separate product stream.

Cool, dry, and particulate-free synthesis gas enters the first absorber unit at approximately 103°F. In this absorber, H_2S is preferentially removed from the fuel gas stream. This is achieved by "loading" the lean Selexol solvent with CO_2. The rich solution leaving the bottom of the absorber is regenerated in a stripper through the indirect application of thermal energy via condensing low-pressure steam in a reboiler. The stripper acid gas stream, consisting of H_2S and CO_2 (with the balance mostly H_2O), is then sent to the Claus unit.

Sweet fuel gas flowing from the first absorber is cooled and routed to the second absorber unit. In this absorber, the fuel gas is contacted with "unloaded" lean solvent. The solvent removes approximately 97 percent of the CO_2 remaining in the fuel gas stream. A CO_2 balance is maintained by hydraulically expanding the CO_2-saturated rich solution and then flashing CO_2 vapor off the liquid at reduced pressure. Sweet fuel gas off the second absorber is warmed and humidified in the fuel gas saturator, reheated and expanded, and then sent to the burner of the combustion turbine.

Sweet, hydrogen-rich fuel gas from the Selexol unit is piped to the bottom of the saturator. The sweet fuel gas rises up through the column while warm water flows downward counter-currently. Internal trays are used to enhance the mass transfer of water vapor into the fuel gas. This process both humidifies the fuel gas and increases its sensible heat content.

Syngas Expander:

After sulfur removal, the sweet fuel gas is saturated with condensate, reheated and depressurized through an expander from 825 psia to 380 psia, which is near the pressure required by the gas turbine. The expander generates 6,570 kW_e.

Sulfur Recovery System:

The sulfur recovery unit is a Claus bypass type sulfur recovery unit utilizing oxygen instead of air. The Claus plant produces molten sulfur by converting approximately a third of the H_2S in the feed to SO_2, then converting the remaining H_2S and SO_2 to elemental sulfur and water. The combination of Claus technology and tail gas recycle to the Selexol results in an overall sulfur recovery of 99 percent. Utilizing oxygen instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The sulfur plant will produce approximately 63 tons of elemental sulfur per day.

CO_2 Compression and Drying:

CO_2 is flashed from the rich solution at two pressures. The bulk of the CO_2 is flashed off at approximately 50 psia, while the remainder is flashed off at atmospheric pressure. The second low-pressure CO_2 stream is "boosted" to 50 psia and then combined with the first CO_2 stream. The combined flow is then compressed in a multiple-stage, intercooled compressor to supercritical conditions at 2,000 psig. During compression, the CO_2 stream is dehydrated with triethylene glycol. The virtually moisture-free supercritical CO_2 steam is then ready for pipeline transportation.

Air Separation Plant:

The air separation plant is designed to produce a nominal output of 2,291 tons/day of 95 percent pure O_2. Most of the oxygen is used in the gasifier. A small portion, approximately 42 tons/day, is used in the Claus plant. The plant is designed with one production train. The air compressor is powered by an electric motor. Approximately 3,638 tons/day of nitrogen are also recovered, compressed and used as dilution in the gas turbine combustor.

In this air separation process, air is compressed to 85 psia and then cooled in a water-scrubbing spray tower. The cooled air enters a reversing heat exchanger, where it is cooled to the liquefaction point prior to entering a double column (high/low pressure) separator.

Flare Stack:

A self-supporting, refractory-lined, carbon steel flare stack is provided to combust and dispose of product gas during startup, shutdown, and upset conditions. The flare stack is provided with multiple pilot burners, fueled by natural gas or propane, with pilot home monitoring instrumentation.

2.11.1.4. Power Generation System

Gas Turbine Generator:

The gas turbine generator selected for this application is the same General Electric MS 7001FA model turbine chosen for the Tampa Electric IGCC Demonstration Project. There are over 140 GE 7FA and GE 9FA units ordered or in operation. This machine is an axial flow, single spool, constant speed unit, with variable inlet guide vanes. The machine is designed for maximum reliability and efficiency with low maintenance. The turbine includes advanced bucket cooling techniques, compressor aerodynamic design and advanced alloys, enabling a higher firing temperature that the previous generation machines. The standard production version of this machine, fired with natural gas, will develop a compressor pressure ratio of 15.2:1 and a rotor inlet temperature of almost 2350°F.

In this service, with syngas from an IGCC plant, the machine requires some modifications to the burner and turbine nozzles in order to properly combust the hydrogen-rich gas and expand the combustion products in the turbine section of the machine. A reduction in rotor inlet temperature of about 50°F results, relative to a production model 7FA machine firing natural gas. This temperature reduction is necessary to not exceed design basis gas path temperatures throughout the expander. Power output for the gas turbine is 187,150 kW at the site inlet conditions.

Steam Generation:

The heat recovery steam generator (HRSG) is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the gas turbine exhaust gas when firing medium-Btu gas. The HP drum produces steam at main steam pressure, while the IP drum produces steam for export to the cold reheat.

The HRSG drum pressures are nominally 1,800 and 420 psia for the HP and IP turbine sections, respectively. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater heating, and also provides deaeration of the condensate.

Steam Turbine Generator and Auxiliaries:

The steam turbine consists of a HP section, an IP section, and one double-flow LP section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing. The LP turbine has a last-stage bucket length of 30 inches.

Main steam from the HRSG and Gasifier Island is combined in a header, and then passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the HRSG for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at 400 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser.

The generator is a hydrogen-cooled synchronous type, generating power at 23 kV. A static, transformer type exciter is provided. The generator is cooled with a hydrogen gas recirculation system using fans mounted on the generator rotor shaft. The heat absorbed by the gas is removed as it passes over finned tube gas coolers mounted in the stator frame. Gas is prevented from escaping at the rotor shafts by a closed-loop oil seal system. The oil seal system consists of a storage tank, pumps, filters, and pressure controls, all skid-mounted.

The steam turbine generator is controlled by a triple-redundant, microprocessor-based electro-hydraulic control system. The system provides digital control of the unit in accordance with programmed control algorithms, color CRT operator interfacing, and datalink interfaces to the balance-of-plant distributed control system (DCS), and incorporates on-line repair capability.

Condensate System:

The condensate system transfers condensate from the condenser hotwell to the deaerator, through the gland steam condenser, gasifier, and the low-temperature economizer section in the HRSG. The system consists of one main condenser; two 50 percent capacity, motor-driven, vertical condensate pumps; one gland steam condenser; and a low-temperature tube bundle in the HRSG. Condensate is delivered to a common discharge header through separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided to maintain minimum flow requirements for the gland

Feedwater System:

The function of the feedwater system is to pump the various feedwater streams from the deaerator storage tank in the HRSG to the respective steam drums. Two 50 percent capacity boiler feed pumps are provided. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

Main and Reheat Steam:

The function of the main steam system is to convey main steam generated HRSG from the HRSG superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey steam from the HP turbine exhaust to the HRSG reheater, and to the turbine reheat stop valves.

Main steam at approximately 1,900 psig / 1,000°F exits the HRSG superheater through a motor-operated stop/check valve and a motor-operated gate valve, and is routed to the HP turbine. Cold reheat steam at approximately 450 psig / 645°F exits the HP turbine, flows through a motor-operated isolation gate valve, to the HRSG reheater. Hot reheat steam at approximately 420 psig / 1,000°F exits the HRSG reheater through a motor-operated gate valve and is routed to the IP turbines.

Circulating Water System:

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the condenser to condense the main turbine exhaust steam. The system also supplies cooling water to the auxiliary cooling system. The heat transferred from the steam to the circulating water in the condenser is removed by a mechanical draft cooling tower.

The system consists of two 50 percent capacity vertical circulating water pumps, a mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counterflow mechanical draft cooling tower.

The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of the condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

In addition to the condenser, additional cooling is required for the air separation unit. This amounts to an additional 175 MM-Btu/hr.

2.11.1.5. Other Balance of Plant Equipment

Coal Handling System:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading will be done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron, and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the two silos.

Technical Requirements and Design Basis:

- Coal burn rate:
  -Maximum coal burn rate = 220,093 lbm/h = 121 tph plus 10 percent margin = 134 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
  -Average coal burn rate = 207,000 lbm/h = 104 tph (based on MCR rate multiplied by an 85 percent capacity factor)
- Coal delivered to the plant by unit trains:
  -One and one-half unit trains per week at maximum burn rate
  -One unit train per week at average burn rate
  -Each unit train shall have 10,000 tons (100-ton cars) capacity
  -Unloading rate = 9 cars/hour (maximum)
  -Total unloading time per unit train = 11 hours (minimum)
  -Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
  -Reclaim rate = 300 tph
- Storage piles with liners, run-off collection, and treatment systems:
  -Active storage = 8,500 tons (72 hours at maximum burn rate)
  -Dead storage = 85,000 tons (30 days at average burn rate)

Table 2.11. 2: Case-11Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	94
Active Storage, tons	8,500
Dead Storage, tons	85,000

Slag Handling:

The plant includes two slag handling systems: one system handles the slag generated at the base of the quench gasifier, and the second handles the slag removed from the syngas in the scrubber.

The coarse slag handling system conveys, stores, and disposes of slag removed from the gasification process. Slag exits through the slag tap into a water bath in the bottom of the quench vessel. A slag crusher receives slag from the water bath and grinds the material into pea-sized fragments. A slag/water slurry that is between 5 and 10 percent solids flows out of the bottom of the quench vessel through a pressure letdown valve into a lined ground level tank. The components listed above, up to the pressure letdown valve, are within the gasifier pressure boundary and at high pressure.

Drag chain conveyors from the bottom of the slag-handling tank remove the cooled, dewatered slag. The slag mixture is discharged to a vibrating screen where the fine slag is removed. The larger screened slag is stored in a storage bin. The bin is sized for a nominal holdup capacity of approximately 72 hours of full-load operation. At periodic intervals, a convoy of slag hauling trucks will transit the unloading station underneath the hopper and remove a quantity of slag for disposal. Approximately 18 truckloads per day are required to remove the total quantity of slag produced by the plant operating at nominal rated power.

The fine slag handling system removes the slag removed by the scrubber. The system consists of a clarifier and rotary drum filter. The slag/water mixture flows by gravity to the clarifier where the solids settle to the bottom. The solids are removed by pumps and transported to the drum filter. The thickened slag/water mixture is further dewatered, and the solids are discharged to a belt conveyor. The conveyor transports the slag to an awaiting truck or dumpster for transport to the coal slurry storage tank for reburn in the gasifier.

Raw Water, Fire Protection, and Cycle Makeup Water Systems:

The raw water system supplies 1,700 gpm of cooling tower makeup, 200 gpm for the cycle makeup, and 15 gpm for service water use and potable water requirements. The pumps will be installed on an intake structure located on the river in close proximity to the plant.

The fire protection system provides water under pressure to the fire hydrants, hose stations, and fixed water suppression system within the buildings and structures. The system consists of pumps, underground and aboveground supply piping, distribution piping, hydrants, hose stations, spray systems, and deluge spray systems. One motor-operated booster pump is supplied on the intake structure of the cooling tower with a diesel engine pump installed on the intake structure located on the river.
The cycle makeup water system provides high quality demineralized water for makeup to the HRSG cycle, and for injection steam to the combustion turbine for control of NOx emissions and auxiliary boiler.

The cycle makeup system will consist of two 100 percent trains, each with a full-capacity activated carbon filter, primary cation exchanger, primary anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. The equipment will be skid-mounted and include a control panel, and associated piping, valves, and instrumentation.

Waste Treatment:

An onsite water treatment facility will treat all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment will be housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge dewatering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water will be provided. A 200,000-gallon storage tank will provide a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, and wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Layout Arrangement:

The development of the reference plant site to incorporate structures required for this technology is based on the assumption of a flat site. The IGCC gasifiers and related structures are arranged in a cluster, with the coal and slurry preparation facilities adjacent to the southeast, as shown in the conceptual general arrangement shown in Appendix II.

The gasifier and its associated process blocks are located west of the coal storage pile. The gas turbine and its ancillary equipment are sited northwest of the gasifier island, in a turbine building. The HRSG and stack are east of the gas turbine, with the steam turbine and its generator in a separate building to the north. Service and administration buildings are located at the west side of the steam turbine building.
The cooling tower heat sink for the steam turbine is located to the east of the steam turbine building. The air separation plant is further to the southwest, with storage tanks for liquid O_2 located near the gasifier and its related process blocks. Sulfur recovery, slag recovery, and wastewater treatment areas are located east and north of the gasifier.

The arrangement described above provides good alignment and positioning for major interfaces; relatively short steam, feedwater, and fuel gas pipelines; and allows good access for vehicular traffic. Transmission line access from the gas turbine and steam turbine step-up transformer to the switchyard is also maintained at short distances.

The air and gas path is developed in a short and direct manner, with ambient air entering an inlet filter/silencer located south of the gas turbine. The clean, hot, medium-Btu gas is conveyed to the turbine combustor for mixing with the air that remains on-board the machine. Turbine exhaust is ducted directly through the HRSG and then the 213-foot (65-meter) stack. The height of the stack is established by application of a good engineering practice rule from 40 CFR 51.00.

Access and construction laydown spaces are freely available on the periphery of the plant.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

2.11.2. Case-11 Overall Plant Performance and Emissions

The Case-11 IGCC plant produces a net output of 201 MWe at a net efficiency of 27.4 percent on an HHV basis. Overall performance for the entire plant is summarized in Table 2.11.3, which includes auxiliary power requirements.

Table 2.11. 3: Case-11 Overall Plant Performance

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine Power	187,150
Sweet Gas Expander Power	6,570
Steam Turbine	96,550
Total	290,270
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	290
Coal Milling	590
Coal Slurry Pumps	200
Slag Handling and Dewatering	110
Air Separation Unit Auxiliaries	23,302
Oxygen Compressor	10,990
Main Nitrogen Compressor	15,460
Claus Oxygen Compressor	30
CO_2 Compressor	25,644
HP Boiler Feedwater Pumps	1,160
IP Boiler Feedwater Pumps	80
LP Boiler Feedwater Pumps	200
Scrubber Pumps	50
Circulating Water Pumps	1,590
Cooling Tower Fans	950
Condensate Pump	140
Double Stage Selexol Unit Auxiliaries	6,100
Gas Turbine Auxiliaries	400
Steam Turbine Auxiliaries	200
Claus Plant Auxiliaries	100
Miscellaneous Balance of Plant	1,000
Transformer Loss	680
TOTAL AUXILIARIES, kWe	89,266
Net Power, kWe	201,004
Net Plant Efficiency, % HHV	27.4%
Net Heat Rate, Btu/kWh (HHV)	12,441
CONSUMABLES	
As-Received Coal Feed, lbm/h	225,822
Thermal Input, kWt	732,714
Gasifier Oxygen (95% pure), lbm/h	187,431
Water (for slurry), lbm/h	168,445

The operation of the combined cycle unit in conjunction with oxygen-blown Chevron-Texaco IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate. A salable by-product is produced in the form of elemental sulfur although no credit was taken for this product in the economic analysis (Section 4). A summary of the plant emissions is presented in Table 2.11.4.

The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the Selexol AGR process. The AGR process removes 99 percent of the sulfur compounds in the fuel gas. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur.

NOx emissions are limited by the use of humidification and nitrogen dilution to at least 9 ppm based on 15 percent oxygen in the flue gas. This is equivalent to 16 ppm at the 10.3 percent oxygen in the flue gas of the design. The ammonia is removed with process condensate prior to the low-temperature AGR process. This helps lower NOx levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of the syngas scrubber and the gas washing effect of the AGR absorber.

CO_2 emissions are the result of 90 percent CO_2 capture.

Table 2.11. 4: Case-11 Overall Plant Emissions

	lb/10^6 Btu	Lbm/hour	lbm/MWh
SO_2	0.042	102.2	0.50
NO_x	0.023	56.0	0.28
Particulate	< 0.002	< 4.9	< 0.024
CO_2	20.0	49,896	248

2.12. Case–12: Indirect Gasification of Coal via Chemical Looping

This case is based on an advanced Chemical Looping gasification process being developed by ALSTOM. This case, without CO_2 capture, provides the basis of comparison to Case-13 (Indirect Gasification of Coal and CO_2 capture via Chemical Looping). In this case an advanced Chemical Looping concept is used to indirectly provide the oxygen for the gasification of coal rather than direct utilization of ambient air or supply of oxygen by other means (Cryogenic Air Separation Unit or Oxygen Transport Membrane). CO_2 is not captured in this concept. The chemical looping concept supplies the oxygen to the gasification process without the large efficiency penalty associated with a cryogenic type Air Separation Unit (Cases 8, 9, 10, 11). Additionally the large investment cost associated with both cryogenic type Air Separation Units or Oxygen Transport Membrane type oxygen supply systems is avoided. The trade off of course is a somewhat more complex gasification process. Through the use of this chemical looping gasification process, Medium Btu Gas (MBG) is produced from an air fired gasificification system.

Power production is provided from a single train GE-7FA gas turbine with a Heat Recovery Steam Generator (HRSG) and an 1,800 psig / 1,000 ^oF / 1,000 ^oF steam cycle. This is the same power production equipment that is used for all other cases in this study.

The overall plant is shown as a simplified block flow diagram in Figure 2.12.1, and Table 2.12.1 shows the key stream conditions. Crushed coal and limestone are supplied to the gasifier system with a conventional feed system. Air is supplied to the gasifier through a primary air fan, and an ID fan takes flue gas from the gasifier and sends it to a stack.

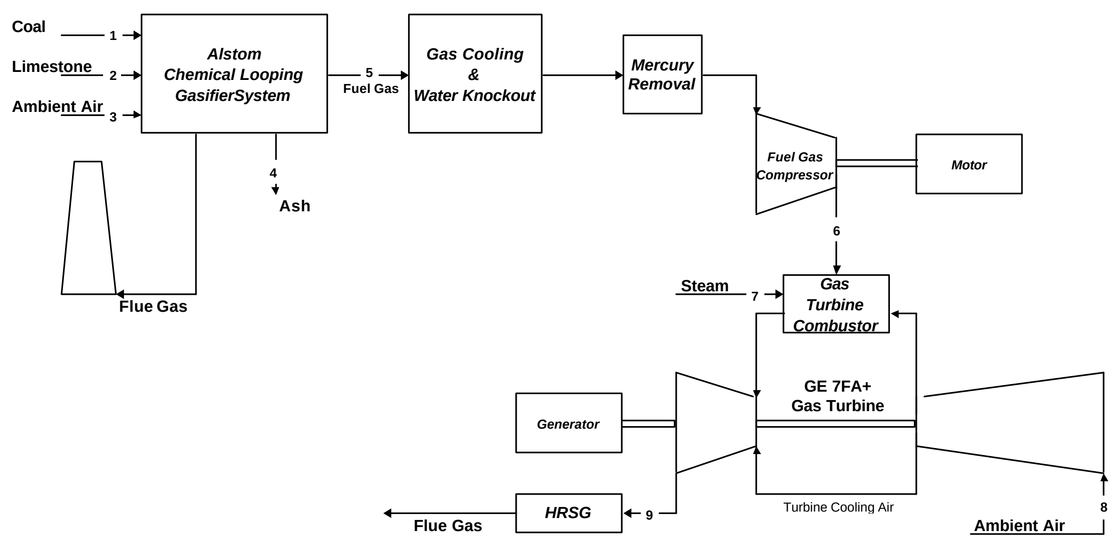


Figure 2.12. 1: Case-12: Simplified Block Flow Diagram

The fuel gas from the gasifier is cooled in a feedwater heater and gas cooler to 100°F to knockout water and it then continues through a mercury removal bed. The gas is compressed to 300 psia in a multi-stage compressor and humidified before going to the combustion turbine. The combustion turbine is steam injected for NOx control. Flue gas from the turbine passes through a heat recovery steam generator (HRSG) where high-pressure steam is generated to produce additional power.

Table 2.12. 1: Case-12: Overall Plant Stream Report

Mole Fraction	1	2	3	4	5	6	7	8	9
Ar	0.0000	0.0000	0.0094	0.0000	0.0000	0.0000	0.0000	0.0094	0.0078
CO	0.0000	0.0060	0.0000	0.0000	0.5231	0.5484	0.0000	0.0000	0.0000
CO_2	0.0000	0.0001	0.0003	0.0000	0.0628	0.0659	0.0000	0.0003	0.0749
H_2	0.0000	0.9283	0.0000	0.0000	0.3520	0.3691	0.0000	0.0000	0.0000
H_2O	0.0000	0.0656	0.0104	0.0000	0.0500	0.0166	1.0000	0.0104	0.1587
N_2	0.0000	0.0000	0.7722	0.0000	0.0000	0.0000	0.0000	0.7722	0.6417
NH_3	0.0000	0.0000	0.0000	0.0000	0.0121	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.2077	0.0000	0.0000	0.0000	0.0000	0.2077	0.1169
Total	0.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (lb_{mol}/hr)	0	0	22,324	0	17,024	16,238	13,795	111,092	133,676
V-L Flowrate (lb/hr)	0	0	644,138	0	327,422	313,457	248,522	3,205,490	3,767,470
Solids Flowrate (lb/hr)	197,428	30,363	0	77,113	0	0	0	0	0
Temperature (°F)	80	225	80	103	1174	293	500	80	1095
Pressure (psia)	14.7	280.0	14.7	63.2	14.7	290.0	250.0	14.7	14.8
Density (lb/ft^3)	---	---	0.07	0.18	0.02	0.69	0.44	0.07	0.02
Average Molecular Weight	---	---	28.85	17.03	19.23	19.30	18.02	28.85	28.18

A brief performance summary for this plant reveals the following results. The Case-12 plant produces a net plant output of about 265 MW. The net plant heat rate and thermal efficiency are calculated to be 8,248 Btu/kWh and 41.4 percent respectively (HHV basis). Carbon dioxide emissions are about 1.71 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.12.3.

2.12.1. Case 12 Gasifier Island Process Description and Equipment

This section describes the Gasifier Island processes for Case-12 and includes a simplified process flow diagram (PFD), material and energy balance and equipment description for the advanced Chemical Looping system. The equipment description includes only the major components included in the Gasifier Island.

It should be emphasized that the advanced Chemical Looping gasification process described for this case is only conceptual at this time. A significant effort encompasing experimental work related to reaction rates, solids regeneration cycles, and fine particulate removal would be necessary to continue development of this gasification process. Additionally, high temperature air heater development would also be required.

2.12.1.1. Gasification Process Description and Process Flow Diagrams

Figure 2.12.2 shows a simplified process flow diagram for the Case-12 Gasifier Island which utilizes a process for the indirect gasification of coal via chemical looping. This process description briefly describes the function of the major equipment and systems included within the Gasifier Island. Selected mass flow rates (lbm/hr) and temperatures (^{o}F) are shown on this figure. Complete data for all state points are shown in Table 2.12.2.

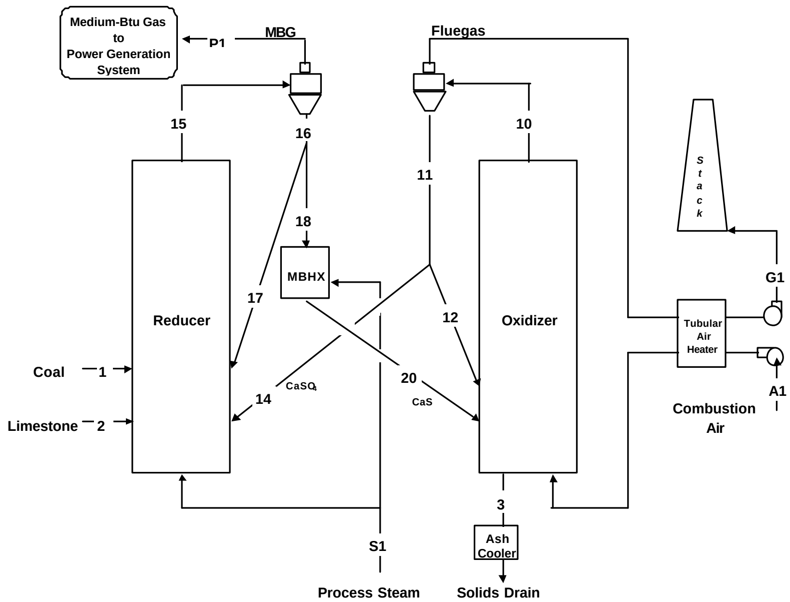


Figure 2.12. 2: Case-12: Simplified Gasifier Island Process Flow Diagram

Oxidizer:

The basic concept for Case-12 is to use a stream of recirculating bed solids as a chemical looping oxygen carrier whereby the oxygen is picked up by the solids from air in the oxidizer vessel. The oxidizer operates at about 2,000 ^{o}F. The basic chemistry in the oxidizer is shown in the following reaction.

$$CaS + 2O_2 \rightarrow CaSO_4 + Heat$$

Therefore the purpose of the oxidizer is to react oxygen from the air with CaS to form hot $CaSO_4$ with a minimal amount of excess air.

There is a high nitrogen content gas stream (Stream 10) leaving the oxidizer, which is cleaned of solids, cooled in a tubular air heater and finally exhausted to the atmosphere after passing through the Induced Draft (ID) fan.

Draining hot solids from the oxidizer vessel through water-cooled fluidized bed ash coolers (Stream 3) controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash coolers is feedwater from the steam cycle.

Oxidizer Solids Recirculation:

The solids separated from the gas/solids mixture leaving the oxidizer (Stream 11) are rich in $CaSO_4$ and are split into two streams. One stream is simply recirculated to the oxidizer (Stream 12) and the other, which is the oxygen source for the gasification reactions, is transported to the reducer (Stream 14).

Reducer:

The oxygen now carried by the solids ($CaSO_4$ in Stream 14) is then reacted sub-stoichiometricly with the carbon and hydrogen contained within the coal (Stream 1) and recycle solids (Stream 17) in the reducer vessel to form a medium-Btu fuel gas stream

(Stream P1). The reducer operates at about 1,700 ^oF and reduces the $CaSO_4$ to CaS. Stream 15, fuel gas comprised of primarily CO, H_2, CO_2 and H_2O vapor and entrained hot solids, flows through a particulate removal device, where hot solids are removed and recirculated. The principal overall reactions, which are endothermic, are shown below:

$$4C + CaSO_4 + Heat \rightarrow 4CO + CaS$$

$$4H\text{ (in Coal)} + CaSO_4 + Heat \rightarrow CaS + 2H_2O$$

$$H_2O + C + Heat \rightarrow H_2 + CO$$

Limestone (Stream 2) is added to the reducer to remove sulfur contained in the coal. The limestone and the sulfur combine to form CaS in the reducer, as shown below, which is used as the oxygen carrier in the chemical looping reactions described above. Solids are removed from the system in the oxidizer as shown in Figure 2.12.2 to avoid a buildup of CaS and remove the captured sulfur.

$$Ca + S \rightarrow CaS$$

The medium-Btu fuel gas stream (Stream 15) leaving the reducer is cleaned of solids, cooled, and provides the feed stream (Stream P1) to the Power Generation System where the fuel is compressed and utilized in a combustion turbine, which is part of a combined cycle.

Reducer Solids Recirculation:

The solids separated out of the gas/solids mixture leaving the reducer (Stream 16) are rich in CaS. This stream is split with the uncooled fraction (Stream 17) recirculated back to the reducer and the remaining fraction (Stream 18) cooled in the Moving Bed Heat Exchanger (MBHE). Exchanging heat with the steam cycle working fluid cools the hot solids in the MBHE. The solids leaving the MBHE, now rich in oxygen deficient CaS are returned to the Oxidizer (Stream 20) to absorb more oxygen, which completes the solids loop.

A small quantity of process steam (Stream S1) is introduced into the MBHE and into the reducer vessel. The purpose of this steam is to help initiate the reducer reactions.

The temperature in the reducer is controlled to the proper level by properly splitting the flow of hot recirculated solids leaving the particulate removal system. One stream (Stream 17) is uncooled and flows directly back to the reducer while a second stream (Stream 18) flows through the Moving Bed Heat Exchanger (MBHE) where the solids are cooled before returning to the oxidizer (Stream 20).

High Temperature Air Heater:

The cooling of the flue gas stream leaving the oxidizer vessel is done in a high temperature tubular air heater where the sensible heat of the high nitrogen content is transferred to the incoming combustion air stream (Stream A1).

2.12.1.2. Material and Energy Balance

Table 2.12.2 shows the Gasifier Island material and energy balance for Case-12. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-12 simplified PFD for the Gasifier Island (Figure 2.12.2). This performance was calculated at MCR conditions for this unit.

The MCR condition for Case-12 is defined as operating the system at a condition that fully loads the combustion turbine.

Table 2.12. 2: Case-12 Gasifier Island Material and Energy Balance

		Inputs					Outputs		
	Stream #	1	2	A1		S1	G1	P1	3
Constituant	(Units)	Coal	Limestone	Primary Air	Infiltration Air	Process Steam	Cold Fluegas	Product Gas	Solids Drain
Carbon	(lbm/hr)	122504	0	0	0	0	0	0	1512
Hydrogen	"	7048	0	0	0	0	0	12074	0
Oxygen	"	6239	0	148098	1481	0	13876	0	0
Nitrogen	"	2882	0	487774	4878	0	492701	0	0
Sulfur	"	4620	0	0	0	0	0	0	0
CO	"	0	0	0	0	0	0	249438	0
CO2	"	0	0	0	0	0	16904	47074	0
H2O	"	7877	0	8266	83	57930	8350	15332	0
NH3	"	0	0	0	0	0	0	3505	0
CaCO3	"	0	28846	0	0	0	0	0	288
CaO	"	0	0	0	0	0	0	0	7919
CaSO4	"	0	0	0	0	0	0	0	19618
Ash	"	46257	1518	0	0	0	0	0	47776
Total	"	197428	30364	644138	6442	57930	531831	327423	77113
Temp	(Deg F)	80	80	80	80	350	225	100	520
Press	(psia)	14.7	14.7	14.7	14.7	150	14.7	14.7	14.7

2.12.1.3. Gasifier Island Equipment

This section describes major equipment included in the Gasifier Island for Case-12. The major components included in the Gasifier Island include the Reducer vessel, Oxidizer vessel, ash coolers, fuel feed system, fuel silos, sorbent feed system, sorbent silo, particulate removal systems, seal pots, external moving bed heat exchanger (MBHE), superheater, reheater, evaporator, economizer, high temperature air heater, and draft system.

Figures 2.12.3 and 2.12.4 show general arrangement drawings of the Case-12 Chemical Looping Gasifier Island**.** The complete Gasifier Island Equipment List for Case-12 is shown in Appendix I. Appendix II shows several additional drawings of the Gasifier (key plan view, gasifier plan view, side elevation, and various sectional views) as well as a drawing of the overall site layout.

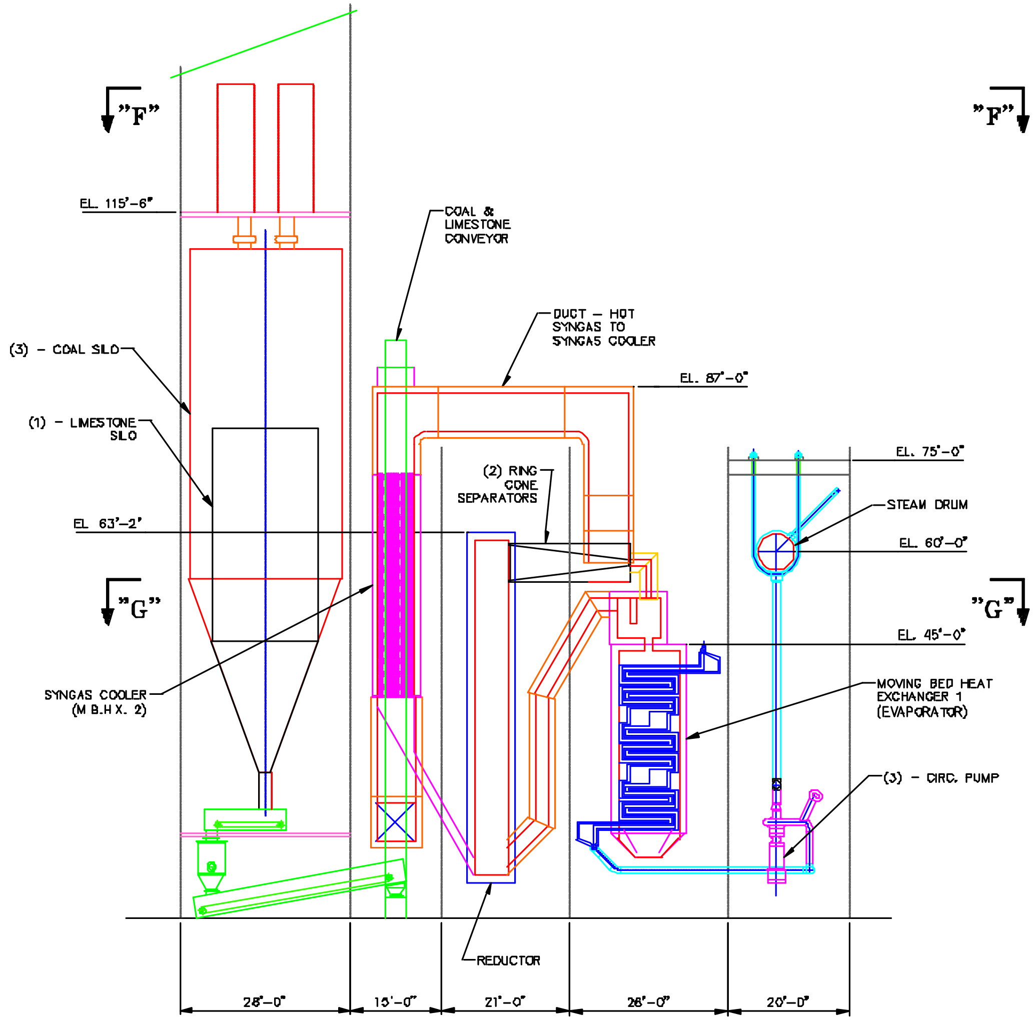


Figure 2.12. 3: Case-12 Gasifier Island General Arrangement Drawing – Side Elevation

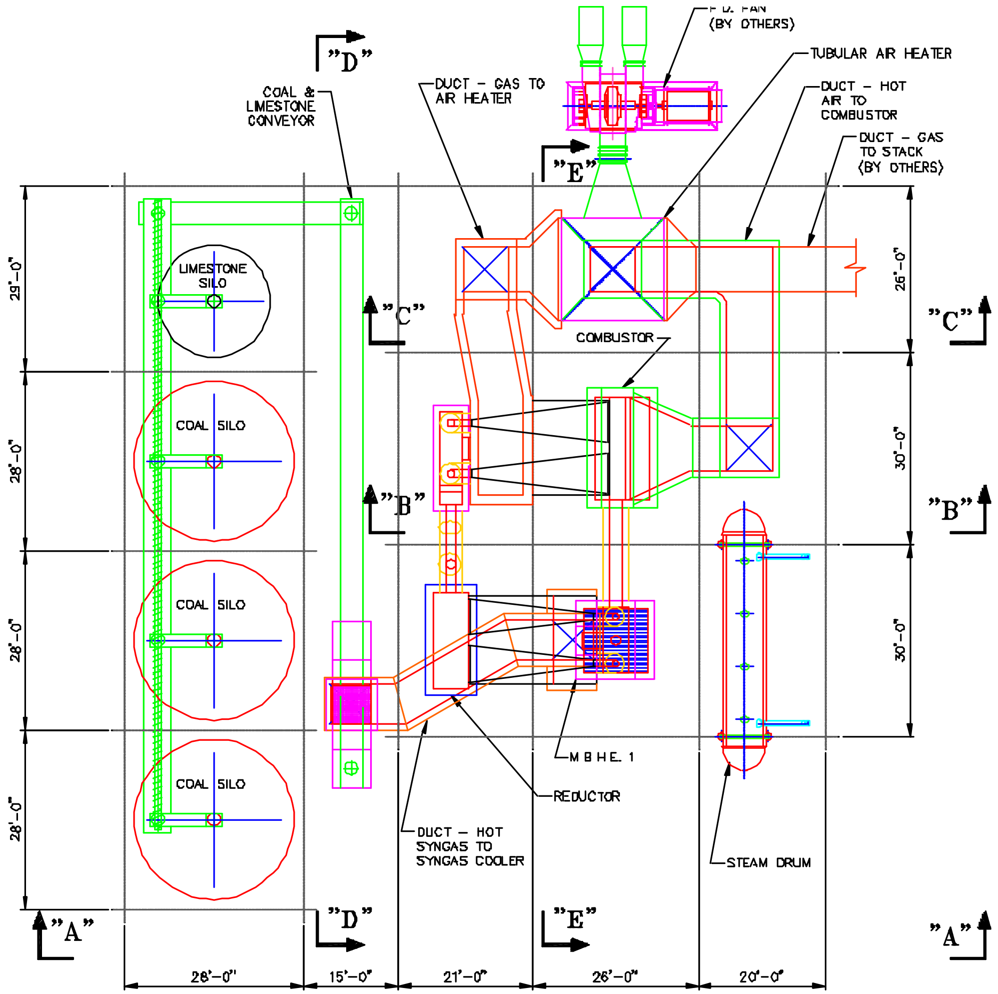


Figure 2.12. 4: Case-12 Gasifier Island General Arrangement Drawing – Plan View

Reducer:

The reducer vessel is designed to sub-stoichiometricly react the oxygen contained in the oxygen carrying solids stream with the feed coal, thus producing a medium-Btu content fuel gas stream, which can be compressed and combusted in a gas turbine. The reducer vessel for Case-12 is about 17 ft wide, 8 ft deep and 56 ft high. Crushed fuel, sorbent, and recycle solids are fed to the lower portion of the reducer.

The reducer vessel is constructed in the same fashion as the Case 7 reducer vessel. It can be described as a rectangular refractory lined vessel with vertical walls. The lower and upper regions are formed with a multilayer refractory liner without any waterwall panels. The lower reducer has penetrations for the admission of fuel, sorbent, and recycle bed material. These penetrations are similar to those used for other cases in this study.

Oxidizer:

The oxidizer is designed to absorb most of the oxygen contained in the incoming air stream. This is done with a supply of oxygen deficient recycle solids (rich in CaS) from the MBHE thus producing an oxygen deficient nitrogen rich flue gas stream and an oxygen rich solids stream (rich in $CaSO_4$) leaving the oxidizer vessel. The oxidizer vessel for Case-12 is about 18 ft wide, 8 ft deep and 52 ft high. Hot air from the air heater, and recycle solids from the MBHE and oxidizer ring cone separator are fed to the lower portion of the oxidizer vessel.

The oxidizer is constructed in the same fashion as the reducer. It can be described as a rectangular refractory lined vessel with vertical walls. The lower and upper regions are formed with a multilayer refractory liner without any waterwall panels. The lower oxidizer vessel includes several penetrations for the admission of hot air and recycled bed material. These penetrations are similar to those used for other cases.

The oxidizer bed temperature is maintained at an optimum level for oxidation efficiency by balancing solids flow between an uncooled stream flowing directly back to the reducer and a cooled stream which flows through the MBHE and then to the oxidizer vessel.

Fuel Feed System:

The fuel feed system for Case-12 is very similar to the system used for other cases in this study. It is designed to transport prepared coal from the storage silos to the lower reducer. The system includes storage silos and silo isolation valves, fuel feeders, feeder isolation valves, and fuel piping connected to the reducer.

Sorbent Feed System:

The limestone feed system for Case-12 is very similar to the system used for other cases in this study. The limestone feed system pneumatically transports prepared limestone from the storage silos to the lower reducer. The system includes the storage silos and silo isolation valves, rotary feeders, blower, and piping from the blower to the reducer injection ports.

Ash Coolers:

The ash cooler design for Case-12 is the same as for the other cases except proportionally larger due to the higher coal throughput. Draining hot solids from the oxidizer through two water-cooled ash coolers controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is provided by feedwater from the steam cycle. The heated water leaving the ash cooler is then combined with water from the economizer located in the HRSG to feed the steam drum.

Particulate Removal:

Fuel gas and entrained solids exit the upper reducer vessel and enter the reducer ring cone separators. Similarly, flue gas and entrained solids exit the oxidizer vessel and enter the oxidizer ring cone separators. These are extremely high efficiency particle separation devices.

Seal Pots:

The seal pots used for Case-12 are of the same design as in other cases. The seal pot is a device that provides a pressure seal between the reducer or oxidizer, which are both at relatively high pressure, and the ring cone separator that is at near atmospheric pressure. The seal pot is a non-mechanical valve, which moves solids collected back to the reducer or oxidizer or other locations. The seal pot is constructed of steel plate with a multiple layer refractory lining with fluidizing nozzles located along the bottom to assist solids flow. Some of the solids flow directly from the seal pot back to the reducer or oxidizer while other solids are diverted through a plug valve to other locations. The diverted solids collected from the reducer flow through the external Moving Bed Heat Exchangers (MBHE) and then back to the oxidizer. The diverted solids collected from the oxidizer flow directly to the reducer.

Convection Pass:

There is no traditional convection pass containing pressure parts in Case-12 and the gas stream leaving the particulate removal system located at the outlet of the oxidizer vessel is simply ducted directly to the High Temperature Tubular Air Heater for heat recovery.

Moving Bed Heat Exchanger:

The external heat exchanger for Case-12 is a single moving bed. The moving bed heat exchanger is not fluidized and contains several immersed tube bundles, which cool the hot solids leaving the reducer seal pot before the cooled solids return to the lower part of the oxidizer. The tube bundles in the MBHE utilize spiral-finned surface and include only a high-pressure evaporator section. Very high heat transfer rates are obtained in the MBHE due to the conduction heat transfer mechanism between the solids and tube. The MBHE is bottom supported and is constructed using steel plate refractory lined enclosure walls. It is rectangular in cross section with a hopper shaped bottom. The solids move through the bed by gravity at a design velocity of about 100 ft/hr. The cooled solids leaving the MBHE are feed to the oxidizer.

Evaporator:

The evaporator section for Case-12 is also located in the MBHE. The evaporator is comprised of three banks of horizontal tubes, which evaporate high-pressure feedwater. The water/steam mixture exiting the evaporator tube banks is supplied to the steam drum through risers where the steam and water phases are separated. The feedwater supplying the evaporator is piped from the steam drum through circulating water pumps and is comprised of a combination of separated saturated water and subcooled water from the HRSG economizer. There is additional evaporator surface located in the HRSG.

Draft System:

The flue gas and fuel gas is moved through the Gasifier Island equipment with the draft system. The draft system includes the combustion air fan, the induced draft (ID) fan, the Fuel Gas Compressor, the associated ductwork, and the expansion joints.

The induced draft and combustion air fan are driven with electric motors and are controlled to operate the oxidizer unit in a balanced draft mode with the oxidizer vessel outlet stream (Stream 10) maintained at a slightly negative pressure (typically, -0.5 inwg).

Similarly, the fuel gas is moved through the reducer vessel with the Fuel Gas Compressor which is controlled to operate the reducer unit in a balanced draft mode with the reducer vessel outlet stream (Stream 15) maintained at a slightly negative pressure (typically, -0.5 inwg).

Air is used for the transport of oxidizer solids while steam and generated syngas are used for the transport of reducer solids.

High Temperature Air Heater:

A tubular regenerative air heater is used to cool the flue gas stream (rich in N_2) leaving the oxidizer particulate removal system by heating the combustion air stream. This is a very high temperature air heater and is considered to be a development item.

2.12.2. Case-12 Balance of Plant Performance and Equipment

This section describes equipment included in the power generation system and other balance of plant equipment including material handling systems, the draft system, the cooling system, electrical systems, and miscellaneous equipment.

2.12.2.1. Power Generation System Performance and Equipment

Gas Turbine Generator:

The gas turbine generator selected for this application is the General Electric MS-7001FA model turbine. There are over 140 GE-7FA and GE-9FA units ordered or in operation. This machine is an axial flow, single spool, constant speed unit, with variable inlet guide vanes. The machine is designed for maximum reliability and efficiency with low maintenance. The turbine includes advanced bucket cooling techniques, compressor aerodynamic design and advanced alloys, enabling a higher firing temperature that the previous generation machines. The standard production version of this machine, fired with natural gas, will develop a compressor pressure ratio of 15.2:1 and a rotor inlet temperature of almost 2,350°F. Power output for the gas turbine is 197,000 kW. The combustion turbine is steam injected for NOx control.

Fuel Gas Compression:

Syngas leaving the Gasifier Island after cooling and mercury removal is compressed to 300 psia, making it suitable for combustion in the gas turbine. The gas is compressed in a multistage intercooled compressor and humidified before going to the combustion turbine. The fuel gas compressor requires about 29,200 kW to compress the full load fuel gas flow.

Steam Cycle Performance:

The steam cycle for Case-12 is shown in a simplified schematic in Figure 2.12.5. This is a double pressure steam cycle with induction steam generated at reheat pressure to optimize heat recovery from the gas turbine exhaust.

For this case, heat is recovered in four (4) locations within the plant. (1) The gas turbine exhaust is cooled in the HRSG, generating high-pressure and intermediate pressure steam. (2) The syngas leaving the Gasifier Island is cooled in the Syngas Cooler, generating high-pressure steam. (3) The Moving Bed Heat Exchanger (MBHE) in the Gasifier Island recovers heat by generating high-pressure steam. (4) The ash cooler within the Gasifier Island.

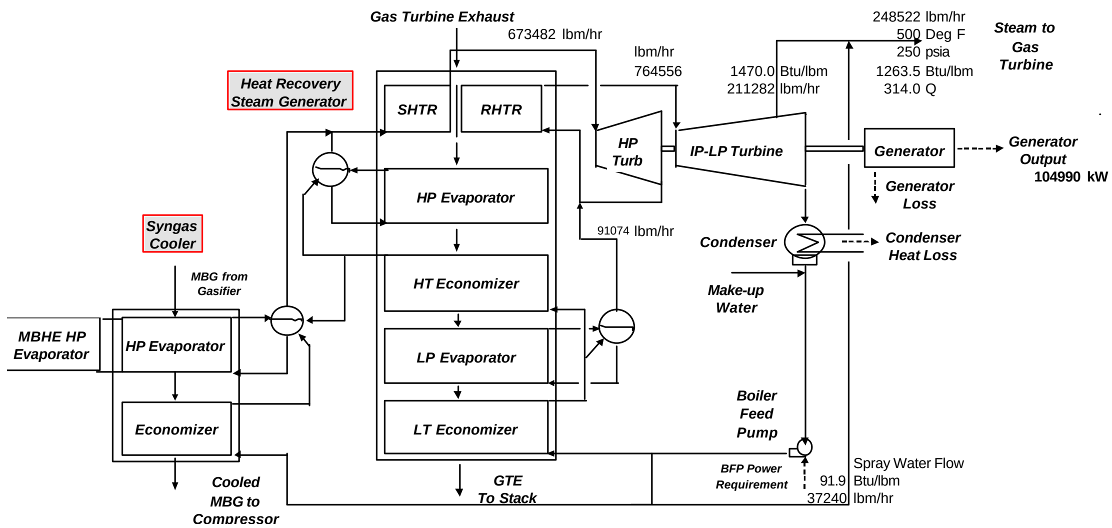


Figure 2.12.5: Case-12 Simplified Steam Cycle Diagram

The steam cycle starts at the condenser hot well, which is a receptacle for the condensed steam from the exhaust of the steam turbine. The condensate flows to the suction of the condensate pumps, which increase the pressure of the fluid by a nominal 250 psi to transport it through the piping system and enable it to enter the open contact heater, or deaerator which is integral within the HRSG.

The condensate entering the deaerator is heated and stripped of noncondensable gases by contact with the steam entering the unit. The steam is condensed and, along with the heated condensate, flows by gravity to a deaerator storage tank. The boiler feedwater pumps take suction from the storage tank and increase the fluid pressure to a nominal 2,200 psig. Both the condensate pump and boiler feed pump are electric motor driven. The boosted condensate flows through the remaining HRSG and Syngas cooler economizer sections.

Within the HRSG, Syngas Cooler and MBHE, the high-pressure feedwater is evaporated. All the evaporated steam is superheated in the HRSG. The high-pressure superheated steam leaving the finishing superheater is expanded through the high-pressure turbine. The high-pressure turbine exhaust is sent to the HRSG for reheating. Additional saturated steam is generated in the HRSG at reheat pressure and is mixed with the HP turbine exhaust before entering the reheater. This combined flow of reheat steam is heated and returned to the intermediate pressure turbine at 405 psia and 1,000 ^{o}F. These conditions (temperatures, pressures) represent common steam cycle operating conditions for power generation systems in use today and are also the same conditions used for all other cases in this study. The reheated steam expands through the intermediate and low-pressure turbines before exhausting to the condenser. The condenser pressure used for Case-12 and all other cases in this study was 3.0 in. Hga.

Nominally, about 250,000 lbm/hr of steam at 250 psia and 500 ^{o}F is supplied to the gas turbine for NOx control. This steam is provided from an extraction point on the LP turbine and sprayed down to 500 ^{o}F with condensate. Makeup water is supplied to the condenser hotwell.

The steam turbine performance analysis results show the generator produces about 105 MW output, and the steam turbine heat rate is 11,416 Btu/kWh. The steam turbine heat rate value may be somewhat misleading, since about 28 percent of the low-pressure turbine inlet steam flow is extracted at 250 psia and used for NO_x control in the gas turbine.

Steam Cycle Equipment:

This section provides a brief description of the major steam cycle equipment (steam turbine, condensate and feedwater systems) utilized for this case.

Steam Turbine:
The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the HRSG passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the HRSG for reheating. The reheated steam flows through the reheat stop valves and intercept valves and enters the IP section at 405 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser. The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for economic unit cycling.

Steam Generation:
Steam is generated in the HRSG and the Gasifier Island in this case. The HRSG is a horizontal gas flow, drum-type, and multi-pressure design that is matched to the characteristics of the gas turbine exhaust gas when firing medium-Btu gas. The HP drum produces steam at main steam pressure, while the IP drum produces steam for export to the cold reheat stream. The HRSG drum pressures are nominally 1,900 and 420 psia for the HP and IP turbine sections, respectively. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater heating, and also provides deaeration of the condensate.

High-pressure steam is also generated within the Gasifier Island in two locations. First, the Syngas Cooler, which cools the MBG leaving the Gasifier prior to fuel gas compression, generates high-pressure steam, which is piped to the HRSG for superheating. Second, a MBHE, used for process cooling within the Gasifier Island, also generates high-pressure steam.

Main and Reheat Steam:
The function of the main steam system is to convey main steam generated in the HRSG and Syngas Cooler from the HRSG superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey cool steam from the HP turbine exhaust to the HRSG reheater, and to convey hot reheated steam from the HRSG to the turbine reheat stop valves.

Main steam at approximately 1,900 psig / 1,000°F exits the HRSG superheater through a motor-operated stop/check valve and a motor-operated gate valve and is routed to the HP turbine. Cold reheat steam at approximately 450 psia / 638°F exits the HP turbine, flows through a motor-operated isolation gate valve, to the HRSG reheater. Hot reheat steam at approximately 420 psig / 1,000°F exits the HRSG reheater through a motor-operated gate valve and is routed to the IP turbine.

Condensate and Feedwater Systems:
The function of the condensate system is to pump condensate from the condenser hot well to the deaerator. The system consists of one main condenser, two 50 percent capacity motor-driven vertical condensate pumps, one gland steam condenser, and one deaerator with a storage tank, which is integral with the HRSG. Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump the various feedwater streams from the deaerator storage tank in the HRSG to the respective steam drums. Two 50 percent capacity boiler feed pumps are provided. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

The feedwater pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing temperature. Feedwater pump suction pressure and temperature are also monitored. In addition, the suction of each boiler feed pump is equipped with a startup strainer. These pumps are electric motor driven.

2.12.2.2. Other Balance of Plant Equipment

The other balance of plant equipment consists of the following areas:

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

Operation Description:
The medium volatile bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading will be done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron, and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1" x 0. The coal is then transferred by conveyor No. 4 to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the two silos.

Technical Requirements and Design Basis:

- Coal burn rate:
- – Maximum coal burn rate = 197,000 lbm/h = 99 tph plus 10 percent margin = 108tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
- – Average coal burn rate = 166,000 lbm/h = 84 tph (based on MCR rate multiplied by an 85 percent capacity factor)
- Coal delivered to the plant by unit trains:
- – Two unit trains per week at maximum or average burn rate
- – Each unit train shall have 10,000 tons (100-ton cars) capacity
- – Unloading rate = 9 cars/hour (maximum)
- – Total unloading time per unit train = 11 hours (minimum)
- – Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
- – Reclaim rate = 300 tph
- – Storage piles with liners, run-off collection, and treatment systems:
- – Active storage = 8,000 tons (72 hours at maximum burn rate)
- – Dead storage = 60,000 tons (30 days at average burn rate)

Table 2.12.3: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	110
Active Storage, tons	8,000
Dead Storage, tons	60,000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,500-ton silo to accommodate 3 days operation.

Ash Handling:

The function of the ash handling system is to convey, prepare, store, and dispose of the bed drain produced on a daily basis by the gasifier. The scope of the system is from the bottom ash hoppers to the truck filling stations.

The bed drain from the gasifier is drained from the bed, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. The silo is sized for a nominal holdup capacity of 36 hours of full-load operation (1,200 tons capacity). At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately

30 truck loads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.12.4: Ash Handling Design Summary

Design Parameter	Value
Bed Drain from Gasifier, lbm/h	77,113
Cooled Ash temperature, °F	520

Draft System:

The following fans, blowers, ductwork and stack provide the draft system for the Gasifier Island:

- Primary air fan:
  This provides forced draft primary airflow to the oxidizer vessel of the chemical looping gasifier system. This fan is a centrifugal type unit, supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.12.5).

Table 2.12.5: Primary Air Fan Specification

Gas Analysis	
Oxygen, wt %	22.89
Nitrogen, wt %	75.83
Water Vapor, wt %	1.28
Carbon Dioxide, wt %	0.00
Sulfur Dioxide, wt %	0.00
Total, wt %	100.00
Operating Conditions	
Mass Flow Rate, lbm/h	644,138
Gas Inlet Temperature, °F	80.0
Inlet Pressure, psia	14.70
Outlet Pressure, psia	16.40
Pressure Rise, in. wg	47.00
Fan Power, kWe	1,170
Motor Horsepower	1,600

- Induced draft fan:
  The ID fan is provided to boost flue gas coming from the oxidizer vessel of the chemical looping gasifier system and flowing out the stack. The ID fan is a centrifugal unit supplied with electric motor drive and inlet damper (see Table).

Table 2.12.6: Induced Draft Fan Specification

Gas Analysis	
Oxygen, wt %	2.61
Nitrogen, wt %	92.64
Water Vapor, wt %	1.57
Carbon Dioxide, wt %	3.18
Sulfur Dioxide, wt %	0.00
Total, wt %	100.00

Operating Conditions	
Mass Flow Rate, lbm/h	531,829
Gas Inlet Temperature, °F	177
Inlet Pressure, in. wg	- 23
Outlet Pressure, psia	17
Pressure Rise, in. wg	40
Fan Power, kWe	740
Motor Horsepower	1,000

- Ducting and Stack:
  One stack is provided with a single 12-foot-diameter steel liner. The stack is constructed of reinforced concrete, with an outside diameter at the base of 40 feet. The stack is 200 feet high.

Table 2.12.7: Stack Design Summary

Design Parameter	Value
Flue Gas Temperature, °F	177
Flue Gas Flow Rate, lbm/h	550,000
Flue Gas Flow Rate, acfm	150,000
Particulate Loading, grains/acfm	nil

Circulating Water System:

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the condenser to condense the main turbine exhaust steam. The system also supplies cooling water to the auxiliary cooling system. A mechanical-draft cooling tower removes the heat transferred from the steam to the circulating water in the condenser.

The system consists of two 50 percent capacity vertical circulating water pumps, a mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counterflow mechanical-draft cooling tower.

The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Waste Treatment:

An onsite water treatment facility will treat all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment will be housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge dewatering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Gasifier building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water will be provided. A 200,000-gallon storage tank will provide a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

All major equipment required for this plant is listed in Appendix I in Section 9.1.12.

Plant layout and plot plan:

The gasification plant layout is arranged functionally to address the flow of material and utilities through the plant site. The site layout drawing is shown in Appendix II Section 9.2.12.

2.12.3. Case-12 Overall Plant Performance and CO_2 Emissions

Overall plant performance and emissions for Case-12 are summarized in Table 2.12.8 and summarized below. The overall system is described previously in Section 2.12.2.

HRSG efficiency is calculated to be 82.77 percent. The steam cycle thermal efficiency including the boiler feed pump debit is about 29.9 percent. This is quite low; however, about 28 percent of the IP/LP steam flow is extracted at 250 psia for gas turbine NOx control.

Total plant auxiliary power is 36,844 kW (about 12 percent of generator output), and the net plant output is about 265 MW.

The net plant heat rate and thermal efficiency are calculated to be 8,248 Btu/kWh and 41.4 percent respectively (HHV basis).

Carbon dioxide emissions are 454,321 lbm/hr or about 1.71 lbm/kWh on a normalized basis.

Table 2.12. 8: Case-12: Overall Plant Performance and Emissions

		Chem Looping w/o CO2 Capture (Case-12)
Auxiliary Power Listing	(Units)	
Induced Draft Fan	(kW)	539
Primary Air Fan	(kW)	1031
Secondary Air Fan	(kW)	n/a
Fluidizing Air Blower	(kW)	n/a
Transport Air Fan	(kW)	n/a
Gas Recirculation Fan	(kW)	n/a
Coal Handling, Preperation, and Feed	(kW)	354
Limestone Handling and Feed	(kW)	209
Limestone Blower	(kW)	157
Ash Handling	(kW)	230
Particulate Removal System Auxiliary Power (baghouse)	(kW)	n/a
Boiler Feed Pump	(kW)	1984
Condensate Pump	(kW)	38
Circulating Water Pump	(kW)	795
Cooling Tower Fans	(kW)	795
Steam Turbine Auxilliaries	(kW)	114
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719
Transformer Loss	(kW)	679
Subtotal	(kW)	7644
	(frac. of Gen. Output)	0.025
Traditional Power Plant Auxiliary Power	(kW)	7644
Air Separation Unit or Fuel Compressor	(kW)	29200
OTM System Compressor Auxiliary Power	(kW)	n/a
CO_2 Removal System Auxiliary Power	(kW)	n/a
Total Auxilary Power	(kW)	36844
	(frac. of Gen. Output)	0.122
Output and Efficiency		
Main Steam Flow	(lbm/hr)	673482
Steam Turbine Heat Rate	(Btu/kwhr)	11416
OTM System Expander Generator Output	(kW)	n/a
Gas Turbine Generator Output		197000
Steam Turbine Generator Output	(kW)	104990
Net Plant Output	(kW)	265146
	(frac. of Case-1 Net Output)	1.37
Boiler Efficiency (HHV)[1]	(fraction)	0.8277
Coal Heat Input (HHV)	(10^6 Btu/hr)	2187
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	2187

[1] Boiler Heat Output / (Qcoal-HHV + Qcredits)
[2] Required for GPS Desicant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4

Net Plant Heat Rate (HHV)	(Btu/kwhr)	8248
Net Plant Thermal Efficiency (HHV)	(fraction)	0.4138
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.17
CO_2 Emissions		
CO_2 Produced	(lbm/hr)	454321
CO_2 Captured	(lbm/hr)	0
Fraction of CO2 Captured	(fraction)	0.00
CO_2 Emitted	(lbm/hr)	454321
Specific CO_2 Emissions	(lbm/kwhr)	1.71
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	0.86
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00

2.13. Case–13: Indirect Gasification of Coal and CO_2 Capture via Chemical Looping

This case provides the CO_2 capture case that is directly comparable to Case-12 (Indirect Gasification of Coal via Chemical Looping). There are two primary chemical loops as well as thermal looping for temperature control utilized in this gasification process.

A chemical looping concept is used to indirectly provide the oxygen for the gasification of coal rather than direct utilization of ambient air. The chemical looping concept supplies the oxygen to the gasification process without the large efficiency penalty associated with a cryogenic type Air Separation Unit (Cases 2, 3, 4, 8, 9, 10, 11). Additionally, the large investment cost associated with both cryogenic Air Separation Units or Oxygen Transport Membrane oxygen supply systems (Case-6) is avoided.

Additionally, CO_2 is captured with a second chemical loop in this concept. These chemical loops provide a very energy efficient method for oxygen transport and CO_2 capture. The trade off, of course, is a more complex gasification process. Through the use of this chemical looping gasification process, Medium Btu Gas (MBG) is produced from an air fired gasifier system.

Power production is provided from a single train GE-7FA gas turbine with a Heat Recovery Steam Generator (HRSG) and an 1,800 psig / 1,000 ^oF / 1,000 ^oF steam cycle. This is the same power production equipment that is used for all other cases in this study.

The overall plant is shown as a block flow diagram in Figure 2.13.1, and Table 2.13.1 shows the key stream conditions. Crushed coal and limestone are supplied to the gasifier system with a conventional feed system. Air is supplied to the gasifier through a primary air fan, and an ID fan takes flue gas from the gasifier and sends it to a stack.

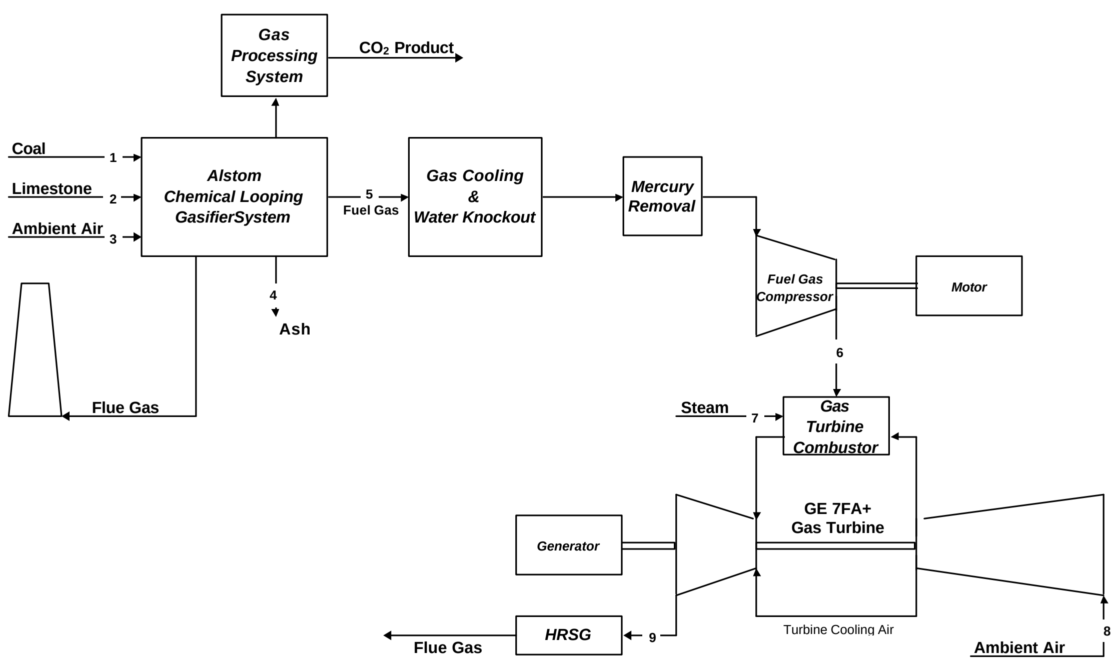


Figure 2.13. 1: Case-13: Simplified Block Flow Diagram

The fuel gas from the gasifier is cooled in a feedwater heater to 100°F to knockout water, and it then continues through a mercury removal bed. The fuel gas is compressed to 300 psia in a multistage compressor and humidified before going to the combustion turbine. The combustion turbine is steam injected for NOx control. Flue gas from the turbine passes through a heat recovery steam generator (HRSG), where high-pressure steam is generated to produce additional power.

The captured CO_2 stream leaving the gasifier is cooled and processed in the Gas Processing System (GPS) to produce a CO_2 stream which is available for use or sequestration.

Table 2.13. 1: Case-13: Overall Plant Stream Report

Mole Fraction	1	2	3	4	5	6	7	8	9
Ar	0.0000	0.0000	0.0094	0.0000	0.0000	0.0000	0.0000	0.0094	0.0079
CO	0.0000	0.0060	0.0000	0.0000	0.0063	0.0060	0.0000	0.0000	0.0000
CO_2	0.0000	0.0001	0.0003	0.0000	0.0001	0.0001	0.0000	0.0003	0.0011
H_2	0.0000	0.9283	0.0000	0.0000	0.9702	0.9283	0.0000	0.0000	0.0000
H_2O	0.0000	0.0656	0.0104	0.0000	0.0100	0.0656	1.0000	0.0104	0.2292
N_2	0.0000	0.0000	0.7722	0.0000	0.0000	0.0000	0.0000	0.7722	0.6479
NH_3	0.0000	0.0000	0.0000	0.0000	0.0135	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.2077	0.0000	0.0000	0.0000	0.0000	0.2077	0.1140
Total	0.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (lb_{mol}/hr)	0	0	24,262	0	16,501	17,244	12,326	112,132	133,646
V-L Flowrate (lb/hr)	0	0	700,077	0	41,987	55,597	222,052	3,235,490	3,513,140
Solids Flowrate (lb/hr)	213,582	32,849	0	83,423	0	0	0	0	0
Temperature (°F)	80	225	80	103	941	225	500	80	1087
Pressure (psia)	14.7	280.0	14.7	63.2	88.2	280.0	250.0	14.7	14.8
Density (lb/ft^3)	---	---	0.07	0.18	0.01	0.12	0.44	0.07	0.02
Average Molecular Weight	---	---	28.85	17.03	2.54	3.22	18.02	28.85	26.29

A brief performance summary for this plant reveals the following results. The Case-13 plant produces a net plant output of about 257 MW. The net plant heat rate and thermal efficiency are calculated to be 9,249 Btu/kWh and 36.9 percent, respectively (HHV basis). Carbon dioxide emissions are about 0.02 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.13.4.

2.13.1. Case 13 Gasifier Island Process Description and Equipment

This section describes the Gasifier Island processes for Case-13 and includes a simplified process flow diagram (PFD), material and energy balance and equipment description. The process and equipment description includes only the major components included in the Gasifier Island.

It should be emphasized that the advanced Chemical Looping gasification and CO_2 capture process described for this case is only conceptual at this time. A significant effort encompasing experimental work related to reaction rates, solids regeneration cycles, and fine particulate removal would be necessary to continue development of this gasification and CO_2 capture process. Additionally, high temperature air heater development work would also be required. Advanced Chemical Looping technology is a technology that shows such promise that ALSTOM has already begun the design of a small "Proof of Concept" pilot-scale facility. Additionally, ALSTOM has responded to a DOE NETL RFP to conduct an extensive test program in this facility (DE-PS26-02NT41613-01).

2.13.1.1. Process Description and Process Flow Diagrams

Figure 2.13.2 shows a simplified process flow diagram for the Case-13 Gasifier Island which utilizes a process for the indirect gasification of coal and CO_2 capture via chemical looping. This process description briefly describes the function of the chemical and thermal loops, major equipment and systems included within the Gasifier Island. Selected mass flow rates (lbm/hr) and temperatures (^oF) are shown on this figure. Complete data for all state points are shown in Table 2.13.2.

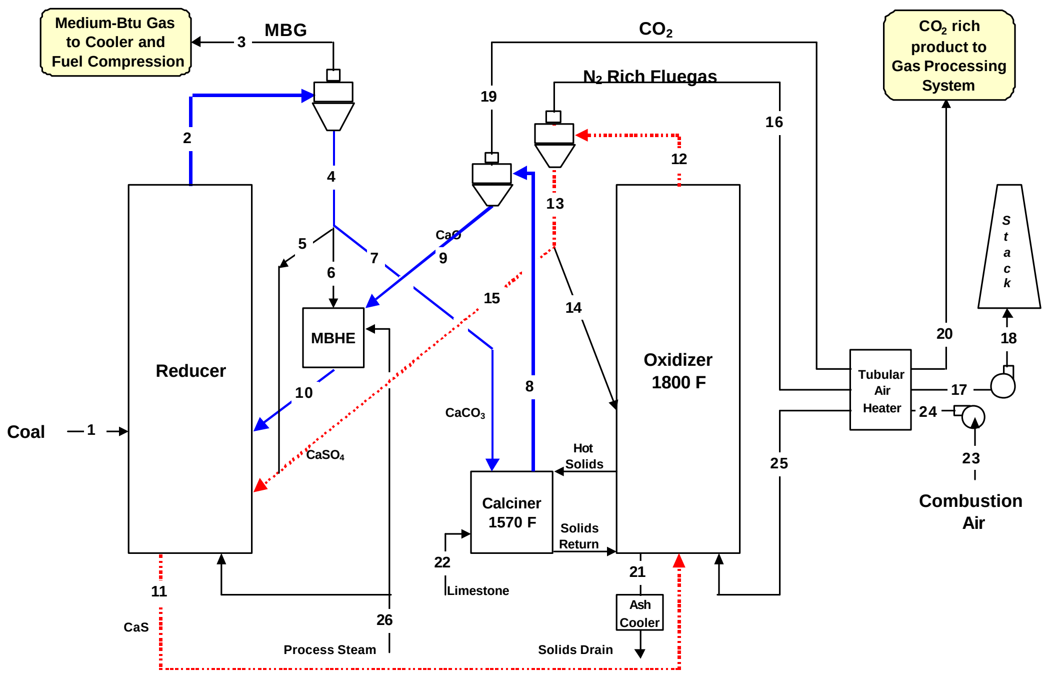


Figure 2.13. 2: Case-13: Simplified Gasifier Island Process Flow Diagram

Three primary reactors are included in the Case-13 processes, the Oxidizer, Reducer, and Calciner. Additionally, there are two primary chemical loops within this process. One loop indirectly supplies oxygen to the reducer for the gasification of the coal. The second loop captures CO_2 from the fuel gas and then releases the captured CO_2 as a second product gas stream, which is compressed and liquefied for sequestration or use. There are also thermal loops included to supply or remove heat as required throughout the processes. The chemical loops will be described first followed by the reactors and other major process equipment and systems.

Oxygen Transport Chemical Loop:

The oxygen transport loop is shown as dashed red lines in Figure 2.13.2 and includes streams 11, 12, 13, and 15. The solids separated from the gas/solids mixture leaving the oxidizer (Stream 13) are rich in $CaSO_4$ and are split into two streams. One stream is transported to the reducer (Stream 15) and the other stream (Stream 14) is recirculated back to the oxidizer. The $CaSO_4$ contained in Stream 15 supplies oxygen to the reducer, where it is reduced to CaS. Stream 11, which is rich in CaS, is returned to the oxidizer to complete the oxygen transport loop.

Limestone (Stream 22) is added to the reducer to react with sulfur contained in the coal. The limestone and the sulfur combine to form CaS in the reducer, as shown below, which is used as the oxygen carrier in the chemical looping reactions described above. Solids are removed from the system in the oxidizer to avoid a buildup of CaS and remove the captured sulfur.

$$Ca + S \rightarrow CaS$$

CO_2 Capture Chemical Loop:

The CO_2 capture chemical loop is shown as bold blue lines in Figure 2.13.2 and includes streams 2, 4, 7, 8, 9 and 10. Starting at the reducer, Stream 10 is a regenerated CaO rich stream that is provided to capture the CO_2 gas that is produced in the reducer. The medium-Btu fuel gas and entrained solids stream leaving the reducer (Stream 2) enter a particulate removal device, where the solids (Stream 4), now rich in $CaCO_3$, are separated from the gas. Stream 4 is then split into three parts with Stream 7 flowing to the calciner. The calciner regenerates CaO from the $CaCO_3$ contained in Stream 7. Heat for this reaction is provided from a stream of hot solids from the oxidizer vessel. The gas solids mixture leaving the calciner (Stream 8) contains the captured CO_2 gas released in the calciner and fine solids that are rich in CaO. This stream enters a particulate removal device where the solids (Stream 9) are separated from the gas (Stream 19). Stream 9 flows through a moving bed heat exchanger and then forms part of Stream 10 to complete the CO_2 capture loop. Stream 19 is the captured CO_2 product stream which is cooled in the tubular air heater and them supplied to the Gas Processing System.

Oxidizer:

The oxidizer is designed to capture oxygen from air utilizing a stream of recirculated bed solids. The bed solids are used as a chemical looping oxygen carrier, whereby the oxygen is picked up by the solids from air in the oxidizer vessel. The oxidizer operates at about 1,800 ^oF. About 90 percent of the oxygen contained in the incoming air is captured by the solids. The basic chemistry in the oxidizer is shown in the following reaction.

$$CaS + 2O_2 \rightarrow CaSO_4 + Heat$$

Therefore, the purpose of the oxidizer is to react the oxygen contained in the air with CaS to form hot $CaSO_4$ with a minimal amount of excess air utilized in the oxidizer. The $CaSO_4$ represents a very effective oxygen carrier due to its high oxygen loading.

Leaving the oxidizer there is a high nitrogen content gas stream (Stream 12), which is cleaned of solids, cooled in a Tubular Air Heater and finally exhausted to the atmosphere through the stack (Stream 18) after passing through Induced Draft (ID) fan.

Draining hot solids (Stream 21) from the oxidizer through water-cooled fluidized bed ash coolers controls solids inventory in the system while recovering heat from the hot ash. The cooling water used to cool the ash in the coolers is feedwater from the steam cycle.

The oxidizer also supplies solids to and receives solids from the Calciner to supply the heat required by the calciner reactions.

Calciner:

The calciner is designed to separate the captured CO_2 from the entering solids stream that are rich in $CaCO_3$ (Stream 7), thereby regenerating the CaO for additional CO_2 capture. The calciner has several streams entering it: (1) a hot solids stream from the oxidizer, (2) a fresh limestone stream (Stream 22), which replaces limestone lost from the system (Stream 21 bed drain), and (3) a stream rich in $CaCO_3$ from the reducer (Stream

7). Two streams leave the calciner: (1) cooled solids stream is returned to the oxidizer and (2) offgas (Stream 8), which contains fine regenerated solids (CaO) that are entrained in the captured CO_2 product.

The calciner is a fluidized bed reactor controlled to operate at about 1,600 °F. The hot solids stream entering the calciner from the oxidizer, at about 1,800 °F, provides the heat required for regeneration of the CaO. Under these conditions the following reaction occurs.

$$CaCO_3 + Heat \rightarrow CaO + CO_2$$

Calciner Particulate Removal:

The CO_2 gas that is released in the calciner (Stream 8) flows through a particulate removal device, where the gas is separated from the entrained solids. Stream 19 (the CO_2 product) then flows to the tubular air heater, where it is cooled by exchanging heat with the incoming combustion air (Stream 24).

The solids leaving the calciner particulate removal device (Stream 9), at about 1,600 °F, are rich in CaO and are piped to the MBHE and then to the reducer (Stream 10). This completes the CO_2 capture solids loop, and the CaO is available to capture more CO_2.

Reducer:

The reducer has several functions to perform. It can be described as a multiple zone reactor. One function of the reducer is to reduce the $CaSO_4$ in the presence of coal, thereby producing a medium-Btu gas. The oxygen carried by the solids ($CaSO_4$ in Stream 15) is reacted substoichiometricly with the carbon and hydrogen contained in the coal (Stream 1) and recycle solids (Streams 5 and 10) in the lower reducer vessel to form a medium-Btu fuel gas (Stream 2). The fuel gas leaves the reducer at about 1,700 °F. Stream 2, fuel gas comprised of primarily H_2 with smaller amounts of CO, CO_2, NH_3 and H_2O vapor, flows through a particulate removal device, where hot solids are removed and recirculated. The principal overall reactions, which are endothermic, are shown below:

$$4C + CaSO_4 + Heat \rightarrow 4CO + CaS$$

$$4H \text{ (in Coal)} + CaSO_4 + Heat \rightarrow CaS + 2H_2O$$

$$H_2O + C + Heat \rightarrow H_2 + CO$$

A second process occurring in the reducer is the shift reaction whereby the following reaction occurs.

$$CO + H_2O \rightarrow H_2 + CO_2$$

This is done to shift most of the carbon into CO_2 for subsequent capture.

A third function of the reducer is to capture CO_2. The CO_2 is captured in the reducer according to the following reaction.

$$CaO + CO_2 \rightarrow CaCO_3 + Heat$$

The medium-Btu fuel gas (Stream 2) leaving the reducer, which is mostly hydrogen, is cleaned of solids, cooled and provides the feed stream (Stream 3) to the Power

Generation System. The fuel is compressed and burned in a combustion turbine, which is part of a combined cycle.

A small quantity of process steam (Stream S1) is introduced into the reducer vessel and the MBHE for purposes of solids activation and to promote the shift reaction.

The temperature in the reducer is controlled to the proper level by splitting the flow of hot recirculated solids leaving the particulate removal system between an uncooled stream (Stream 5) that flows directly back to the reducer and the Moving Bed Heat Exchanger (MBHE), where the solids are cooled before returning to the reducer (Stream 10).

High Temperature Air Heater:

The cooling of the flue gas stream leaving the oxidizer vessel is done in a high temperature tubular air heater where the sensible heat of the high nitrogen content stream is transferred to the incoming combustion air stream (Stream A1).

2.13.1.2. Material and Energy Balance

Table 2.13.2 shows the Gasifier Island material and energy balance for Case-13. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-13 simplified PFD for the Gasifier Island (Figure 2.13.2). This performance was calculated at MCR conditions for this unit.

The MCR condition for Case-13 is defined as operating the system at a condition that fully loads the combustion turbine.

Table 2.13. 2: Case-13 Gasifier Island Material and Energy Balance

		Inputs					Outputs			
	Stream #	1	22	23		26	17	3	20	21
Constituant	(Units)	Coal	Limestone	Primary Air	Infiltration Air	Process Steam	Cold Fluegas	Product Gas	CO_2	Solids Drain
Carbon	(lbm/hr)	132528	0	0	0	0	0	0	0	1636
Hydrogen	"	7625	0	0	0	0	0	32274	0	0
Oxygen	"	6749	0	160216	1602	0	15011	0	0	0
Nitrogen	"	3118	0	530881	5309	0	536243	0	0	0
Sulfur	"	4998	0	0	0	0	0	0	0	0
CO	"	0	0	0	0	0	0	2909	0	0
CO2	"	0	0	0	0	0	18287	42	470298	0
H2O	"	8522	0	8984	90	220742	9075	2971	0	0
NH3	"	0	0	0	0	0	0	3791	0	0
CaCO3	"	0	31206	0	0	0	0	0	0	312
CaO	"	0	0	0	0	0	0	0	0	8567
CaSO4	"	0	0	0	0	0	0	0	0	21223
Ash	"	50042	1642	0	0	0	0	0	0	51685
Total	"	213582	32849	700081	7002	220742	578617	41988	470298	83422
Temp	(Deg F)	80	80	80	80	360	177	941	135	520
Press	(psia)	14.7	14.7	14.7	14.7	150	14.7	88.2	14.7	14.7

2.13.1.3. Gasifier Island Equipment

This section describes major equipment included in the Gasifier Island for Case-13. The major components included in the Gasifier Island include the Reducer vessel, Oxidizer vessel, Calciner Vessel, ash coolers, fuel feed system, fuel silos, sorbent feed system, sorbent silo, particulate removal systems, seal pots, external moving bed heat exchangers (MBHE #1 & #2), evaporator, high temperature air heater, and draft system.

Figures 2.13.3 and 2.13.4 show general arrangement drawings of the Case-13 advanced Chemical Looping Gasifier Island**.** The complete Gasifier Island Equipment List for Case-13 is shown in Appendix I. Appendix II shows several additional drawings of the Gasifier (key plan view, gasifier plan view, side elevation, and various sectional views) as well as the overall plant site plan.

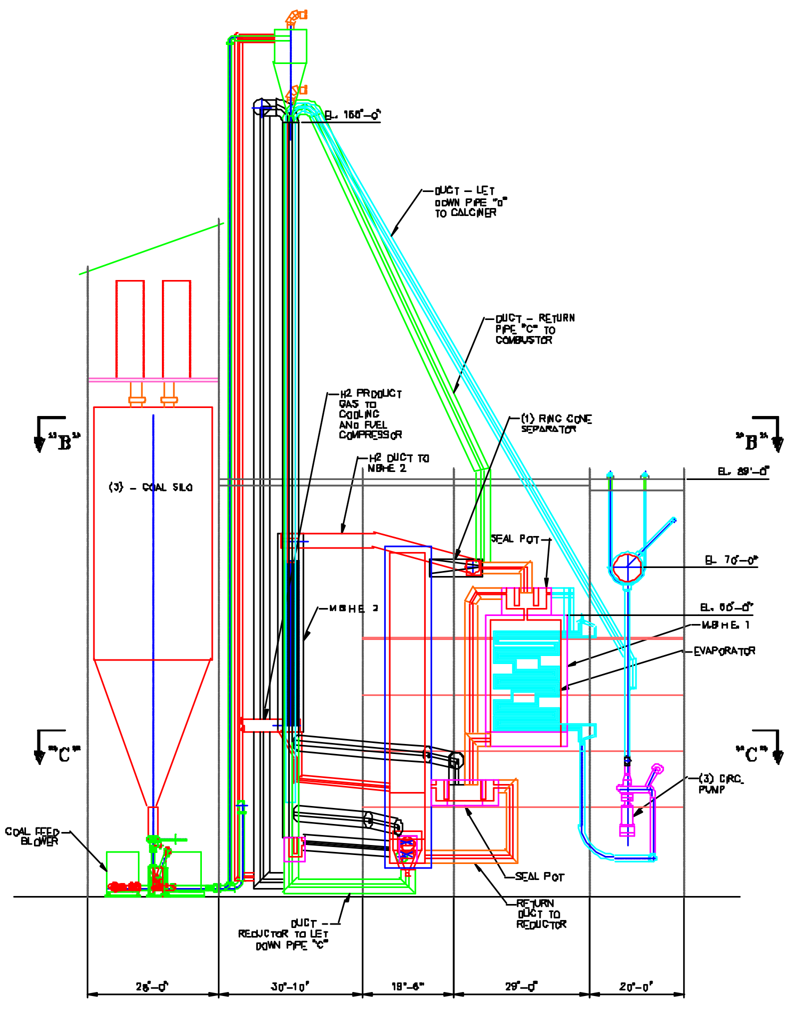


Figure 2.13. 3: Case-13 Gasifier Island General Arrangement Drawing – Side Elevation

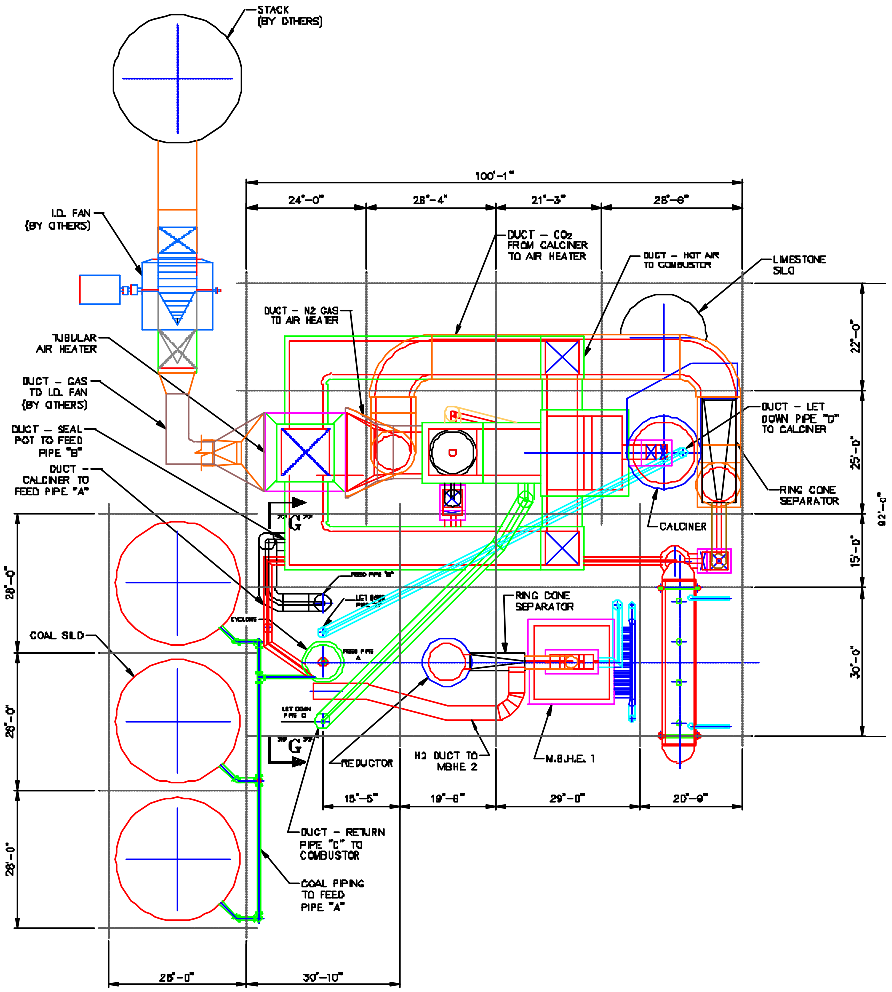


Figure 2.13. 4: Case-13 Gasifier Island General Arrangement Drawing – Plan View

Reducer:

The reducer vessel is designed to sub-stoichiometricly react the oxygen contained in the oxygen carrying solids stream with the feed coal, thus producing a medium-Btu content fuel gas stream which can be compressed an combusted in a gas turbine. Additionally, it is designed to capture CO_2. The reducer vessel for Case-13 is about 10 ft in diameter and about 68 ft high. Crushed fuel and recycle solids are fed to the lower portion of the reducer.

The reducer vessel is constructed in the same fashion as the Case 12 and Case 7 reducers. It can be described as a cylindrical refractory lined vessel with vertical walls. The lower and upper regions are formed with a multilayer refractory liner without any waterwall panels. The lower reducer has penetrations for the admission of fuel, and recycle bed material. These penetrations are similar to those used for other cases in this study.

Oxidizer:

The oxidizer is designed to absorb most of the oxygen contained in the incoming air stream with oxygen deficient recycle solids (CaS) supplied from the MBHE, thus producing an oxygen deficient, nitrogen rich flue gas stream leaving the oxidizer vessel. The oxidizer vessel for Case-13 is about 15 ft wide, 7 ft deep and 38 ft high. Hot air from the air heater and recycle solids from the reducer, calciner and oxidizer particulate removal system, are fed to the oxidizer vessel.

The oxidizer is constructed in the same fashion as the reducer. It can be described as a rectangular refractory lined vessel with vertical walls. The lower and upper regions are formed with a multilayer refractory liner without any waterwall panels. The lower oxidizer vessel includes several penetrations for the admission of hot air and recycle bed material. These penetrations are similar to those used for other cases.

Calciner:

The calciner is designed to separate the captured CO_2 from the entering $CaCO_3$ rich stream thereby regenerating the CaO. The calciner vessel for Case-13 is cylindrical and has an inside diameter of about 12-ft and a height of about 24-ft. This vessel requires penetrations for recycle solids from the particulate removal system and oxidizer as well as a make-up limestone stream.

The calciner is constructed in the same fashion as the reducer vessel. It can be described as a cylindrical refractory lined vessel with vertical walls. It is formed with a multilayer refractory liner without any waterwall panels.

CO_2 and entrained solids exit the calciner vessel and enter the ring cone separator where the hot CO_2 is separated from the fine CaO particles.

Fuel Feed System:

The fuel feed system for Case-13 is very similar to the system used for other cases in this study. It is designed to transport prepared coal from the storage silos to the lower reducer. The system includes the storage silos and silo isolation valves, fuel feeders, feeder isolation valves, and fuel piping to the reducer.

Sorbent Feed System:

The limestone feed system for Case-13 is very similar to the system used for other cases in this study. The limestone feed system pneumatically transports prepared limestone from the storage silos to the calciner. The system includes the storage silos and silo isolation valves, rotary feeders, blower, and piping from the blower to the calciner injection ports.

Ash Coolers:

The ash cooler design for Case-13 is the same as for the other cases except proportionally larger due to the higher coal throughput. Draining hot solids from the oxidizer through two water-cooled ash coolers controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is provided by feedwater from the steam cycle. The heated water leaving the ash cooler is then combined with water from the economizer located in the HRSG to feed the steam drum.

Particulate Removal:

Fuel gas and entrained solids exit the upper reducer vessel and enter the reducer ring cone separators. Similarly, the flue gas and entrained solids exit the oxidizer vessel and enter the oxidizer ring cone separators. Also, the separated CO_2 and entrained solids leaving the calciner vessel, enter the calciner ring cone separators. These ring cone separators are extremely high efficiency particle separation devices.

Seal Pots:

The seal pots used for Case-13 are of the same design as in other cases. The seal pot is a device that provides a pressure seal between the reducer or oxidizer, which is at relatively high pressure, and the ring cone separator that is at near atmospheric pressure. The seal pot is a non-mechanical valve, which moves solids collected back to the reducer or oxidizer or other locations. The seal pot is constructed of steel plate with a multiple layer refractory lining with fluidizing nozzles located along the bottom to assist solids flow. Some of the solids flow directly from the seal pot back to the reducer or oxidizer while other solids are diverted through a plug valve to other locations. The diverted solids collected from the reducer flow through the external Moving Bed Heat Exchangers (MBHE), and then back to the oxidizer. The diverted solids collected from the oxidizer flow directly to the reducer.

Convection Pass:

There is no traditional convection pass containing pressure parts in Case-13 and the gas stream leaving the particulate removal system located at the outlet of the oxidizer vessel is simply ducted directly to the High Temperature Tubular Air Heater for heat recovery.

Moving Bed Heat Exchanger:

There are two external moving bed heat exchangers utilized in Case-13. The moving bed heat exchangers are not fluidized and contain several immersed tube bundles. Very high heat transfer rates are obtained both MBHE’s due to the conduction heat transfer mechanism between the solids and tubes. The MBHE’s are bottom supported and are constructed using steel plate refractory lined enclosure walls. They are rectangular in cross section with a hopper shaped bottoms to facilitate transport of the solids leaving the MBHE. The solids move through the bed by gravity at a design velocity of about 100 ft/hr.

MBHE #1 cools the hot solids leaving the reducer seal pot before the cooled solids return to the lower part of the oxidizer. The tube bundles in MBHE #1 utilize spiral-finned surface and include only a three-bank evaporator section. The cooled solids leaving MBHE #1 are fed to the oxidizer.

MBHE #2 cools the product gas (primarily hydrogen) leaving the reducer while preheating the incoming crushed coal stream. The tube bundles in MBHE #2 are vertically oriented and therefore do not include spiral fins. The heated coal stream leaving MBHE #2 is fed to the reducer.

Evaporator:

The evaporator section for Case-13 is located in MBHE #1. The evaporator is comprised of three banks of horizontal tubes, which evaporate high-pressure feedwater. The water/steam mixture exiting the evaporator tube banks is supplied to the steam drum through risers where the steam and water phases are separated. The feedwater supplying the evaporator is piped from the steam drum through circulating water pumps and is comprised of a combination of separated saturated water and subcooled water from the HRSG economizer.

Draft System:

The flue gas and fuel gas is moved through the Gasifier Island equipment with the draft system. The draft system includes the combustion air fan, the induced draft (ID) Fan, the Fuel Gas Compressor, the associated ductwork, and the expansion joints.

The induced draft and combustion air fan are driven with electric motors and are controlled to operate the oxidizer unit in a balanced draft mode with the oxidizer vessel outlet stream (Stream 10) maintained at a slightly negative pressure (typically, -0.5 inwg).

Similarly, the fuel gas is moved through the reducer vessel with the Fuel Gas Compressor which is controlled to maintain the reducer vessel outlet stream (Stream 3) at about 6 ATM pressure.

High Temperature Air Heater:

A tubular regenerative air heater is used to cool the flue gas stream leaving the oxidizer particulate removal system by heating the combustion air stream. This is a very high temperature air heater and is considered a development item.

2.13.2. Case 13 Gas Processing System Process Description and Equipment

The purpose of this system is to process the CO_2 rich flue gas stream leaving the Case-13 Gasifier Island to provide a liquid CO_2 product stream of suitable purity for an EOR application.

The Case-13 CO_2 capture system is designed for about 95 percent CO_2 capture. Cost and performance estimates were developed for all the systems and equipment required to cool, extract, clean, compress and liquefy the CO_2, to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO_2 specification for EOR, given in Table 2.0.1, was used as the basis for the CO_2 capture system design.

2.13.2.1. Process Description

The following describes a CO_2 recovery system that compresses and then cools a CO_2 rich gas stream from an advanced air-fired CFB type boiler to a pressure high enough so CO_2 can be liquefied. The resulting liquid CO_2 is pumped to a high pressure so it can be economically transported for sequestration or usage. Pressure in the transport pipeline will be maintained above the critical pressure of CO_2 to avoid 2-phase flow.

The key process parameters (pressures, temperatures, duties etc.) are shown in the material and energy balance tables and will not be repeated here except in selected instances.

Figure 2.13.5 shows the Flue Gas Cooling process flow diagram and Figure 2.13.6 shows the Flue Gas Compression and Liquefaction process flow diagram.

Flue Gas Cooling:

Please refer to Figure 2.13.5 (drawing D 12173-13001-0).

The feed to the Gas Processing System is the flue gas stream that leaves the High Temperature Air Heater of the Gasifier Island. At this point, the flue gas is near the dew point of H_2O. All of the flue gas leaving the boiler is cooled to 100 °F in Gas Cooler DA-101 which operates slightly below atmospheric pressure. A significant amount of water condenses out in this cooler. Excess condensate is blown down to the cooling water system. A single vessel has been provided for this cooler.

The Gas Cooler is configured in a packed tower arrangement where the flue gas is contacted with cold water in countercurrent fashion. Warm water from the bottom of the contactor is recycled back to the top of the contactor by Water Pump GA-101 after first cooling it in an external water cooled heat exchanger, Water Cooler EB-101 (plate and frame exchanger). The cooling water for this exchanger comes from the new cooling tower.

Because the flue gas may carry a small amount of fly ash, the circulating water is filtered in Water Filter FD-101A-F to prevent solids build-up in the circulating water. Condensate blowdown is filtered and is taken out downstream of the filter. However, the stream is not cooled and is split off before EB-101. Thus the heat load to the cooling tower is minimized.

From the Gas Cooler the gas stream then is boosted in pressure by the ID fan followed by a split of the gas into two streams. This design was developed to minimize the length of ducting operating at a slight vacuum and to minimize the temperature of the gas being recycled back to the boiler. The mass flow rate of the gas recirculation stream is about 52 percent of the flow rate of the product gas stream, which proceeds to the gas compression area. The recycle stream is sized to provide an oxygen content of about 70 percent by volume in the oxidant stream supplying the boiler. The Gas Cooler minimizes the volumetric flow rate to, and the resulting power consumption of, the Flue Gas Compression equipment located downstream.

Three-Stage Gas Compression System:

Please refer to Figure 2.13.6 (drawing D 12173-13002-0).

The compression section, where the CO_2 rich stream is compressed to 311 psia by a three-stage centrifugal compressor, includes Gas Compressor GB-2301. This three-stage compressor set includes a series of gas coolers (aftercoolers) located after each compression stage. Following the third stage aftercoolers, the stream is then further cooled in a propane chiller to a temperature of –26 °F. Note that both the trim cooling water and water for the propane condenser comes from the cooling tower. Following the compression and liquefaction steps, the pressure of the liquid is boosted to 2,018 psia by CO_2 Pipeline Pump GA-2301. This stream is now available for sequestration or usage.

The volumetric flow to the compressor inlet is about 65,000 ACFM. The discharge pressures of the stages have been balanced to give reasonable power distribution and discharge temperatures across the various stages. They are:

- 1st Stage 40 psia
- 2nd Stage 110 psia
- 3rd Stage 311 psia

Power consumption for this large compressor has been estimated assuming adiabatic efficiency of 75 percent.

The hot gas stream from each compressor stage is first cooled in an air cooler to 120 ^{o}F in Flue Gas Compressor 1st/ 2nd / 3rd Stage Aftercoolers (EC-2301A/B/C, EC-2302A/B, EC-2303). The gas is then further cooled by water-cooled heat exchangers to 95 ^{o}F in Flue Gas Compressor 1st/ 2nd Stage Trim Coolers (EA-2301A/B and EA-2302). The gas compressor's 3rd stage cooler (EA-2303) cools the gas to 90 ^{o}F to reduce the size of the dryers. Due to their large size, many of these heat exchangers consist of multiple shells. Because of highly corrosive conditions, the process side of the coolers must be stainless steel.

Because the flue gas stream leaving the Gasifier Island is nearly saturated, some water condenses out in the three aftercoolers. The sour condensate is separated from the gas in knockout drums (FA-2300/1/2/4) equipped with mist eliminator pads. The condensate from these drums is drained to the cooling tower or to waste water treatment. To prevent corrosion, these drums have a stainless steel liner.

Flue gas leaving the 3rd stage discharge knockout drum (FA-2304) is fed to Flue Gas Drier PA-2351 where additional moisture is removed.

Gas Drying:

Please refer to Figure 2.7.6 (drawing D 12173-13002-0).

It is necessary to dry the CO_2 stream to meet the product specification. A molecular sieve drier has been selected.

The performance of a fixed-bed drier improves as pressure increases. This favors locating the drier at the discharge of the compressor. However, as the operating pressure of the drier increases, so does the design pressure of the equipment. This favors low-pressure operation. But, at low pressure the diameter or number of the drier vessels grows, increasing the cost of the vessel. For this design the drier has been optimally located downstream of the 3rd stage compressor. The CO_2 Drier system consists of four vessels filled with molecular sieve. One vessel is on line while the others are being regenerated. The flow direction is down during operation and up during regeneration.

The drier is regenerated with CO_2 exiting the online driers. After regeneration, heating is stopped while the gas flow continues. This cools the bed down to the normal operating range. The regeneration gas and the impurities contained in it are vented to the atmosphere.

Regeneration of a molecular sieve bed requires relatively high temperature and, because HP steam pressure may fluctuate, a gas-fired heater has been specified for this service.

Flue Gas Filter FD-102 has been provided at the drier outlet to remove any fines that the gas stream may pick up from the desiccant bed.

CO_2 Condensation:

Please refer to Figure 2.13.6 (drawing D 12173-13002-0).
From the CO_2 Drier, the gas stream is cooled down further to -26 ^oF with propane refrigeration in CO_2 Condenser EA-2304A-F.

CO_2 Pumping and CO_2 Pipeline:

Please refer to Figure 2.13.6 (drawing D 12173-13002-0).
The CO_2 product must be increased in pressure to 2,000 psig. A multistage heavy-duty pump (GA-2301) is required for this service. This is a highly reliable derivative of an API-class boiler feed-water pump.

It is important that the pipeline pressure be always maintained above the critical pressure of CO_2 such that single-phase (dense-phase) flow is guaranteed. Therefore, the pressure in the line should be controlled with a pressure controller and the associated control valve located at the destination end of the line.

2.13.2.2. Process Flow Diagrams

Two process flow diagrams are shown below for these systems:

- Figure 2.13.5 (drawing D 12173-13001-0) Flue Gas Cooling PFD
- Figure 2.13.6 (drawing D 12173-13002-0) CO_2 Compression and Liquefaction PFD

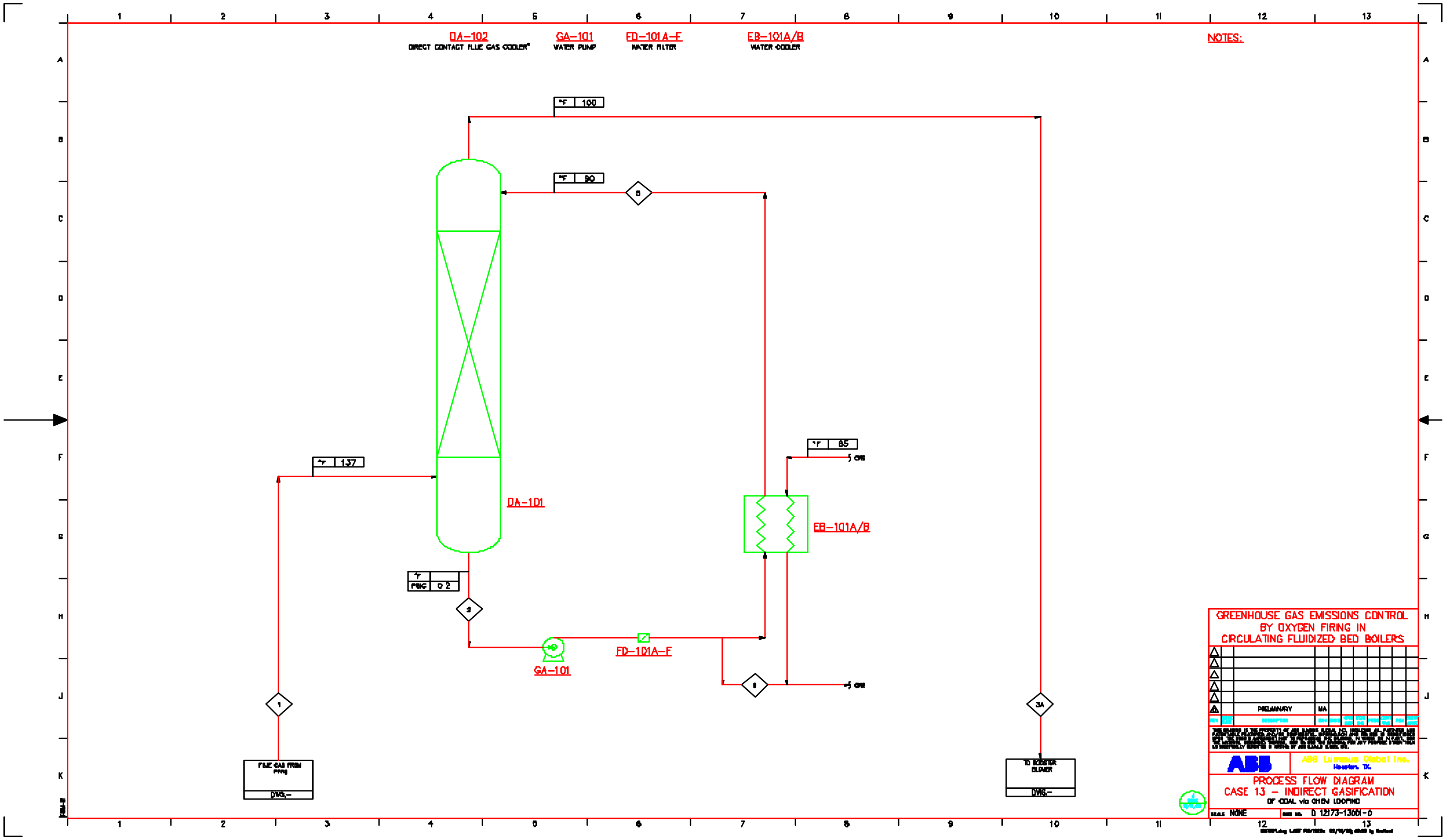


Figure 2.13. 5: Process Flow Diagram for Case-13: Flue Gas Cooling

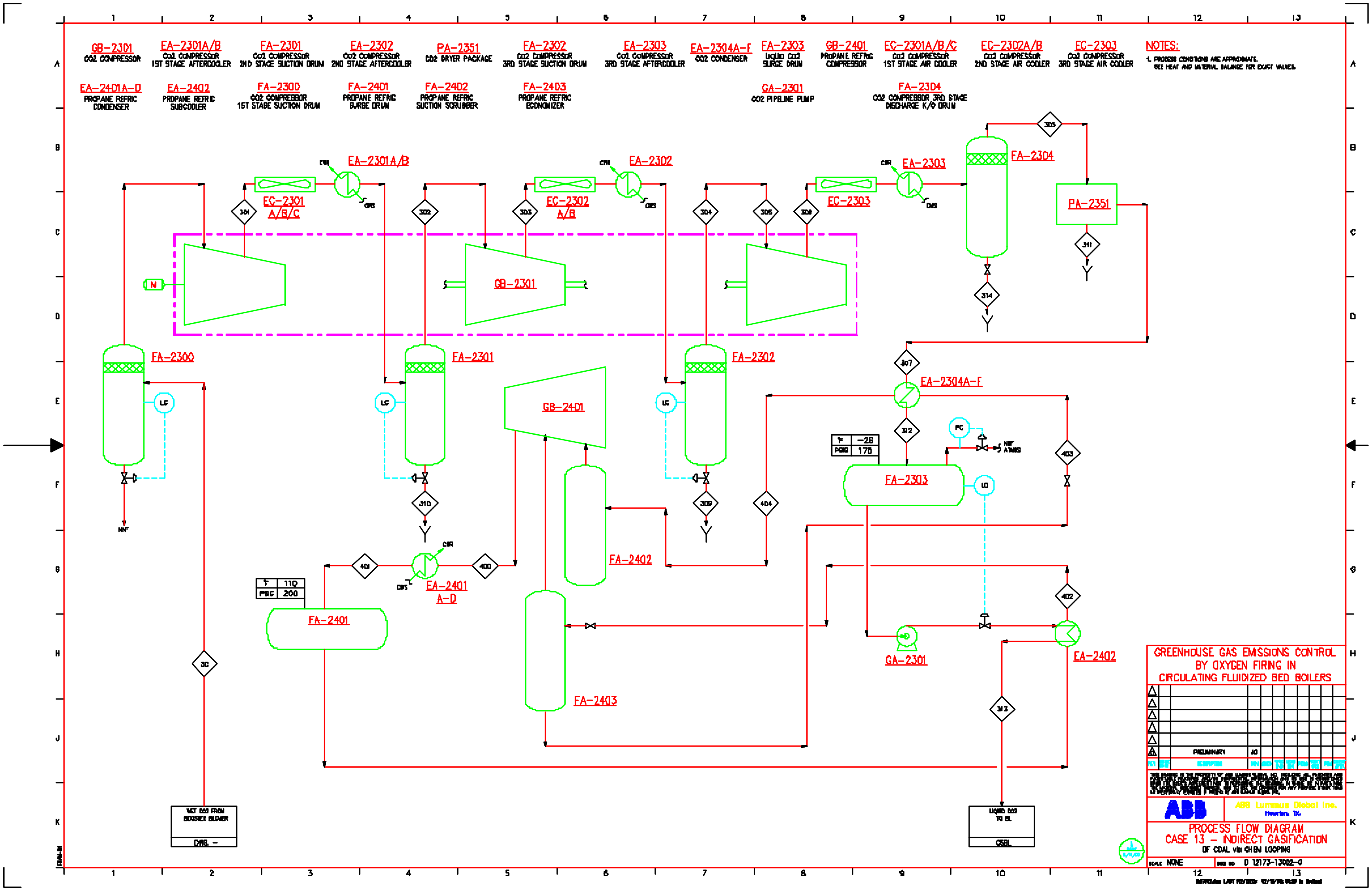


Figure 2.13. 6: Process Flow Diagram for Case-13: CO_2 Compression and Liquefaction

2.13.2.3. Material and Energy Balance

Table 2.13.3 shows the material and energy balance for the Case-13 Gas Processing System.

Table 2.13.3: Case-13 Gas Processing System Material & Energy Balance

STREAM NAME		To quench columns	From Quench columns	Excess water	Quench water out	Quench water in	To liquefaction train	First stage discharge	To second stage	First stage water KO	2nd stage discharge	To 3rd stage	2nd stage water KO
PFD STREAM NO.		**1**	**3a**	**6**	**2**	**5**	**3c**	**301**	**302**	**310**	**303**	**304**	**309**
VAPOR FRACTION	Molar	1.000	1.000	0.000	0.000	0.000	1.000	1.000	1.000	0.000	1.000	1.000	0.000
TEMPERATURE	°F	170.0	100	130	130	90	100	279	95	95	301	95	95
PRESSURE	PSIA	14.7	14	55	14	45	14	40	34	34	110	104	104
MOLAR FLOW RATE	lbmol/hr	20,877	14,255.34	6,614.93	186,195	179,580.3	14,255.63	14,255.63	13,583.67	671.95	13,583.67	13,365.58	218.10
MASS FLOW RATE	lb/hr	718,306	598,952	119,205	3,355,574	3,236,204	598,952	598,952	586,819	12,120	586,819	582,881	3,944
ENERGY	Btu/hr	1.05E+08	6.30E+07	-9.16E+07	-2.58E+09	-2.62E+09	6.34E+07	8.69E+07	5.91E+07	-9.73E+06	8.53E+07	5.70E+07	-3.15E+06
COMPOSITON	Mol %												
CO2		62.76%	91.90%	0.02%	0.02%	0.02%	91.89%	91.89%	96.43%	0.09%	96.43%	98.00%	0.27%
H2O		36.51%	7.03%	99.98%	99.98%	99.98%	7.04%	7.04%	2.45%	99.91%	2.45%	0.86%	99.73%
Nitrogen		0.61%	0.90%	0.00%	0.00%	0.00%	0.90%	0.90%	0.94%	0.00%	0.94%	0.96%	0.00%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		0.12%	0.17%	0.00%	0.00%	0.00%	0.17%	0.17%	0.18%	0.00%	0.18%	0.18%	0.00%
SO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR													
MOLAR FLOW RATE	lbmol/hr	18,102.4	12,360.6	-	-	-	12,360.9	12,360.9	11,778.2	-	11,778.2	11,589.1	-
MASS FLOW RATE	lb/hr	622,834	519,343	-	-	-	519,343	519,343	508,824	-	508,824	505,408	-
STD VOL. FLOW	MMSCFD	164.87	112.57	-	-	-	112.58	112.58	107.27	-	107.27	105.55	-
ACTUAL VOL. FLOW	ACFM	137,905	90,312	-	-	-	84,172	40,585	33,955	-	14,375	10,656	-
MOLECULAR WEIGHT	MW	34.41	42.02	-	-	-	42.02	42.02	43.20	-	43.20	43.61	-
DENSITY	lb/ft³	0.08	0.10	-	-	-	0.10	0.21	0.25	-	0.59	0.79	-
VISCOSITY	cP	0.0131	0.0145	-	-	-	0.0146	0.0199	0.0150	-	0.0214	0.0153	-
LIGHT LIQUID													
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID													
MOLAR FLOW RATE	lbmol/hr	-	-	5,736	161,447	155,712	-	-	-	582.64	-	-	189.11
MASS FLOW RATE	lb/hr	-	-	103,361	2,909,574	2,806,071	-	-	-	10,509.48	-	-	3,419.93
STD VOL. FLOW	BPD	-	-	7,093	199,642	192,551	-	-	-	721	-	-	235
ACTUAL VOL. FLOW	GPM	-	-	209.60	5,900.52	5,592.66	-	-	-	20.98	-	-	6.82
DENSITY	lb/ft³	-	-	61.48	61.48	62.56	-	-	-	62.44	-	-	62.49
VISCOSITY	cP	-	-	0.5043	0.5046	0.7606	-	-	-	0.7185	-	-	0.7503
SURFACE TENSION	Dyne/Cm	-	-	66.91	66.92	70.83	-	-	-	70.30	-	-	70.18

STREAM NAME		From 3rd stage	To drier	3rd stage water KO	From drier / To condenser	Water from drier	From condenser	From product pump	To pipeline	Refrig compressor discharge	From refrig condenser	From subcooler	Refrig to CO2 condenser	Refrig from CO2 condenser
PFD STREAM NO.		**306**	**305**	**314**	**307**	**311**	**312**	**308**	**313**	**400**	**401**	**402**	**403**	**404**
VAPOR FRACTION	Molar	1.000	1.000	0.000	1.000	0.631	0.993	0.000	0.000	1.000	0.000	0.000	0.209	0.993
TEMPERATURE	°F	287	90	90	90	418	-33	-7	82	164	110	49	-33	-33
PRESSURE	PSIA	311	305	305	300	300	19	2,018	2,015	222	215	212	19	19
MOLAR FLOW RATE	lbmol/hr	13,365.58	13,292.01	73.57	13,250.27	41.73	14,481.79	13,250.27	13,250.27	15,409.78	15,409.78	15,409.78	14,481.79	14,481.79
MASS FLOW RATE	lb/hr	582,881	581,543	1,341	580,791	752	638,596	580,791	580,791	679,519	679,519	679,519	638,596	638,596
ENERGY	Btu/hr	8.03E+07	5.26E+07	-1.06E+06	5.25E+07	4.77E+04	6.80E+07	-3.39E+07	-5.51E+06	1.18E+08	1.15E+07	-1.69E+07	-2.21E+07	6.80E+07
COMPOSITON	Mol %													
CO2		98.00%	98.54%	0.80%	98.85%	0.00%	0.00%	98.85%	98.85%	0.00%	0.00%	0.00%	0.00%	0.00%
H2O		0.86%	0.31%	99.20%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.96%	0.97%	0.00%	0.97%	0.00%	0.00%	0.97%	0.97%	0.00%	0.00%	0.00%	0.00%	0.00%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%
Oxygen		0.18%	0.18%	0.00%	0.18%	0.00%	0.00%	0.18%	0.18%	0.00%	0.00%	0.00%	0.00%	0.00%
SO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR														
MOLAR FLOW RATE	lbmol/hr	11,589.1	11,525.3	-	11,489.1	22.8	12,475.0	-	-	13,361.6	-	-	2,623.2	12,475.0
MASS FLOW RATE	lb/hr	505,408	504,248	-	503,596	411	550,108	-	-	589,202	-	-	115,675	550,108
STD VOL. FLOW	MMSCFD	105.55	104.97	-	104.64	0.21	113.62	-	-	121.69	-	-	23.89	113.62
ACTUAL VOL. FLOW	ACFM	4,790.89	3,299.20	-	3,353.74	10.87	48,062	-	-	5,480.35	-	-	10,106.42	48,062
MOLECULAR WEIGHT	MW	43.61	43.75	-	43.83	18.02	44.10	-	-	44.10	-	-	44.10	44.10
DENSITY	lb/ft³	1.76	2.55	-	2.50	0.63	0.19	-	-	1.79	-	-	0.19	0.19
VISCOSITY	cP	0.0216	0.0158	-	0.0158	0.0162	0.0065	-	-	0.0103	-	-	0.0065	0.0065
LIGHT LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	82.00	11,489.14	11,489.14	-	13,361.62	13,361.62	9,933.78	82.00
MASS FLOW RATE	lb/hr	-	-	-	-	-	3,615.81	503,596	503,596	-	589,202	589,202	438,040	3,615.81
STD VOL. FLOW	BPD	-	-	-	-	-	489	41,771	41,771	-	79,626	79,626	59,198	489
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	12.57	929.04	1,362.34	-	2,548.86	2,269.99	1,522.89	12.57
DENSITY	lb/ft³	-	-	-	-	-	35.86	67.58	46.09	-	28.82	32.36	35.86	35.86
MOLECULAR WEIGHT	MW	-	-	-	-	-	44.10	43.83	43.83	-	44.10	44.10	44.10	44.10
VISCOSITY	cP	-	-	-	-	-	0.1849	0.1520	0.0555	-	0.0835	0.1165	0.1849	0.1849
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	14.66	13.32	0.85	-	4.81	8.78	14.66	14.66
HEAVY LIQUID														
MOLAR FLOW RATE	lbmol/hr	-	-	63.79	-	13.35	0.00	-	-	-	-	-	(0.00)	0.00
MASS FLOW RATE	lb/hr	-	-	1,162.48	-	240.57	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	80	-	17	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	2.31	-	0.57	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	62.78	62.78	-	52.31	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	0.7775	0.7775	-	0.1246	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	70.28	70.28	-	34.48	-	-	-	-	-	-	-	-

2.13.2.4. Gas Processing System Utilities

The following tables define the cooling water, natural gas, and electrical requirements for the Gas Processing System.

Table 2.13.4: Case-13 Gas Processing System Cooling Water and Fuel Gas Requirements

COOLING WATER

REV	Equipment TAG NO	SERVICE	No. Installed	DUTY MMBTU/HR	INLET TEMP, F	OUTLET TEMP, F	FLOWRATE LB/HR
	EA-2301	FG Comp 1 stg trim cooler	1	7.69	85	103	427,288
	EA-2302	FG Comp 1 stg trim cooler	1	4.57	85	103	253,895
	EA-2303	FG Comp 1 stg trim cooler	1	4.12	85	103	229,125
	EA-2402	Refrig Condenser	1	86.61	85	100	5,773,955
	EB-101	Water cooler	1	110.35	85	105	5,517,582
		TOTAL COOLING WATER		213.35			12,201,845

FUEL GAS FUEL GAS VALUE BASIS: 930 BTU/SCF (LHV)

REV	Equipment TAG NO	SERVICE	ONLINE FACTOR	COMPR HP	HEAT RATE BTU/HP-HR	DUTY MMBTU/HR	EFFICIENCY %	FLOWRATE (Peak) MMSCFD	FLOWRATE (Peak) SCFH	FLOW (Avg) MMSCFD
	FH-101	Mole sieve regeneration	72%			9.69	80%	0.312	13,019	0.225
		TOTAL FUEL GAS				9.69		0.312	13,019	0.225

Table 2.13.5: Case-13 Gas Processing System Electrical Requirements

Number of Trains	Item Number	Service	Number Operating per train	Power (ea) including 0.95 motor eff	Total all trains
				(kW)	(kW)
1	EC-101	Flue Gas Compressor 1st Stage Aftercooler	1	74	74
1	EC-102	Flue Gas Compressor 2nd Stage Aftercooler	1	62	62
1	EC-103	Flue Gas Compressor 3rd Stage Aftercooler	1	56	56
1	PA-2352	Drier Package	1	426	426
1	GB-101	1 Stage	1	5943	5943
1		2 Stage	1	6621	6621
1		3 Stage	1	5874	5874
1	GB-102	1 stage	1	5761	5761
1		2 stage	1	5458	5458
1	GA-103	CO2 Pipeline pump	1	936	936
		Total			**31209**

2.13.2.5. Gas Processing System Equipment

The equipment list for the Gas Processing System is provided in Appendix I, Section 9.1.13.2.

2.13.3. Case 13 Balance of Plant Performance and Equipment

This section describes equipment included in the power generation system and other balance of plant equipment including material handling systems, the draft system, the cooling system, electrical systems, and miscellaneous equipment.

2.13.3.1. Power Generation System Performance and Equipment

Gas Turbine Generator:

The gas turbine generator selected for this application is the General Electric MS-7001FA model turbine. There are over 140 GE-7FA and GE-9FA units ordered or in operation. This machine is an axial flow, single spool, constant speed unit, with variable inlet guide vanes. The machine is designed for maximum reliability and efficiency with low maintenance. The turbine includes advanced bucket cooling techniques, compressor aerodynamic design and advanced alloys, enabling a higher firing temperature that the previous generation machines. The standard production version of this machine, fired with natural gas, will develop a compressor pressure ratio of 15.2:1 and a rotor inlet temperature of almost 2350°F.

For Case-13 (same as with Cases 9 and 11), with syngas from a plant, the machine requires some modifications to the burner and turbine nozzles in order to properly combust the hydrogen-rich gas and expand the combustion products in the turbine section of the machine. A reduction in rotor inlet temperature of about 50°F results, relative to a production model 7FA machine firing natural gas. This temperature reduction is necessary to not exceed design basis gas path temperatures throughout the expander. Power output for the gas turbine is 197,000 kW. The combustion turbine is also steam injected for NOx control.

Fuel Gas Compression:

Syngas leaving the Gasifier Island after cooling and mercury removal is compressed to 300 psia, making it suitable for combustion in the gas turbine. The gas is compressed in a multi-stage-intercooled compressor and humidified before going to the combustion turbine. The fuel gas compressor requires about 13,080 kW to compress the full load gas flow for this case.

Steam Cycle Performance:

The steam cycle for the Case-13 is shown in a simplified schematic in Figure 2.13.7. This is a double pressure steam cycle with induction steam generated at reheat pressure to optimize heat recovery from the gas turbine exhaust.

For this case heat is recovered in four locations within the plant. (1) The gas turbine exhaust is cooled in the HRSG generating high-pressure and intermediate pressure steam. (2) The syngas leaving the Gasifier Island is cooled prior to compression in the Syngas Cooler generating high-pressure steam. (3) There is a Moving Bed Heat Exchanger (MBHE #1) within the Gasifier Island that recovers heat by generating and superheating high-pressure steam. (4) The CO_2 rich stream leaving the Gasifier Island is cooled prior to compression in the CO_2 Cooler generating high-pressure steam.

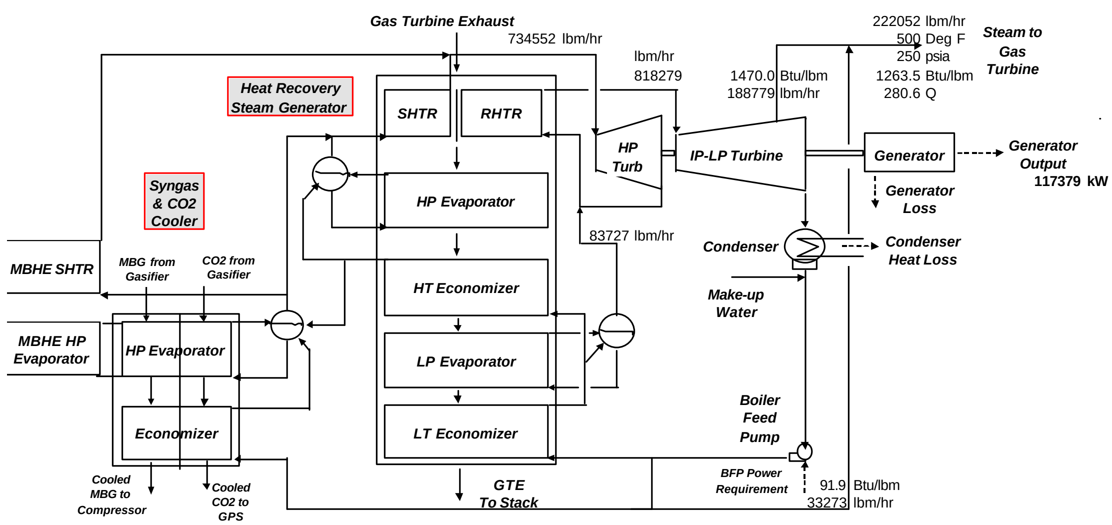


Figure 2.13.7: Case-13 Simplified Steam Cycle Diagram

The steam cycle starts at the condenser hot well, which is a receptacle for the condensed steam from the exhaust of the steam turbine. The condensate flows to the suction of the condensate pumps, which increase the pressure of the fluid by a nominal 250 psi to transport it through the piping system and enable it to enter the open contact heater, or deaerator which is integral within the HRSG.

The condensate entering the deaerator is heated and stripped of noncondensable gases by contact with the steam entering the unit. The steam is condensed and, along with the heated condensate, flows by gravity to a deaerator storage tank. The boiler feedwater pumps take suction from the storage tank and increase the fluid pressure to a nominal 2200 psig. Both the condensate pump and boiler feed pump are electric motor driven. The boosted condensate flows through the remaining HRSG and Syngas cooler economizer sections.

Within the HRSG, Syngas Cooler, CO_2 Cooler, and MBHE #1 the high-pressure feedwater is evaporated. All the evaporated steam is superheated in the HRSG and MBHE #1. The high-pressure superheated steam leaving the finishing superheaters is expanded through the high-pressure turbine. The high-pressure turbine exhaust is sent to the HRSG for reheating. Additional intermediate pressure steam is generated in Th HRSG at reheat pressure and is mixed with the HP turbine exhaust before entering the reheater. Reheat steam is heated and returned to the intermediate pressure turbine at 405 psia and 1,000^{o}F. These conditions (temperatures, pressures) represent common steam cycle operating conditions for power generation systems in use today. The reheated steam expands through the intermediate and low-pressure turbines before exhausting to the condenser. The condenser pressure used for Case-13 and all other cases in this study was 3.0 in. Hga.

Nominally, about 220,000 lbm/hr of steam at 250 psia and 500^{o}F is supplied to the gas turbine for NOx control. This steam is provided from an extraction point on the LP turbine where it is sprayed down to 500 ^{o}F with condensate. Makeup water is supplied to the condenser hotwell.

The steam turbine performance analysis results show the generator produces about 117 MW output and the steam turbine heat rate is 10,947 Btu/kWh. The steam turbine heat rate value is however somewhat misleading since about 23 percent of the low-pressure turbine inlet steam flow is extracted at 250 psia and used for NO_x control in the gas turbine.

Steam Cycle Equipment:

This section provides a brief description of the major equipment (steam turbine, condensate and feedwater systems) utilized for the steam cycle of this case.

Steam Turbine:
The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the HRSG passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheated steam flows through the reheat stop valves and intercept valves and enters the IP section at 405 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser. The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for economic unit cycling.

Steam Generation:
Steam is generated in the HRSG and the Gasifier Island in this case. The HRSG is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the gas turbine exhaust gas when firing medium-Btu gas. The HP drum produces steam at main steam pressure, while the IP drum produces steam for export to the cold reheat. The HRSG drum pressures are nominally 1,900 and 420 psia for the HP and IP turbine sections, respectively. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater heating, and also provides deaeration of the condensate.

High-pressure steam is also generated within the Gasifier Island in three locations. First, the Syngas Cooler, which cools the MBG leaving the Gasifier prior to fuel gas compression, generates high-pressure steam, which is piped to the HRSG for superheating. Second, the CO_2 Cooler, which cools the MBG leaving the Gasifier prior to gas compression, generates high-pressure steam, which is piped to the HRSG for superheating. Finally, there is a MBHE used for process cooling located within the Gasifier Island, which also generates and superheats high-pressure steam.

Main and Reheat Steam:
The function of the main steam system is to convey main steam generated in the HRSG, Syngas Cooler and CO_2 Cooler from the HRSG superheater outlet and MBHE superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey steam from the HP turbine exhaust to the HRSG reheater, and to the turbine reheat stop valves.

Main steam at approximately 1,900 psig / 1,000°F exits the HRSG superheater through a motor-operated stop/check valve and a motor-operated gate valve, and is routed to the HP turbine. Cold reheat steam at approximately 450 psia / 638°F exits the HP turbine, flows through a motor-operated isolation gate valve, to the HRSG reheater. Hot reheat steam at approximately 420 psig / 1,000°F exits the HRSG reheater through a motor-operated gate valve and is routed to the IP turbine.

Condensate and Feedwater Systems:
The function of the condensate system is to pump condensate from the condenser hot well to the deaerator. The system consists of one main condenser; two 50 percent capacity, motor-driven vertical condensate pumps; one gland steam condenser, and one deaerator with a storage tank, which is integral with the HRSG. Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump the various feedwater streams from the deaerator storage tank in the HRSG to the respective steam drums. Two 50 percent capacity boiler feed pumps are provided. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

The feedwater pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing temperature. Feedwater pump suction pressure and temperature are also monitored. In addition, the suction of each boiler feed pump is equipped with a startup strainer.

2.13.3.2. Other Balance of Plant Equipment

The other balance of plant equipment consists of the following areas:

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

Operation Description:
The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading will be done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron, and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1" x 0. The coal is then transferred by conveyor No. 4 to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the two silos.

Technical Requirements and Design Basis:

- Coal burn rate:
- – Maximum coal burn rate = 214,000 lbm/h = 107 tph plus 10 percent margin = 117tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
- – Average coal burn rate = 180,000 lbm/h = 91 tph (based on MCR rate multiplied by an 85 percent capacity factor)
- Coal delivered to the plant by unit trains:
- – Two unit trains per week at maximum or average burn rate
- – Each unit train shall have 10,000 tons (100-ton cars) capacity
- – Unloading rate = 9 cars/hour (maximum)
- – Total unloading time per unit train = 11 hours (minimum)
- – Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
- – Reclaim rate = 300 tph
- – Storage piles with liners, run-off collection, and treatment systems:
- – Active storage = 8,600 tons (72 hours at maximum burn rate)
- – Dead storage = 65,000 tons (30 days at average burn rate)

Table 2.13.6: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	120
Active Storage, tons	8,600
Dead Storage, tons	65,000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,500 ton silo to accommodate 3 days operation.

Ash Handling:

The function of the ash handling system is to convey, prepare, store, and dispose of the bed drain produced on a daily basis by the gasifier. The scope of the system is from the bottom ash hoppers to the truck filling stations.

The bed drain from the gasifier is drained from the bed, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. The silo is sized for a nominal holdup capacity of 36 hours of full-load operation (1,200 tons capacity). At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately 30 truck loads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.13.7: Ash Handling Design Summary

Design Parameter	Value
Ash from Gasifier, lbm/h	83,423
Ash temperature, ^oF	520

Draft System:

The following fans, blowers, ductwork and stack provide the draft system for the Gasifier Island:

- Primary air fan:
  This provides forced draft primary airflow to the gasifier. This fan is a centrifugal type unit, supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.13.8).

Table 2.13.8: Primary Air Fan Specification

Gas Analysis	
Oxygen, wt %	22.89
Nitrogen, wt %	75.83
Water Vapor, wt %	1.28
Carbon Dioxide, wt %	0.00
Sulfur Dioxide, wt %	0.00
Total, wt %	100.00
Operating Conditions	
Mass Flow Rate, lbm/h	700,077
Gas Inlet Temperature, °F	80.0
Inlet Pressure, psia	14.70
Outlet Pressure, psia	16.40
Pressure Rise, in. wg	47.00
Fan Power, kWe	1,250
Motor Horsepower	1,700

- CO_2 Induced draft fan:
  The CO_2 ID fan is provided to boost CO_2 coming from the Gasifier Island and flowing to the Gas Processing System. The CO_2 ID fan is a centrifugal unit supplied with electric motor drive and inlet damper (see Table 2.13.9).

Table 2.13.9: CO_2 Induced Draft Fan Specification

Gas Analysis	
Oxygen, wt %	0.00
Nitrogen, wt %	0.00
Water Vapor, wt %	0.00
Carbon Dioxide, wt %	100.00
Sulfur Dioxide, wt %	0.00
Total, wt %	100.00
Operating Conditions	
Mass Flow Rate, lbm/h	470,295
Gas Inlet Temperature, °F	135
Inlet Pressure, psia	7
Outlet Pressure, psia	14.7
Pressure Rise, in. wg	214
Fan Power, kWe	4,400
Motor Horsepower	6,000

- Induced draft fan:
  The ID fan is provided to boost flue gas coming from the Gasifier Island and flowing out the stack. The ID fan is a centrifugal unit supplied with electric motor drive and inlet damper (see Table 2.13.10).

Table 2.13.10: Induced Draft Fan Specification

Gas Analysis	
Oxygen, wt %	2.61
Nitrogen, wt %	92.64
Water Vapor, wt %	1.57
Carbon Dioxide, wt %	3.18
Sulfur Dioxide, wt %	0.00
Total, wt %	100.00
Operating Conditions	
Mass Flow Rate, lbm/h	578,015
Gas Inlet Temperature, °F	177
Inlet Pressure, in. wg	- 23
Outlet Pressure, psia	17
Pressure Rise, in. wg	40
Fan Power, kWe	790
Motor Horsepower	1,100

- Ducting and Stack:
  One stack is provided for the gasifier island with a single 12-foot-diameter steel liner. The stack is constructed of reinforced concrete, with an outside diameter at the base of 40 feet. The stack is 200 feet high.

Table 2.13.11: Stack Design Summary

Design Parameter	Value
Flue Gas Temperature, ^oF	177
Flue Gas Flow Rate, lbm/h	600,000
Flue Gas Flow Rate, acfm	160,000
Particulate Loading, grains/acfm	nil

Circulating Water System:

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the condenser to condense the main turbine exhaust steam. The system also supplies cooling water to the auxiliary cooling system. The heat transferred from the steam to the circulating water in the condenser is removed by a mechanical draft cooling tower.

The system consists of two 50 percent capacity vertical circulating water pumps, a mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counterflow mechanical draft cooling tower.

The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Waste Treatment:

An onsite water treatment facility will treat all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment will be housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge dewatering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft^2 is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Gasifier building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water will be provided. A 200,000-gallon storage tank will provide a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

All major equipment required for this plant is listed in Appendix I in Section 9.1.13.

Plant layout and plot plan:

The gasification plant layout is arranged functionally to address the flow of material and utilities through the plant site. The site layout drawing is shown in Appendix II Section 9.2.13.

2.13.4. Case-13 Overall Plant Performance and CO_2 Emissions

Overall plant performance and emissions for Case-13 are summarized in Table 2.13.12. The Case-12 (Base Case Chemical Looping) values are also listed along side for comparison purposes.

HRSG efficiency for Case-13 is calculated to be 82.72 percent (HHV basis) as compared to 82.77 percent for Case-12.

The steam cycle thermal efficiency including the boiler feed pump debit is about 31.2 percent as compared to 29.9 percent for Case-12. The slight increase is due to some additional Gasifier island high-pressure steam generation, which is required with the CO_2 capture system.

The net plant heat rate and thermal efficiency for Case-13 are calculated to be 9,249 Btu/kWh and 36.9 percent respectively (HHV basis). This compares to 8,248 Btu/kWh and 41.4 percent respectively for Case-12.

Auxiliary power for Case-13 is 57,548 kW (about 18.3 percent of generator output). The large auxiliary power increase, as compared to Case 12, is due primarily to the large power requirement of the gas compression equipment included in the Gas Processing System of Case-13. Case-12, which does not capture CO_2, does not incur this penalty.

The resulting net plant output for Case-13 is 256,830 kW or about 97 percent of the Case-12 output.

Coal heat input for Case-13 is about 8 percent higher than Case-12.

Carbon dioxide emissions for Case-13 are 6,028 lbm/hr or about 0.02 lbm/kWh on a normalized basis. This represents about 1.5 percent of the Case-12 normalized CO_2 emissions and a CO_2 avoided value of 1.69 lbm/kWh.

Table 2.13. 12: Case-13: Overall Plant Performance and Emissions

	(Units)	Chem Looping w/o CO2 Capture (Case-12)	Chem Looping with CO2 Capture (Case-13)
Auxiliary Power Listing			
Induced Draft Fan	(kW)	539	586
Primary Air Fan	(kW)	1031	1121
Secondary Air Fan	(kW)	n/a	n/a
Fluidizing Air Blower	(kW)	n/a	n/a
Transport Air Fan	(kW)	n/a	n/a
Gas Recirculation Fan	(kW)	n/a	n/a
Coal Handling, Preperation, and Feed	(kW)	354	383
Limestone Handling and Feed	(kW)	209	227
Limestone Blower	(kW)	157	170
Ash Handling	(kW)	230	249
Particulate Removal System Auxiliary Power (baghouse)	(kW)	n/a	n/a
Boiler Feed Pump	(kW)	1984	2236
Condensate Pump	(kW)	38	41
Circulating Water Pump	(kW)	795	1216
Cooling Tower Fans	(kW)	795	1216
Steam Turbine Auxilliaries	(kW)	114	129
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719
Transformer Loss	(kW)	679	707
Subtotal	(kW)	7644	8999
	(frac. of Gen. Output)	0.025	0.029
Traditional Power Plant Auxiliary Power	(kW)	7644	8999
Air Separation Unit or Fuel Compressor	(kW)	29200	13080
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a
CO_2 Removal System Auxiliary Power	(kW)	n/a	35469
Total Auxilary Power	(kW)	36844	57548
	(frac. of Gen. Output)	0.122	0.183
Output and Efficiency			
Main Steam Flow	(lbm/hr)	673482	734552
Steam Turbine Heat Rate	(Btu/kwhr)	11416	10947
OTM System Expander Generator Output	(kW)	n/a	n/a
Gas Turbine Generator Output		197000	197000
Steam Turbine Generator Output	(kW)	104990	117379
Net Plant Output	(kW)	265146	256830
	(frac. of Case-1 Net Output)	1.37	1.33
Boiler Efficiency (HHV)[1]	(fraction)	0.8277	0.8272
Coal Heat Input (HHV)	(10^6 Btu/hr)	2187	2366
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	9.7
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	2187	2375

[1] Boiler Heat Output / (Qcoal-HHV + Qcredits)
[2] Required for GPS Desicant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4

	(Units)	Chem Looping w/o CO2 Capture (Case-12)	Chem Looping with CO2 Capture (Case-13)
Net Plant Heat Rate (HHV)	(Btu/kwhr)	8248	9249
Net Plant Thermal Efficiency (HHV)	(fraction)	0.4138	0.3690
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.17	1.04
CO_2 Emissions			
CO_2 Produced	(lbm/hr)	454321	492600
CO_2 Captured	(lbm/hr)	0	486572
Fraction of CO2 Captured	(fraction)	0.00	0.99
CO_2 Emitted	(lbm/hr)	454321	6028
Specific CO_2 Emissions	(lbm/kwhr)	1.71	0.02
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	0.86	0.01
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.69

3. COST ANALYSIS

The plant capital cost estimate summaries, including engineering, procurement, and construction (EPC basis), are shown in this section for the thirteen (13) power plants included in this study. A detailed cost breakdown for each case is included in Appendix III. The EPC basis does not include owner's costs. Owner's costs are, however, included in the economic analysis (Section 4). The costs are expressed in July 2003 dollars. The level of accuracy for these conceptual level designs is expected to be about +/- 30 percent. These plants are assumed to be constructed on a common Greenfield site in the Gulf Coast region of southeastern Texas.

The boundary limit for these plants includes the complete plant facility within the "fence line". It includes the coal receiving and water supply systems and terminates at the high-voltage side of the main power transformers. Also, for the cases with CO_2 capture, the boundary terminates at the outlet flange of the CO_2 product pipe.

The EPC costs for the seven combustion cases include all required equipment including the traditional Boiler Island equipment, and Balance of Plant equipment (steam turbine, condensate and feedwater, draft system, gas clean-up, material handling, cooling, electrical, instrumentation and control, and misc.). Additionally, for all the CO_2 removal concepts (Cases 2 - 7), the non-traditional equipment required for CO_2 capture, compression and liquefaction is included, and the Air Separation equipment is also included where required (Cases 2, 3, 4, 6). Sections 3.1 – 3.7 show these costs for the seven combustion cases.

The EPC costs for the four Texaco based IGCC cases and the two Chemical Looping gasification cases include all required equipment. This includes the gasifier island equipment, Air Separation equipment (Texaco IGCC cases only), fuel compression equipment (advanced Chemical Looping cases only), combustion turbine equipment and Balance of Plant equipment (steam turbine, condensate and feedwater systems, material handling systems, cooling systems, electrical, instrumentation and control, and misc.). Additionally, for the CO_2 removal concepts (Cases 9, 11 and 13), the non-traditional equipment required for CO_2 capture, compression and liquefaction is included. Sections 3.8 – 3.13 show these costs for the four IGCC and two advanced Chemical Looping cases.

The costs include equipment, materials, labor, indirect construction costs, and engineering. The labor cost to install the equipment and materials was estimated on the basis of labor man-hours. The labor costing approach was a multiple contract labor basis with the labor cost including direct and indirect labor cost plus fringe benefits and allocations for contractor expenses and markup.

These costs include professional services and "other costs". Professional services consist of the cost for engineering, construction management, and startup assistance. The engineering services include all preliminary and detailed engineering and design for the total plant scope. It includes specifying equipment for purchase, procurement, performing project scheduling and cost control services for the project; providing engineering and design liaison during the construction period; and providing startup support. Construction management services cost includes a field management staff capable of performing all field contract administration; field inspection and quality assurance; project construction control; safety and medical services as required; field and construction insurance administration, field office clerical and administrative support. The "other costs" category includes a cost allowance for freight costs, heavy haul, insurance's and taxes, and indirect startup spares.

The capital cost estimates for these plants were calculated based on a combination of vendor-furnished quotes, and cost estimating database values. The Boiler Island costs for Cases 1-7 and the advanced Chemical Looping Island costs for Cases 12 and 13 were estimated based on calculated material weights for all components, conceptual equipment arrangement drawings, and equipment lists, which were developed as a part of the conceptual design of the required equipment.

Operating and Maintenance (O&M) costs are calculated for each plant and are listed as either fixed or variable. The fixed costs are those costs, which are incurred irrespective of the number of hours of plant operation whereas the variable costs are directly proportional to the operating hours. These costs are calculated separately for the traditional power plant equipment, the oxygen supply systems (ASU or OTM), and the Gas Processing Systems (GPS's) where applicable. The O&M costs for the ASU or OTM were calculated by Praxair and the O&M costs for the GPS's were calculated by Lummus.

The O&M costs for the traditional power plant equipment was developed quantitatively by Parsons and ALSTOM. Operating labor cost was calculated based on the number of operator jobs (O.J.) required. The average labor rate used to determine the annual cost was 30.90 $/hr, with a labor burden of 30 percent. The labor administration and overhead cost was assessed at a rate of 25 percent of the O&M labor. Maintenance cost was evaluated as a percentage of the initial capital cost. Maintenance costs for the Combustion Turbines (Cases 8-13) were calculated as a function of the operating hours.

Consumable costs including fuel, limestone, water, and chemicals were determined on the basis of individual flow rates as listed in the material and energy balances, individual unit costs (listed below), and the plant annual operating hours. Waste disposal cost was also based on flow rates from the material and energy balances, unit costs, and operating hours.

- Coal cost - 1.25 $/MM-Btu
- Natural Gas cost - 4.00 $/MM-Btu
- Limestone cost - 10.00 $/Ton
- Water cost - 1.00 $/1,000 gallons
- Water Treatment Chemicals cost - 0.16 $/lbm
- Ash Disposal cost - 8.00 $/Ton

By-product credits were not considered for these cases except in the economic sensitivity study (Section 4.3.3) where a credit for CO_2 product was calculated using values ranging from 0.0 to 20.0 $/Ton of CO_2.

A summary of costs (Capital and O&M) for the thirteen cases are shown in Table 3.0.1. A breakdown of the costs for each case is shown later in this section and a detailed cost breakdown for each case is included in Appendix III. Figure 3.0.1 shows the specific EPC costs ($/kW) for each case.

Table 3.0. 1 EPC Plant Costs and O&M Costs – Summary of all Cases

Study Case	Total Investment Cost, EPC Basis		Operating & Maintenance (O&M) Costs				
			Fixed		Variable @ 80% Capacity		Total, $
	$x1000	$/kW	$	$/kW	$	$/kWh	
Case 1, Air-fired CFB w/o CO_2 Capture	251,804	1,304	5,657,635	29.31	5,587,188	0.0041	11,244,823
Case 2, O_2-Fired CFB w/ASU & CO_2 Capture	328,589	2,443	7,853,885	58.39	8,820,048	0.0094	16,673,933
Case 3, O_2-Fired CFB w/ASU & Flue Gas Sequestration	320,638	2,369	8,060,787	59.55	8,653,810	0.0091	16,714,598
Case 4, O_2-Fired CMB w/ASU & CO_2 Capture	337,402	2,553	7,899,070	59.77	8,889,066	0.0096	16,788,137
Case 5, Air-Fired CFB w/Carbonate Reg. Process & CO_2 Capture	270,232	1,677	5,799,465	35.98	8,264,460	0.0073	14,063,925
Case 6, O_2-Fired CMB w/OTM & CO_2 Capture	468,919	2,375	6,537,784	33.11	10,133,605	0.0073	16,671,389
Case 7, CMB Chemical Looping Combustion w/CO_2 Capture	273,568	1,663	5,797,471	35.25	8,014,747	0.0070	13,812,218
Case 8, Built & Operating IGCC w/o CO_2 Capture	411,731	1,565	10,180,299	38.70	7,745,766	0.0042	17,926,065
Case 9, Built & Operating IGCC w/ CO_2 Capture	502,330	2,179	12,138,670	52.66	9,201,958	0.0057	21,340,627
Case 10, Commercially Offered IGCC w/o CO_2 Capture	341,468	1,451	9,343,766	39.71	6,899,778	0.0042	16,243,544
Case 11, Commercially Offered IGCC w/CO_2 Capture	412,377	2,052	11,067,713	55.06	9,110,706	0.0065	20,178,419
Case 12, Chemical Lopping Gasification w/o CO_2 Capture	296,991	1,120	6,487,709	24.47	5,989,858	0.0032	12,477,567
Case 13, Chemical Lopping Gasification w/ CO_2 Capture	355,132	1,383	7,915,922	30.82	9,888,018	0.0055	17,803,941

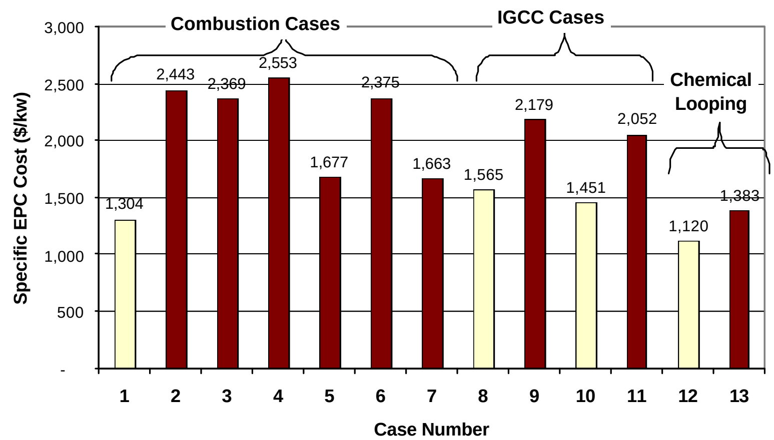


Figure 3.0. 1: EPC Plant Costs – Summary of all Cases

Specific investment costs for the entire spectrum of CO_2 capture cases range from 1,383 – 2,553 $/kW as compared to 1,120 – 1,565 $/kW without capture. Taken as groups, the effect of CO_2 capture on specific investment cost ($/kW-net) were as follows:

- Combustion Cases – $/kW increase ranging from about 28 percent (Cases 5 & 7) to about 96 percent for cryogenic cases 2, 3, and 4.
- IGCC Cases – $/kW increase of about 40 percent for both cases
- Advanced Chemical Looping – $/kW increase of about 23 percent

Overall plant costs and the associated specific plant costs ($/kW) can vary quite significantly for any given plant design depending on several factors. Some of the more important factors are listed below.

- Plant Location and Site Conditions
- Construction Labor Basis
- Coal Analysis
- Ambient Conditions

For the cases in this study, the design coal analysis, design ambient conditions, plant location and site conditions are described in the beginning of Section 2 under Plant

Design Basis. The construction labor basis used is Gulf Coast non-union. The sensitivity of plant specific cost to construction labor basis is indicated by observing that for these studies, changing from Gulf Coast non-union to Ohio River Valley union basis, for example, would increase the EPC plant costs by about 20 percent.

Cost Estimation Basis:

The following assumptions were made in developing the EPC cost estimates for each concept evaluated:

- Investment costs are expressed in July 2003 US dollars
- Construction labor rates are based on Gulf Coast non-union rates
- The plant is constructed on a Greenfield site in southeastern Texas
- All costs are based on mature level (n^{th} plant) commercial design
- Owners costs (including interest during construction, start-up fuel, land, land rights, plant licensing, permits, etc.) are not included in the investment costs but are included in the Cost of Electricity analysis
- Ash is to be shipped off site with provisions for short-term storage only
- Outdoor installation for Gas Processing System (GPS)
- Investment in new utility systems is outside the scope
- CO_2 pipeline is outside the scope
- No special limitations for transportation of large equipment
- No protection against unusual airborne contaminants (dust, salt, etc.)
- No unusual wind storms
- No earthquakes
- No piling required
- All releases can go to atmosphere – no flare provided
- CO_2 Pump designed to API standards, all other pumps conform to ANSI
- All GPS heat exchangers designed to TEMA “C”
- All GPS vessels are designed to ASME Section VIII, Div 1.
- Annual operating time is 7008 h/yr (80 percent capacity factor).
- The investment cost estimate was developed as a factored estimate based on a combination of vendor quotes and in-house data for the major equipment. Such an estimate can be expected to have accuracy of +/-30 percent.
- No purchases of utilities or charges for shutdown time have been charged against the project.

Other exclusions from the EPC investment cost estimate are as follows:

- CO_2 pipeline offsite
- Fuels required for startup
- Relocation or removal of buildings, utilities, and highways
- Permits
- Land and land rights
- Soil investigation
- Environmental Permits
- Disposal of hazardous or toxic waste
- Disposal of existing materials
- Custom's and Import duties
- Sales/Use tax.
- Forward Escalation
- Capital spare parts
- Chemical loading facilities
- GPS Buildings except for Compressor building and electrical substation.

- Financing cost
- Owners costs
- Guards during construction
- Site Medical and Ambulance service
- Cost & Fees of Authorities
- Overhead High voltage feed lines
- Cost to run a natural gas pipeline to the plant
- Excessive piling

Operating and maintenance (O&M) costs were also calculated for all systems. The variable operating and maintenance (VOM) costs for the new equipment included such categories as chemicals and desiccants, waste handling, maintenance material and labor, supplemental fuel usage, and contracted services. The fixed operating and maintenance (FOM) costs for the new equipment includes operating labor only.

3.1. Case-1 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-1 plant (Table 3.1.1) without CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.1.2.

Table 3.1. 1: Case-1 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	14,193	74
2	FUEL & SORBENT PREP. & FEED	5,626	29
3	FEEDWATER & MISC. BOP SYSTEMS	20,361	105
4	FLUIDIZED BED BOILER	79,409	411
5	FLUE GAS CLEANUP	8,781	45
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	13,306	69
8	STEAM TURBINE GENERATOR	37,654	195
9	COOLING WATER SYSTEM	11,835	61
10	ASH/SPENT SORBENT HANDLING SYSTEMS	9,157	47
11	ACCESSORY ELECTRIC PLANT	18,141	94
12	INSTRUMENTATION & CONTROL	11,045	57
13	IMPROVEMENT TO SITE	5,291	27
14	BUILDINGS & STRUCTURES	17,005	88
	TOTAL COST	**251,804**	**1,304**

Table 3.1. 2: Case-1 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

ANNUAL O&M EXPENSES

Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture

Cost Base: Jul-03
Net Plant Heat Rate (Btu/kWh): 9,611
Net Power Output (kW): 193,037
Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	*5,657,635*	*29.31*
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	*5,587,188*	*0.0041*
AIR SEPARATION UNIT (ASU)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	-	0.00
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	-	*0.000000*
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	-	0.00
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	-	*0.0000*
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**5,657,635**	**29.31**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**5,587,188**	**0.00413**

3.2. Case-2 Investment Costs and Operating and Maintenance Cost

The capital cost estimate of the Case-2 plant (Table 3.2.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.2.2.

Table 3.2. 1: Case-2 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	13,955	104
2	FUEL & SORBENT PREP. & FEED	5,526	41
3	FEEDWATER & MISC. BOP SYSTEMS	21,317	158
4	FLUIDIZED BED BOILER & AIR SEP. UNIT	114,291	850
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	60,137	447
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	965	7
8	STEAM TURBINE GENERATOR	37,777	281
9	COOLING WATER SYSTEM	11,869	88
10	ASH/SPENT SORBENT HANDLING SYSTEMS	8,045	60
11	ACCESSORY ELECTRIC PLANT	21,815	162
12	INSTRUMENTATION & CONTROL	10,545	78
13	IMPROVEMENT TO SITE	5,304	39
14	BUILDINGS & STRUCTURES	17,043	127
	TOTAL COST	**328,589**	**2,443**

Table 3.2. 2: Case-2 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture

Net Plant Heat Rate (Btu/kWh): 13,548

Net Power Output (kW): 134,514

Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	5,435,950	40.41
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	5,786,394	0.0061
AIR SEPARATION UNIT (ASU)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	2,111,335	15.70
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	397,253	0.000421
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	306,600	2.28
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	2,636,401	0.0028
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**7,853,885**	**58.39**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**8,820,048**	**0.00936**

3.3. Case-3 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-3 plant (Table 3.3.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.3.2.

Table 3.3. 1: Case-3 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	13,955	103
2	FUEL & SORBENT PREP. & FEED	5,526	41
3	FEEDWATER & MISC. BOP SYSTEMS	21,007	155
4	FLUIDIZED BED BOILER & AIR SEP. UNIT	114,291	844
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	52,534	388
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	965	7
8	STEAM TURBINE GENERATOR	37,777	279
9	COOLING WATER SYSTEM	11,869	88
10	ASH/SPENT SORBENT HANDLING SYSTEMS	8,007	59
11	ACCESSORY ELECTRIC PLANT	21,815	161
12	INSTRUMENTATION & CONTROL	10,545	78
13	IMPROVEMENT TO SITE	5,304	39
14	BUILDINGS & STRUCTURES	17,043	126
	TOTAL COST	**320,638**	**2,369**

Table 3.3. 2: Case-3 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 3 - 1x200 gr. MW O2-Fired CFB w/ASU & Flue Gas Sequestration

Net Plant Heat Rate (Btu/kWh): 13,492

Net Power Output (kW): 135,351

Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	*5,642,852*	*41.69*
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	*5,633,350*	*0.0059*
AIR SEPARATION UNIT (ASU)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	2,111,335	15.60
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	20,000	2.108504E-05
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	306,600	2.27
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	2,623,207	0.002766
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**8,060,787**	**59.55**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**8,276,557**	**0.00873**

3.4. Case-4 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-4 plant (Table 3.4.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.4.2.

Table 3.4. 1: Case-4 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	13,952	106
2	FUEL & SORBENT PREP. & FEED	5,530	42
3	FEEDWATER & MISC. BOP SYSTEMS	21,333	161
4	FLUIDIZED BED BOILER & AIR SEP. UNIT	125,076	946
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	60,011	454
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	972	7
8	STEAM TURBINE GENERATOR	37,793	286
9	COOLING WATER SYSTEM	11,874	90
10	ASH/SPENT SORBENT HANDLING SYSTEMS	8,045	61
11	ACCESSORY ELECTRIC PLANT	21,851	165
12	INSTRUMENTATION & CONTROL	10,549	80
13	IMPROVEMENT TO SITE	5,305	40
14	BUILDINGS & STRUCTURES	17,047	129
	TOTAL COST	**339,338**	**2,567**

Table 3.4. 2: Case-4 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

ANNUAL O&M EXPENSES

Case 4 - 1x200 gr. MW O2-Fired CMB w/ASU & CO2 Capture

Cost Base: Jul-03

Net Plant Heat Rate (Btu/kWh): 13,894

Net Power Output (kW): 132,168

Capacity Factor (%): 80

	Annual Cost, $	Annual Unit Cost
BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	*5,490,815*	*41.54*
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	*5,871,616*	*0.00634*
AIR SEPARATION UNIT (ASU)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	2,111,335	15.97
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	20,000	2.159283E-05
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	306,600	2.32
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	2,629,490	0.002839
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**7,908,750**	**59.84**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**8,521,106**	**0.00920**

3.5. Case-5 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-5 plant (Table 3.5.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.5.2.

Table 3.5. 1: Case-5 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	14,274	89
2	FUEL & SORBENT PREP. & FEED	5,678	35
3	FEEDWATER & MISC. BOP SYSTEMS	20,420	127
4	FLUIDIZED BED BOILER	63,123	392
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	51,382	319
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	9,605	60
8	STEAM TURBINE GENERATOR	36,924	229
9	COOLING WATER SYSTEM	11,630	72
10	ASH/SPENT SORBENT HANDLING SYSTEMS	7,498	47
11	ACCESSORY ELECTRIC PLANT	16,856	105
12	INSTRUMENTATION & CONTROL	10,800	67
13	IMPROVEMENT TO SITE	5,229	32
14	BUILDINGS & STRUCTURES	16,813	104
	TOTAL COST	**270,232**	**1,677**

Table 3.5. 2: Case-5 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.

ANNUAL O&M EXPENSES

Cost Base: Jul-03

Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Case 5 - 1x200 gr. MW Air-Fired CFB w/Carbonate Regeneration Process & CO2 Capture

Net Plant Heat Rate (Btu/kWh): 11,307

Net Power Output (kW): 161,184

Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	*5,492,865*	*34.08*
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	*5,628,059*	*0.0050*
AIR SEPARATION UNIT (ASU)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	-	0.00
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	-	0.0000
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	306,600	1.90
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	2,636,401	0.002334
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**5,799,465**	**35.98**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**8,264,460**	**0.0073**

3.6. Case-6 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-6 plant (Table 3.6.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.6.2.

Table 3.6. 1: Case-6 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	15,987	81
2	FUEL & SORBENT PREP. & FEED	6,377	32
3	FEEDWATER & MISC. BOP SYSTEMS	23,386	118
4	FLUIDIZED BED BOILER & Oxygen Trans Membrane	177,102	897
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	76,627	388
6	COMBUSTION TURBINE ACCESSORIES	27,521	139
7	HRSG DUCTING & STACK	15,921	81
8	STEAM TURBINE GENERATOR	41,116	208
9	COOLING WATER SYSTEM	12,905	65
10	ASH/SPENT SORBENT HANDLING SYSTEMS	8,221	42
11	ACCESSORY ELECTRIC PLANT	28,913	146
12	INSTRUMENTATION & CONTROL	11,295	57
13	IMPROVEMENT TO SITE	5,582	28
14	BUILDINGS & STRUCTURES	17,966	91
	TOTAL COST	**468,919**	**2,375**

Table 3.6. 2: Case-6 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 6 - 1x200 gr. MW O2-Fired CMB w/OTM & CO2 Capture

Net Plant Heat Rate (Btu/kWh): 11,380

Net Power Output (kW): 197,435

Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	*5,879,295*	*29.78*
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	*7,020,308*	*0.0051*
OXYGEN TRANSPORT MEMBRANE (OTM)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	351,889	1.78
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	100,000	7.227395E-05
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	306,600	1.55
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	3,013,297	2.177829E-03
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**6,537,784**	**33.11**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**10,133,605**	**0.00732**

3.7. Case-7 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-7 plant (Table 3.7.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.7.2.

Table 3.7. 1: Case-7 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	13,789	84
2	FUEL & SORBENT PREP. & FEED	5,436	33
3	FEEDWATER & MISC. BOP SYSTEMS	20,786	126
4	FLUIDIZED BED BOILER	68,437	416
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	55,117	335
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	5,191	32
8	STEAM TURBINE GENERATOR	36,938	225
9	COOLING WATER SYSTEM	11,633	71
10	ASH/SPENT SORBENT HANDLING SYSTEMS	7,177	44
11	ACCESSORY ELECTRIC PLANT	16,185	98
12	INSTRUMENTATION & CONTROL	10,844	66
13	IMPROVEMENT TO SITE	5,230	32
14	BUILDINGS & STRUCTURES	16,805	102
	TOTAL COST	**273,568**	**1,663**

Table 3.7. 2: Case-7 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 7 - 1x200 gr. MW CMB Chemical Looping w/CO2 Capture

Net Plant Heat Rate (Btu/kWh): 11,051

Net Power Output (kW): 164,484

Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	*5,490,871*	*33.38*
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	*5,421,500*	*0.0047*
AIR SEPARATION UNIT (ASU)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	-	0
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	-	0
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	306,600	1.86
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	2,593,247	0.00225
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**5,797,471**	**35.25**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**8,014,747**	**0.0070**

3.8. Case-8 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-8 IGCC plant (Table 3.8.1) without CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.8.2.

Table 3.8. 1: Case-8 Total Plant Investment Cost Summary

Account Number	Account Description	Case-8 ($x1000)	Case-8 $/kW
1	Coal Receiving and Handling	15,310	58
2	Coal Preparation and Feed	14,661	56
3	Feedwater & Miscellaneous BOP Systems	15,540	59
4	Gasifier & Accessories	116,250	442
4a	Air Separation Unit	33,992	129
5a	Shift, Gas Cooling, Humidification and Acid Gas Removal	31,232	119
5b	CO2 Compression, Purification, and Liquefaction	0	0
6	Combustion Turbine & Auxiliaries	49,623	189
7	Heat Recovery Boiler & Stack	21,418	81
8	Steam Turbine Generator	22,289	85
9	Cooling Water System	16,030	61
10	Slag Recovery & Handling	23,186	88
11	Accessory Electric Plant	23,672	90
12	I&C	11,371	43
13	Site Improvements	5,309	20
14	Buildings & Structures	11,848	45
	Total Plant Cost	**411,731**	**1,565**

Table 3.8. 2: Case-8 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Case 8 Built & Operating IGCC w/o CO_2 Capture

Cost Base: Jul-03
Net Plant Heat Rate (Btu/kWh): 9,069
Net Power Output (kW): 263,087
Capacity Factor (%): 80

GASIFIER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	9.0	9.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	12.0	12.0

	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	4,222,670	16.05
Maintenance Labor Costs (calc'd)	3,921,569	14.91
Administrative & Support Labor (calc'd)	2,036,060	7.74
TOTAL FIXED OPERATING COSTS	***10,180,299***	***38.70***

	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $	Annual Unit Cost $/kW-net
Maintenace Materisl Cost (calc'd)					4,705,882	0.0026
Consumables						
Water (1000 gallons)		2,926	1.00		854,392	0.0005
Chemicals						
Makeup Chemicals & Catalysts					705,882	0.0004
Subtotal Chemicals					705,882	0.0004
Waste Disposal						
Slag Disposal		631.38	8.00		1,474,904	0.000800
Catalyst Disposal					4,706	0.000003
Subtotal Solid Waste Disposal					1,479,610	0.000803
TOTAL VARIABLE OPERATING COST					***7,745,766***	***0.0042***

3.9. Case-9 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-9 IGCC plant (Table 3.9.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.9.2.

Table 3.9. 1: Case-9 Total Plant Investment Cost Summary

Account		Case-9	
Number	Account Description	($x1000)	$/kW
1	Coal Receiving and Handling	16,961	74
2	Coal Preparation and Feed	16,243	70
3	Feedwater & Miscellaneous BOP Systems	15,554	67
4	Gasifier & Accessories	114,358	496
4a	Air Separation Unit	36,951	160
5a	Shift, Gas Cooling, Humidification and Acid Gas Removal	62,875	273
5b	CO2 Compression, Purification, and Liquefaction	52,418	227
6	Combustion Turbine & Auxiliaries	49,670	215
7	Heat Recovery Boiler & Stack	21,439	93
8	Steam Turbine Generator	22,014	96
9	Cooling Water System	16,526	72
10	Slag Recovery & Handling	25,204	109
11	Accessory Electric Plant	23,562	102
12	I&C	11,381	49
13	Site Improvements	5,314	23
14	Buildings & Structures	11,859	51
	Total Plant Cost	**502,330**	**2,179**

Table 3.9. 2: Case-9 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Case 9 Built & Operating IGCC w/ CO_2 Capture

Cost Base: Jul-03

Net Plant Heat Rate (Btu/kWh): 11,467

Net Power Output (kW): 230,515

Built & Operating IGCC w/ CO_2 Capture

Capacity Factor (%): 80

GASIFIER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	11.0	11.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	14.0	14.0

	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	4,926,449	21.37
Maintenance Labor Costs (calc'd)	4,784,487	20.76
Administrative & Support Labor (calc'd)	2,427,734	10.53
TOTAL FIXED OPERATING COSTS	***12,138,670***	***52.66***
Maintenace Materisl Cost (calc'd)	5,741,384	0.0036

Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $	Annual Unit Cost $/kW-net
Water (1000 gallons)		3,000	1.00		876,000	0.0005
Chemicals						
Makeup Chemicals & Catalysts					941,176	0.0006
Subtotal Chemicals					941,176	0.0006
Waste Disposal						
Slag Disposal		699.48	8.00		1,633,985	0.001011
Catalyst Disposal					9,412	0.000006
Subtotal Solid Waste Disposal					1,643,397	0.001017
TOTAL VARIABLE OPERATING COST					***9,201,958***	***0.0057***

3.10. Case-10 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-10 IGCC plant (Table 3.10.1) without CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.10.2.

Table 3.10. 1: Case-10 Total Plant Investment Cost Summary

Account Number	Account Description	Case-10 ($x1000)	Case-10 $/kW
1	Coal Receiving and Handling	15,619	66
2	Coal Preparation and Feed	13,595	58
3	Feedwater & Miscellaneous BOP Systems	11,921	51
4	Gasifier & Accessories	62,842	267
4a	Air Separation Unit	32,357	138
5a	Shift, Gas Cooling, Humidification and Acid Gas Removal	36,189	154
5b	CO2 Compression, Purification, and Liquefaction	0	0
6	Combustion Turbine & Auxiliaries	53,307	227
7	Heat Recovery Boiler & Stack	18,479	79
8	Steam Turbine Generator	19,193	82
9	Cooling Water System	12,197	52
10	Slag Recovery & Handling	16,843	72
11	Accessory Electric Plant	22,953	98
12	I&C	10,749	46
13	Site Improvements	4,753	20
14	Buildings & Structures	10,471	45
	Total Plant Cost	**341,468**	**1,451**

Table 3.10. 2: Case-10 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Case10 Commercially Offered IGCC w/o CO_2 Capture

Cost Base: Jul-03

Net Plant Heat Rate (Btu/kWh): 9,884

Net Power Output (kW): 235,294

Capacity Factor (%): 80

GASIFIER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	9.0	9.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	12.0	12.0

	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	4,222,670	17.95
Maintenance Labor Costs (calc'd)	3,252,342	13.82
Administrative & Support Labor (calc'd)	1,868,753	7.94
TOTAL FIXED OPERATING COSTS	***9,343,766***	***39.71***

	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $	Annual Unit Cost
Maintenace Materisl Cost (calc'd)					3,902,811	0.0024
Consumables						
Water (1000 gallons)		2,926	1.00		854,392	0.0005
Chemicals						
Makeup Chemicals & Catalysts					705,882	0.0004
Subtotal Chemicals					705,882	0.0004
Waste Disposal						
Slag Disposal		613.008	8.00		1,431,987	0.000868
Catalyst Disposal					4,706	0.000003
Subtotal Solid Waste Disposal					1,436,693	0.000871
TOTAL VARIABLE OPERATING COST					***6,899,778***	***0.0042***

3.11. Case-11 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-11 IGCC plant (Table 3.11.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), and allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.11.2.

Table 3.11. 1: Case-11 Total Plant Investment Cost Summary

Account Number	Account Description	Case-11 ($x1000)	Case-11 $/kW
1	Coal Receiving and Handling	16,795	84
2	Coal Preparation and Feed	14,619	73
3	Feedwater & Miscellaneous BOP Systems	11,976	60
4	Gasifier & Accessories	62,692	312
4a	Air Separation Unit	34,387	171
5a	Shift, Gas Cooling, Humidification and Acid Gas Removal	48,581	242
5b	CO2 Compression, Purification, and Liquefaction	49,587	247
6	Combustion Turbine & Auxiliaries	55,476	276
7	Heat Recovery Boiler & Stack	18,563	92
8	Steam Turbine Generator	18,924	94
9	Cooling Water System	12,254	61
10	Slag Recovery & Handling	17,994	90
11	Accessory Electric Plant	22,838	114
12	I&C	11,461	57
13	Site Improvements	5,066	25
14	Buildings & Structures	11,164	56
	Total Plant Cost	**412,377**	**2,052**

Table 3.11. 2: Case-11 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Case 11 Commercially Offered IGCC w/o CO_2 Capture

Cost Base: Jul-03

Net Plant Heat Rate (Btu/kWh): 12,441

Net Power Output (kW): 201,004

Capacity Factor (%): 80

GASIFIER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90 $/hour
Operating Labor Burden:	30.00 %
Labor O-H change Rate:	25.00 %

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	11.0	11.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	14.0	14.0

	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	4,926,449	24.51
Maintenance Labor Costs (calc'd)	3,927,722	19.54
Administrative & Support Labor (calc'd)	2,213,543	11.01
TOTAL FIXED OPERATING COSTS	***11,067,713***	***55.06***
Maintenace Materisl Cost (calc'd)	4,713,266	0.0033

Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $	Annual Unit Cost $/kW-net
Water (1000 gallons)		3,000	1.00		876,000	0.0006
Chemicals						
Makeup Chemicals & Catalysts					941,176	0.0007
Subtotal Chemicals					941,176	0.0007
Waste Disposal						
Slag Disposal		701.664	8.00		1,639,087	0.001164
Catalyst Disposal					941,176	0.000668
Subtotal Solid Waste Disposal					2,580,264	0.001832
TOTAL VARIABLE OPERATING COST					***9,110,706***	***0.0065***

3.12. Case-12 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-12 Chemical Looping Gasification plant (Table 3.12.1) without CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), and allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.12.2.

Table 3.12. 1: Case-12 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	16687	63
2	FUEL & SORBENT PREP. & FEED	6578	25
3	FEEDWATER & MISC. BOP SYSTEMS	8729	33
4	CHEMICAL LOOPING GASIFIER	53266	201
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	0	-
6	COMBUSTION TURBINE AND FUEL COMP.	102332	386
7	HRSG DUCTING & STACK	24433.208	92
8	STEAM TURBINE GENERATOR	15513.12	59
9	COOLING WATER SYSTEM	5235.3	20
10	ASH/SPENT SORBENT HANDLING SYSTEMS	8682.96	33
11	ACCESSORY ELECTRIC PLANT	22657.6	85
12	INSTRUMENTATION & CONTROL	10842	41
13	IMPROVEMENT TO SITE	5232	20
14	BUILDINGS & STRUCTURES	16803	63
	TOTAL COST	**296,991**	**1,120**

Table 3.12. 2: Case-12 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 12 - CMB Chemical Looping Gasification w/o CO2 Capture

Net Plant Heat Rate (Btu/kWh): 8,248

Net Power Output (kW): 265,146

Capacity Factor (%): 80

	Annual Cost, $	Annual Unit Cost
GASIFIER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	*8,814,236*	*33.24*
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	*8,223,325*	*0.0044*
AIR SEPARATION UNIT (ASU)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	-	0
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	-	0
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	-	0.00
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	-	0.00000
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**8,814,236**	**33.24**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**8,223,325**	**0.0044**

3.13. Case-13 Investment Costs and Operating and Maintenance Costs

The capital cost estimate of the Case-13 Chemical Looping Gasification plant (Table 3.13.1) with CO_2 capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), and allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 3.13.2.

Table 3.13. 1: Case-13 Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		$x1000	$/kW
1	FUEL & SORBENT HANDLING	18,021	70
2	FUEL & SORBENT PREP. & FEED	7,105	28
3	FEEDWATER & MISC. BOP SYSTEMS	10,127	39
4	FLUIDIZED BED BOILER	63,572	248
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	64,880	253
6	COMBUSTION TURBINE ACCESSORIES	77,716	303
7	HRSG DUCTING & STACK	24,433	95
8	STEAM TURBINE GENERATOR	17,994	70
9	COOLING WATER SYSTEM	5,236	20
10	ASH/SPENT SORBENT HANDLING SYSTEMS	9,380	37
11	ACCESSORY ELECTRIC PLANT	23,790	93
12	INSTRUMENTATION & CONTROL	10,843	42
13	IMPROVEMENT TO SITE	5,230	20
14	BUILDINGS & STRUCTURES	16,805	65
	TOTAL COST	**355,132**	**1,383**

Table 3.13. 2: Case-13 Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 13 - CMB Chemical Looping Gasification w/ CO2 Capture

Net Plant Heat Rate (Btu/kWh): 9,249

Net Power Output (kW): 256,830

Capacity Factor (%): 80

	Annual Cost, $	Annual Unit Cost
GASIFIER ISLAND AND BALANCE OF PLANT O&M COSTS		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	*9,613,728*	*37.43*
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	*8,431,762*	*0.0047*
AIR SEPARATION UNIT (ASU)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	-	0
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE O&M COSTS	-	0
GAS PROCESSING SYSTEM (GPS)		
	Annual Cost, $	Annual Unit Cost, $/kW
TOTAL FIXED O&M COSTS	306,600	1.19
	Annual Cost, $	Annual Unit Cost, $/kWh
TOTAL VARIABLE OPERATING COST	3,380,486	0.00188
TOTAL PLANT O&M COSTS		
	Annual Cost, $	**Annual Unit Cost, $/kW**
TOTAL FIXED OPERATING COSTS	**9,920,328**	**38.63**
	Annual Cost, $	**Annual Unit Cost, $/kWh**
TOTAL VARIABLE OPERATING COST	**11,812,248**	**0.0066**

4. ECONOMIC ANALYSIS

A comprehensive economic evaluation comparing the various CO_2 capture concepts with the appropriate Base Case study unit (without CO_2 capture) was performed. These comparisons were done for three types of plants (Combustion, IGCC, and advanced Chemical Looping) and therefore multiple Base Cases were used. An economic sensitivity analysis was also completed.

The purpose of the evaluation was to quantify the impact of CO_2 capture on the Cost of Electricity (COE) of new Greenfield coal fired plants including Combustion, IGCC, and advanced Chemical Looping type units. Additionally a comparison between all cases is also provided. The economic evaluation results are presented as both total and incremental Costs of Electricity (levelized basis). The incremental costs of electricity of the new plants are incremental relative to the appropriate Base Case (Case 1 for the Combustion concepts, Cases 8 and 10 for the Texaco IGCC concepts, and Case 12 for the advanced Chemical Looping concept). CO_2 mitigation costs were also determined in this analysis for each case.

The model used to perform the economic evaluations was the proprietary ALSTOM Power Plant Laboratories' Project Economic Evaluation Pro-Forma. This cash flow model, developed by the Company's Project & Trade Finance group, has the capability to analyze the economic effects of different technologies based on differing efficiencies, investment costs, operating and maintenance costs, fuel costs, and cost of capital assumptions. Various categories of results are available from the model. In addition to cost of electricity, net present value, project internal rate of return, payback period, and other evaluation parameters are available.

4.1. Economic Analysis Assumptions

Numerous financial assumptions were required in performing the economic evaluations. These assumptions are listed in Table 4.1.1. Three technology types, with slightly different financial assumptions, were identified for this analysis. Cases 1, 2, 3, 4, and 6 were included in the CFB and CMB technology type. Cases 5 and 7 were included in the Chemical Looping technology type. Cases 8, 9, 10, 11, 12 and 13 were included in the gasification technology type. The parameters that vary among the three technology types are the construction period, owner's initial spares/consumables, and the funding drawdown during the construction period. All other financial inputs are equivalent for all cases.

The 30-month construction period of the CFB systems is known from the experience in the industry. The construction period for the CMB systems is thought to be similar to the CFB systems since the system complexity is very similar. The CFB/CMB cases include Cases 2, 3, 4, and 6. The 36 month construction period for the air-fired CFB with carbonate regeneration (Case-5) and CMB Chemical Looping combustion (Case-7) systems is slightly longer because these technologies are thought to be somewhat more complex (similar to today's advanced coal-fired systems). Finally, the gasification systems (Cases 8 - 13) are significantly more complex and, as such, will have a longer construction period of 48 months (Holt, 2000).

Furthermore, the complexity of a gasification system will require the owner to have more spare and consumable equipment than what is required of today's advanced coal-fired systems (Holt, 2000). Therefore, the percent of EPC cost assumed for this category is increased from 1 percent for CFB/CMB and CMB Chemical Looping systems to 3 percent for gasification systems.

Table 4.1.1 summarizes the primary technical and financial assumptions used in the model for the three technologies: CFB/CMB, chemical looping, and gasification. Items that are indicated as "Case Sensitive" are discussed in the corresponding case study section of this report. Items shaded in yellow represent parameters that were varied in the economic sensitivity study (Section 4.3.3).

Table 4.1.1: Economic Evaluation Study Assumptions

	Technology Type		
	CFB and CMB Systems - Cases 1, 2, 3, 4, 6	**Chemical Looping Systems - Cases 5, 7**	**Gasification Systems - Cases 8, 9, 10, 11, 12 , 13**
POWER GENERATION			
Net output (MW)	Case Sensitive	Case Sensitive	Case Sensitive
Capacity factor (%)	100%	100%	100%
Availability factor (%)	80%	80%	80%
Net plant heat rate, HHV (Btu per kWh)	Case Sensitive	Case Sensitive	Case Sensitive
Degradation factor (%)	0.0%	0.0%	0.0%
TIME FRAME			
Construction period (months)	30	36	48
Depreciation Term (years)	30	30	30
Analysis Horizon (years)	30	30	30
PROJECT COSTS			
EPC Price ($1000s)	Case Sensitive	Case Sensitive	Case Sensitive
Fixed O&M costs ($ per kW)	Case Sensitive	Case Sensitive	Case Sensitive
Variable O&M costs (cents per kWh)	Case Sensitive	Case Sensitive	Case Sensitive
Owner's EPC Contingency	0.0%	0.0%	0.0%
Initial spares and consumables	1.0%	1.0%	3.0%
Insurance			
Insurance during Construction	1.0%	1.0%	1.0%
Insurance during first year of operation	0.5%	0.5%	0.5%
Development Costs			
Development Costs & Fees	4.0%	4.0%	4.0%
Reimburseable Dev't Costs	3.0%	3.0%	3.0%
Advisory Fees	3.0%	3.0%	3.0%
Financial and Legal Fees	3.0%	3.0%	3.0%
Start-up Fuel	0.5%	0.5%	0.5%
Fuel Stock Pile	0.5%	0.5%	0.5%
Other Costs	0.5%	0.5%	0.5%
Total Initial Project Costs (% of EPC)	17.0%	17.0%	19.0%
FUEL COST			
Coal Price ($ per MMBtu)	1.25	1.25	1.25
Natural Gas Price ($ per MMBtu)	4.00	4.00	4.00
PROJECT CREDITS			
CO_2 Sell Price ($ per ton)	0.00	0.00	0.00
N_2 Sell Price ($ per ton)	0.00	0.00	0.00
ESCALATION FACTORS			
Coal Price	0.0%	0.0%	0.0%
Variable O&M	0.0%	0.0%	0.0%
Fixed O&M (including payroll)	0.0%	0.0%	0.0%
Consumer Price Index	0.0%	0.0%	0.0%
FINANCING ASSUMPTIONS			
Equity	50.0%	50.0%	50.0%
Debt	50.0%	50.0%	50.0%

	Technology Type		
	CFB and CMB Systems - Cases 1, 2, 3, 4, 6	**Chemical Looping Systems - Cases 5, 7**	**Gasification Systems - Cases 8, 9, 10, 11, 12 , 13**
DEBT PORTFOLIO			
Interest Rates (Financed) [1]			
During Construction			
Base Rate	1.32%	1.32%	1.32%
Swap/Reinvestment cushion	1.28%	1.28%	1.28%
Fixed Rate Margin	3.00%	3.00%	3.00%
All-In Fixed Rate	5.60%	5.60%	5.60%
During Operation			
Base Rate	1.32%	1.32%	1.32%
Swap/Reinvestment cushion	1.28%	1.28%	1.28%
Fixed Rate Margin	2.50%	2.50%	2.50%
All-In Fixed Rate	5.10%	5.10%	5.10%
Up-front Fee (Financed)	2.0%	2.0%	2.0%
Commitment Fee	1.0%	1.0%	1.0%
Grace Period (months)	0	0	0
Loan Tenor (years after construction)	30	30	30
TAXES			
Corporate Tax	20.0%	20.0%	20.0%
Tax holiday (years after commissioning)	0.0%	0.0%	0.0%
Customs Duty	0.0%	0.0%	0.0%
Customs Clearance Fee	0.0%	0.0%	0.0%
COST OF CAPITAL ASSUMPTIONS			
Discount Factor	10.0%	10.0%	10.0%
PROGRESS PAYMENT SCHEDULES			
Month			
1	10%	10%	10%
8	15%		
10		15%	7.5%
16			7.5%
17	25%		
20		25%	
22	20%		
25			12.5%
26	20%	20%	
30	10%		
31		20%	
32			12.5%
36		10%	20%
41			20%
48			10%
Total	100%	100%	100%

[1] **Wall Street Journal, 4/23/03, London Interbank Offered Rate (LIBOR) Swap Curve**

4.2. Cost of Electricity Calculation

Levelized cost of electricity (COE) was used as one criterion to compare the systems in this study. The cost of electricity result consists of the following components: financial, fixed O&M, variable O&M, and fuel. The cash flow model is structured to calculate the corresponding annual cash flows for each of these items over the evaluation life of the project. The annual expenses are distributed over the corresponding net annual electricity generated (kWh/year) in order to determine a unit cost (cents/kWh). These costs are subsequently levelized to get a corresponding value of each component over the plant life. In other words, each of the cash flow streams are converted to annuity payments corresponding to a constant value over the life of the study.

The **financial** component of the COE represents the costs which are associated with payment of the original engineered, procured and constructed (EPC) price, all associated owner's costs, custom's and financing fees, and interests accrued both during construction and during operation. The **fixed O&M** component represents the costs that occur regardless of whether the unit is in operation or not. The **variable O&M** component represents the incremental costs which occur when the unit is in operation. The **fuel** cost component represents the cost of the fuel, which is consumed by a given technology.

4.3. Economic Analysis Results

The economic analysis results of the combustion systems (Cases 1 - 7) are discussed in Section 4.3.1. The gasification and advanced Chemical Looping technology systems (Cases 8 - 13) are discussed in Section 4.3.2. The case studies are compared using several evaluation criterion including levelized cost of electricity (COE), incremental COE with respect to the appropriate reference plant without CO_2 capture, and by mitigated costs of avoided CO_2.

The equation for the incremental COE is defined as:

$$Incremental\,COE = \left(COE_{CP} - COE_{ref}\right)$$

Where:

COE = levelized Cost of Electricity (cents / kWh),
CP = Capture Plant, and
ref = Reference Plant.

The equation for the mitigation cost is defined as:

$$Mitigation\,Cost = \frac{COE_{CP} - COE_{ref}}{CO2_{ref} - CO2_{CP}}$$

Where:

Mitigation Cost = \$/ton of CO_2 Avoided,
COE = levelized Cost of Electricity (\$ / kWh),
CO_2 = Carbon dioxide emitted (ton / kWh),
CP = Capture Plant, and
ref = Reference Plant.

The reference plants (Case-1 for the Combustion systems, Cases 8 and 10 for the IGCC systems and Case-12 for the advanced Chemical Looping systems) for the CO_2 capture technologies are included in Tables 4.3.1 and 4.3.2. The incremental COE and mitigation cost results for the CO_2 capture technologies are provided in Tables 4.3.1 and 4.3.2 as well and discussed in the Economics Study Summary and Conclusions section.

4.3.1. Combustion Cases

The levelized COE for the combustion based systems (Cases 1 - 7) are summarized in Table 4.3.1. The air-fired CFB system (Case-1) is the reference plant against which the combustion-based systems with CO_2 capture (Cases 2 - 7) are compared. The levelized COE for Case 1 is about 4.5 cents/kWh. All economic evaluation criterion, levelized COE, incremental COE and mitigated CO_2 costs, indicate Case 7 – Chemical Looping combustion and Case 5 – Air-fired CFB with carbonate regeneration have the lowest production costs of the combustion system technologies studied. These cases (5 and 7) showed increases in levelized COE of slightly less than 30 percent. The cryogenic cases (2, 3, and 4) showed increases in levelized COE of about 76 – 85 percent. Case 6 (OTM) falls in-between with an increase in levelized COE of about 58 percent.

Table 4.3. 1: Combustion-Based Systems (Cases 1, 2, 3, 4, 5, 6, and 7) – Economic Analysis Summary.

	Case 1 - Air Fired CFB w/o CO_2 Capture	Case 2 - O_2 Fired CFB w/ ASU & CO_2 Capture	Case 3 - O_2 Fired CFB w/ ASU & Flue Gas Sequestration	Case 4 - O_2 Fired CMB w/ ASU & CO_2 Capture	Case 5 - Air Fired CFB w/ Carbonate Regneration & CO_2 Capture	Case 6 - O_2 Fired CMB w/ OTM & CO_2 Capture	Case 7 - Chemical Looping Combustion w/ CO_2 Capture
Levelized Cost of Electricity at 80% Availability Factor (cents / kWh)							
Financial	2.5	4.7	4.5	4.9	3.3	4.4	3.3
Fixed O&M	0.4	0.8	0.8	0.9	0.5	0.5	0.5
Variable O&M	0.4	1.0	0.9	1.0	0.7	0.7	0.7
Fuel	1.2	1.7	1.7	1.7	1.4	1.4	1.4
Total	4.5	8.3	8.0	8.4	5.9	7.1	5.8
Incremental Cost of Electricity at 80% Availability Factor (cents / kWh)	Reference Plant	3.7	3.4	3.9	1.4	2.5	1.3
Mitigated Cost ($ / ton CO2 avoided at 80% Availability Factor)	Reference Plant	41	35	43	14	27	13

4.3.2. Gasification Cases

The levelized COE values for the IGCC and advanced Chemical Looping systems (Cases 8 through 13) are summarized in Table 4.3.2. The reference plants (Cases 8, 10, and 12) are indicated with the respective gasification systems (built and operating IGCC, commercially offered gasification, and chemical looping gasification) to which the respective gasification systems with CO_2 capture are compared against. Both incremental COE and mitigated cost analyses indicate Case 13 – advanced Chemical Looping gasification has the lowest production costs of the gasification systems. The IGCC cases (9 and 11) showed increases in levelized COE of about 36 and 38 percent respectively. The advanced Chemical Looping case (13) showed an increase in levelized COE of about 21 percent.

Table 4.3. 2: Gasification Systems (Cases 8, 9, 10, 11, 12, and 13) – Economic Analysis Summary.

	Texaco Built and Operating IGCC		Texaco Commercially Offered IGCC		ALSTOM Chemical Looping Gasification	
	Case 8 - w/o CO_2 Capture	Case 9 - w/ CO_2 Capture	Case 10 - w/o CO_2 Capture	Case 11 - w/ CO_2 Capture	Case 12 - w/o CO_2 Capture	Case 13 - w/ CO_2 Capture
Levelized Cost of Electricity at 80% Availability Factor (cents / kWh)						
Financial	3.2	4.4	3.0	4.2	2.3	2.9
Fixed O&M	0.6	0.8	0.6	0.8	0.5	0.6
Variable O&M	0.4	0.6	0.4	0.6	0.4	0.7
Fuel	1.1	1.4	1.2	1.6	1.0	1.2
Total	5.3	7.2	5.2	7.2	4.3	5.2
Incremental Cost of Electricity at 80% Availability Factor (cents / kWh)	Reference Plant for Case 9	1.9	Reference Plant for Case 11	2.0	Reference Plant for Case 13	0.9
Mitigated Cost ($ / ton CO2 avoided at 80% Availability Factor)	Reference Plant for Case 9	23	Reference Plant for Case 11	23	Reference Plant for Case 13	11

4.3.3. Sensitivity Study Results

Sensitivity analyses were conducted for all case studies to determine the effect on COE of variation of selected base parameter values by ± 25 percent and CO_2 by-product selling price up to $20 per ton. These parameters (shaded in yellow in Table 4.1.1) are capacity factor, EPC price, coal price, CO_2 credit sell price, equity rate, corporate tax rate, and the discount rate for cost of capital. The base parameter values represent the point where all the sensitivity curves intersect (point 0, 0). Selected sensitivity analysis “spider plots” for selected Cases 2, 9, and 13 are provided in the following section. The complete package of sensitivity results for all case studies are provided in Appendix IV.

In general, for the variable ranges studied, CO_2 selling price, capacity factor, plant investment cost, and discount rate, in order of decreasing significance, have the greatest effect on the COE. The discount rate becomes more significant as the construction period increases as observed by comparing between the combustion cases (30-month construction period) versus the IGCC and advanced Chemical Looping cases (48-month construction period).

4.3.3.1. Case 2 - Oxygen-Fired CFB with Air Separation Unit and CO2 Capture

Results for the Case 2 COE sensitivity study are shown in Figure 4.3.1. The tabulated results for Case 2 are provided in Appendix IV. The levelized COE for the base parameter values is 8.3 cents per kWh. Levelized COE ranges from a low of 5.6 to a high of 10.1 cents per kWh. CO_2 mitigation costs ranged from 12 to 62 $/ton of CO_2 avoided (reference plant is Case 1) with the baseline value at 41 $/ton of CO_2 avoided.

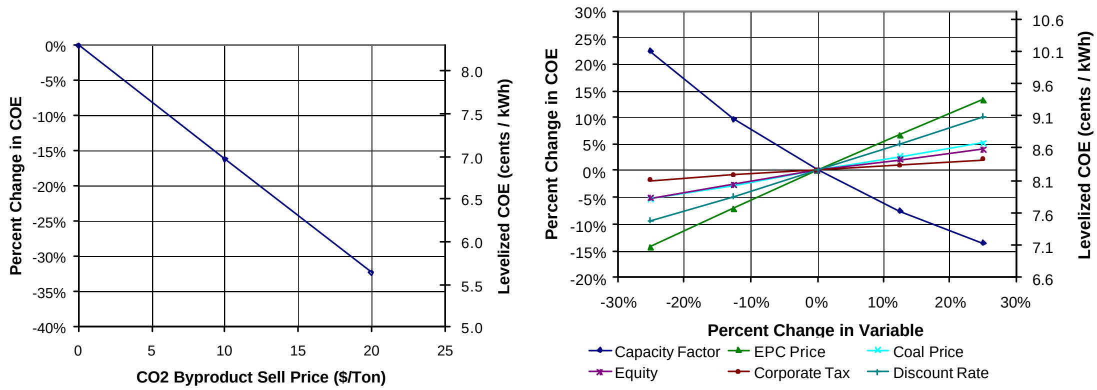


Figure 4.3. 1: Case 2 - Oxygen-Fired CFB with ASU and CO_2 Capture Economic Sensitivity Results

4.3.3.2. Case 9 - Built and Operating Texaco IGCC with CO_2 Capture

Results for the Case 9 COE sensitivity study are shown in Figure 4.3.2. The tabulated results for Case 9 are provided in Appendix IV. The levelized COE for the base parameter values is 7.2 cents per kWh. Levelized COE ranges from a low of 5.1 to a high of 8.9 cents per kWh. CO_2 mitigation costs ranged from -3 to 45 $/ton of CO_2 avoided (reference plant is Case 8) with the baseline value at 23 $/ton of CO_2 avoided.

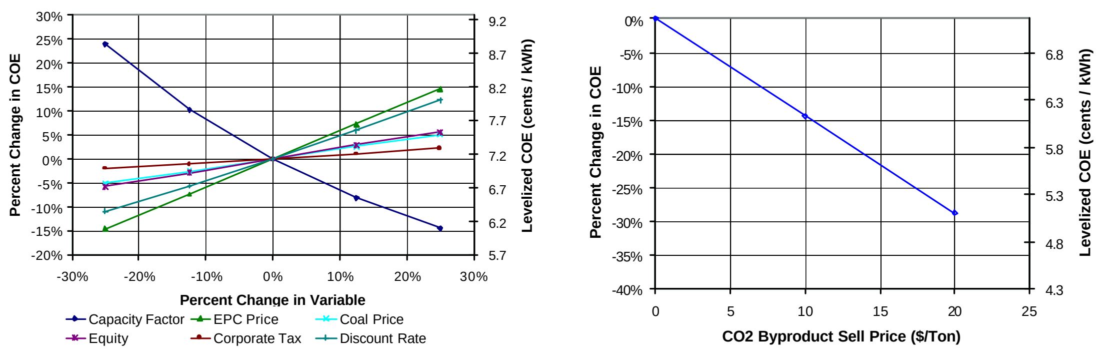


Figure 4.3. 2: Case 9 - Built and Operating Texaco IGCC with CO_2 Capture Economic Sensitivity Results

4.3.3.3. Case 13 - ALSTOM Advanced Chemical Looping with CO2 Capture

Results for the Case 13 COE sensitivity study are shown in Figure 4.3.3. The tabulated results for Case 13 are provided in Appendix IV. The levelized COE for the base parameter values is 5.2 cents per kWh. Levelized COE ranges from a low of 3.3 to a high of 6.3 cents per kWh. CO_2 mitigation costs ranged from -11 to 24 $/ton of CO_2 avoided (reference plant is Case 12) with the baseline value at 11 $/ton of CO_2 avoided.

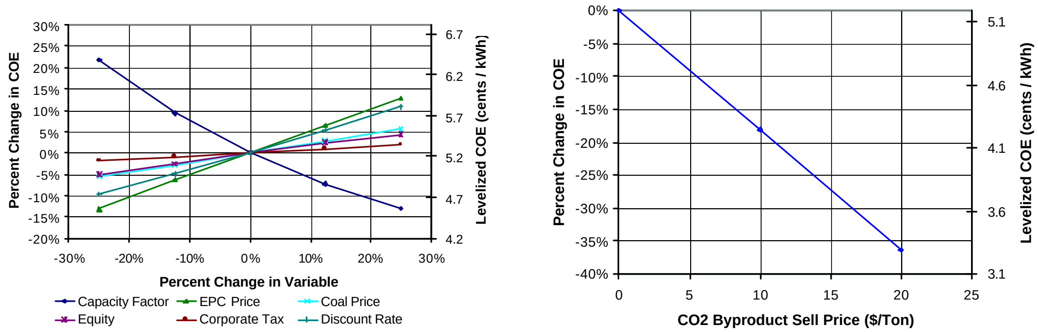


Figure 4.3. 3: Case 13 – ALSTOM Advanced Chemical Looping with CO_2 Capture Economic Sensitivity Results

4.4. Economic Study Summary and Conclusions

The economic study results are summarized by comparing the levelized COE, incremental COE, and CO_2 mitigation cost results for these CO_2 capture technologies as shown in Figures 4.4.1 - 4.4.3, respectively. Figure 4.4.1 shows the levelized COE for all cases studied and Figure 4.4.2 shows incremental COE (relative to the appropriate base case).

The incremental COE for the combustion cases (Cases 2 to 7) as compared to the air-fired CFB reference plant without CO_2 capture (Case 1) ranges from 1.3 to 3.9 cents / kWh. Similarly, the CO_2 mitigation costs range from 13 to 43 $/ton of CO_2 avoided. Case-7 (chemical looping combustion) represents the best combustion case, using either criterion, followed closely by Case-5 (high temperature carbonate regeneration). The cryogenic systems (Cases 2, 3, and 4) fall into a group with incremental COE ranging from 3.4 to 3.9 cents / kWh. Similarly, the CO_2 mitigation costs range from 35 to 43 $/ton of CO_2 avoided. Case 6, which utilizes an OTM for oxygen supply, falls in-between with an incremental COE of 2.5 cents/kWh and a CO_2 mitigation cost of 27 $/ton.

The incremental COE for the gasification cases (Cases 9, 11, 13) as compared to the respective reference plant without CO_2 capture (Case 8, 10, 12) ranges from 0.9 to 2.0 cents / kWh. Similarly, the CO_2 mitigation costs range from 11 to 23 $/ton of CO_2 avoided. Case-13 (advanced Chemical Looping) is clearly the best gasification case using either criterion.

Figure 4.4. 1: Levelized Cost of Electricity Comparison for all Cases

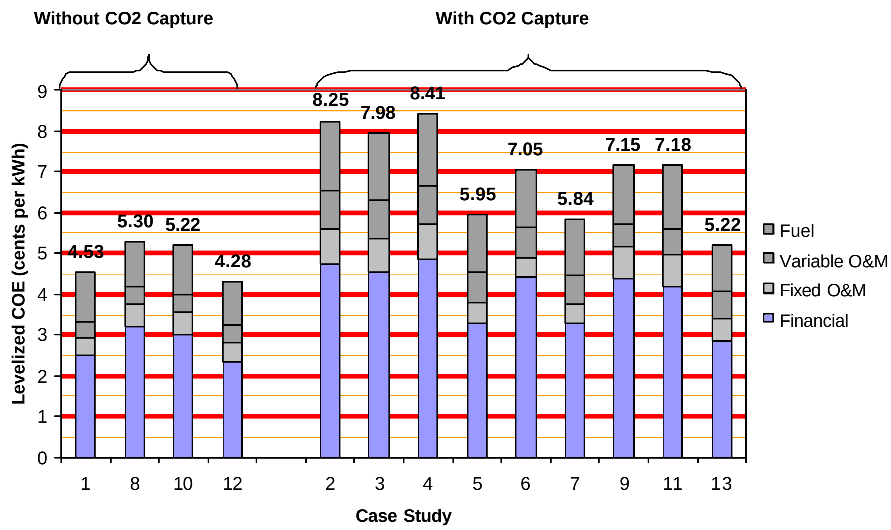


Figure 4.4. 2: Incremental COE of CO_2 Capture Technologies

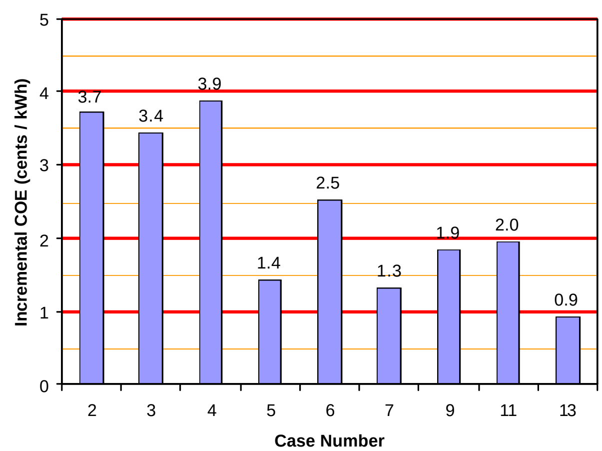
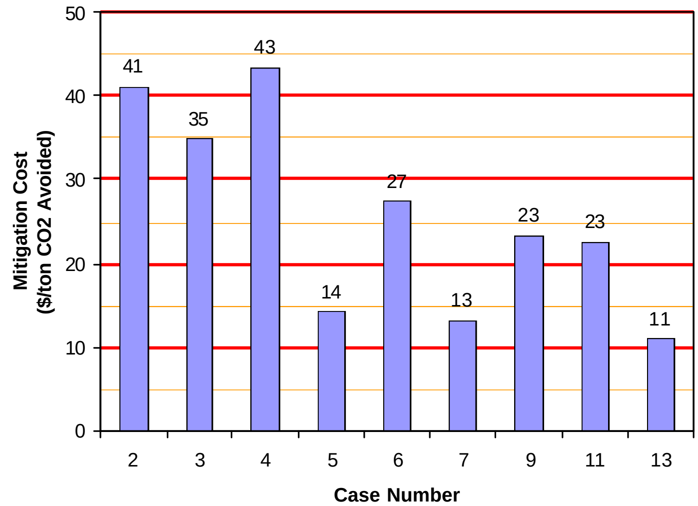
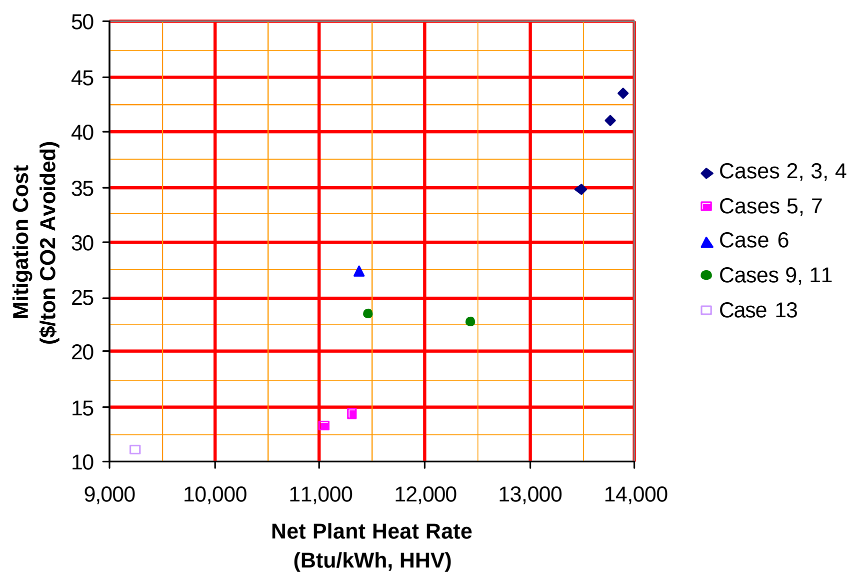


Figure 4.4. 3: Mitigation Costs of CO_2 Capture Technologies



An interesting result is shown in Figure 4.4.4 where the CO_2 mitigation costs are plotted versus the net plant heat rate (NPHR) for all of the CO_2 capture technologies studied. As a technology is less efficient (increasing NPHR), its COE is higher which increases its mitigation cost. Conversely, the technologies with the higher efficiencies (lower NPHR) generally have the lower CO_2 mitigation costs. The strong correlation results highlight the importance of efficiency in the economics of these CO_2 capture technologies.

Figure 4.4. 4: Mitigation Cost versus NPHR of CO_2 Capture Technologies



Cases 2, 3, and 4 (upper right) are the oxygen-fired systems with cryogenic air separation, which increases costs and decreases plant efficiency. Case 6 has the oxygen transport membrane system that helps to reduce cost and increase efficiency. Cases 9 and 11 are the IGCC systems, which are lower in cost and more efficient than Cases 2, 3, and 4. Cases 5 and 7 (air-fired CMB with carbonate regeneration and CMB chemical looping) have comparable efficiencies to the IGCC systems and Case-6 but with lower costs. Case 13, Chemical Looping Gasification, is the most efficient system studied in this work and has the lowest costs.

Advanced Chemical Looping Gasification (Case 13) is calculated to be the best CO_2 capture alternative evaluated in this study based on levelized COE, incremental COE, and mitigation cost of avoided CO_2. The levelized COE for advanced Chemical Looping Gasification with CO_2 capture is 5.2 cents / kWh. This is 10 percent lower than the corresponding value for Case 7 (Chemical Looping combustion – lowest of the combustion cases) and 27 percent lower than the corresponding value for Case 9 (built and operating IGCC – lowest of the IGCC systems). As compared to the air-fired CFB system (Case-1), the advanced Chemical Looping case (Case 13) is only about 15 percent higher with respect to levelized COE. If the CO_2 by-product can be sold at $10 / ton, the levelized COE drops to 4.3 cents / kWh which is about 5 percent lower than today's air-fired CFB system (Case-1) without CO_2 capture at 4.5 cents / kWh and equivalent to advanced Chemical Looping gasification without CO_2 capture (Case 12).

5. COMPARATIVE ANALYSIS

This section provides a summary and comparison of several important outputs from this study as well as a comparison to other work reported in the literature. The outputs highlighted include plant performance parameters, CO_2 emissions, plant investment costs, O&M costs, cost of electricity (COE), and CO_2 mitigation costs.

This section is structured in such a way as to first compare the combustion based system results (Cases 1-7) from this study amongst themselves followed by a comparison of the gasification system results (Cases 8-13) amongst themselves. Next, the Combustion system results are compared to the gasification based system results. Finally, the results from this study are compared with selected results from the literature.

5.1. Combustion Cases

This section summarizes and compares overall system performance, CO_2 emissions, costs, and economics for the seven combustion cases (Cases 1-7).

5.1.1. Combustion Cases: Performance and CO_2 Emissions Comparison

This section summarizes overall system performance and CO_2 emissions for the combustion cases (Cases 1-7). Table 5.1.1 shows a fairly detailed comparison of plant performance and CO_2 emissions for both the CO_2 recovery concepts (Cases 2-7) and the Base Case (Case-1) that employs no CO_2 recovery system.

Additionally, selected results from this table are illustrated and compared in Figures 5.1.1– 5.1.7. The comparisons shown in the figures are Boiler Efficiency, Coal Heat Input, Steam Cycle Efficiency, Gas Processing System Auxiliary Power, Total Plant Auxiliary Power, Net Plant Output, Plant Thermal Efficiency, and Plant CO_2 Emissions.

Table 5.1. 1: Combustion Cases Plant Performance Summary & Comparison

	(Units)	CFB Air Fired (Case 1)	CFB Cryogenic O_2 Fired (Case 2)	CFB Cryogenic O_2 Fired (Case 3)	CMB Cryogenic O_2 Fired (Case 4)	CMB Air Fired High Temp Carb Proc (Case 5)	CMB with OTM O_2 Fired (Case 6)	CMB Chemical Looping (Case 7)
Auxiliary Power Listing								
Induced Draft Fan	(kW)	2285	511	511	515	7679	626	1117
Primary Air Fan	(kW)	2427	n/a	n/a	n/a	2015	n/a	2559
Secondary Air Fan	(kW)	1142	n/a	n/a	n/a	n/a	n/a	n/a
Fluidizing Air Blower	(kW)	920	209	209	209	n/a	209	n/a
Transport Air Fan	(kW)	n/a	n/a	n/a	1865	244	2409	46
Gas Recirculation Fan	(kW)	n/a	341	341	344	n/a	795	n/a
Coal Handling, Preparation, and Feed	(kW)	300	292	292	294	293	363	293
Limestone Handling and Feed	(kW)	200	195	195	196	217	242	173
Limestone Blower	(kW)	150	146	146	147	163	181	130
Ash Handling	(kW)	200	195	195	196	205	242	189
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	151	151	152	n/a	186	n/a
Boiler Feed Pump	(kW)	3715	3715	3715	3715	3756	3715	3757
Condensate Pump	(kW)	79	79	79	79	80	92	80
Circulating Water Pump	(kW)	1400	1877	1729	1889	1436	2006	1563
Cooling Tower Fans	(kW)	1400	1877	1729	1889	1436	2006	1563
Steam Turbine Auxiliaries	(kW)	200	206	206	207	187	253	187
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719	719	719	719	719	719
Transformer Loss	(kW)	470	472	472	472	456	525	456
Subtotal	(kW)	16007	10983	10687	12888	18887	14570	12833
	(frac. of Gen. Output)	0.077	0.052	0.051	0.061	0.093	0.062	0.063
Auxiliary Power Summary								
Traditional Power Plant Auxiliary Power	(kW)	16007	10983	10687	12888	18887	14570	12833
Air Separation Unit or Fuel Compressor	(kW)	n/a	37505	37505	37800	n/a	n/a	n/a
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a	n/a	n/a	n/a	110920	n/a
CO_2 Removal System Auxiliary Power	(kW)	n/a	26905	26364	27200	22878	33434	25453
Total Auxiliary Power	(kW)	16007	75393	74556	77888	41765	158923	38287
	(frac. of Gen. Output)	0.077	0.359	0.355	0.371	0.206	0.196	0.189
Output and Efficiency								
Main Steam Flow	(lbm/hr)	1400555	1400555	1400555	1400555	1400555	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8256	8256	8275	8397	8758	8404
OTM System Expander Generator Output	(kW)	n/a	n/a	n/a	n/a	n/a	122659	n/a
Gas Turbine Generator Output		n/a	n/a	n/a	n/a	n/a	n/a	n/a
Steam Turbine Generator Output	(kW)	209041	209907	209907	210056	202949	233699	202770
Net Plant Output	(kW)	193034	134514	135351	132168	161183	197435	164483
	(frac. of Case-1 Net Output)	1.00	0.70	0.70	0.68	0.84	1.02	0.85
Boiler Efficiency (HHV)[1]	(fraction)	0.8946	0.9412	0.9412	0.9366	0.9217	0.9404	0.9242
Coal Heat Input (HHV)	(10^6 Btu/hr)	1855	1806	1806	1820	1815	2242	1810
Natural Gas Heat Input (HHV)[2]	(10^6 Btu/hr)	n/a	16.5	20.6	16.6	7.9	4.8	7.9
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	1855	1822	1826	1836	1822	2247	1818

[1] Boiler Heat Output / (Qcoal-HHV + Qcredits)

[2] Required for GPS Desiccant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4

	(Units)	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6	Case 7
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	13546	13492	13894	11307	11380	11051
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.2520	0.2530	0.2456	0.3019	0.2999	0.3088
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.71	0.71	0.69	0.85	0.84	0.87
CO_2 Emissions								
CO_2 Produced	(lbm/hr)	385427	376995	377466	379959	359997	466301	384453
CO_2 Captured	(lbm/hr)	0	352377	375095	352380	359030	437084	383420
Fraction of CO2 Captured	(fraction)	0.00	0.93	0.99	0.93	1.00	0.94	1.00
CO_2 Emitted	(lbm/hr)	385427	24618	2371	27579	967	29217	1033
Specific CO_2 Emissions	(lbm/kwhr)	2.00	0.18	0.02	0.21	0.01	0.15	0.01
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.09	0.01	0.10	0.00	0.07	0.00
Avoided CO_2 Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.81	1.98	1.79	1.99	1.85	1.99

Boiler Efficiency:

Figure 5.1.1 compares boiler efficiencies among the seven combustion cases. Case-1, the air-fired Base Case, is lower than the other cases primarily due to a higher dry gas loss. The higher dry gas loss is the result of higher gas flow and/or temperature exiting the Boiler Island. Cases 2, 3, 4, and 6 are all oxygen-fired and therefore the exiting flue gas flow rates from the Boiler Island are much lower than for Case-1 resulting in the reduced dry gas losses for these cases. These cases also have their flue gas streams cooled to lower temperatures than Case-1 also contributing to the reduction in dry gas loss. The flue gas flow rates are much lower for these cases since nearly all of the nitrogen is removed in the ASU system. Boiler efficiency improvement falls in the 5-6 percent range. Cases 5 and 7 are air fired but have flue gas streams which are cooled to lower temperatures than Case-1 and therefore have lower dry gas losses. These cases show boiler efficiency improvements of about 2.7 and 2.9 percent respectively.

Figure 5.1. 1: Combustion Cases Plant Boiler Efficiency Comparison

Coal Heat Input:

Figure 5.1.2 compares coal heat input to the boilers for the seven combustion cases. Case-1, the air-fired Base Case, has a slightly higher coal heat input than any of the other cases except Case-6 primarily due to the fact that the boiler efficiency is slightly lower for this case as explained above.

The boiler heat output is nearly the same for all cases except Case-6 since all the cases (including Case-6) use a steam cycle that is nearly identical. The only differences in the steam cycles for the combustion cases are in the low level heat recovery systems.

The significant increase in boiler heat output for Case-6 is because a substantial amount of heat is transferred to the high temperature air heater in this case. This air heater is used to pre-heat the air supplied to the Oxygen Transport Membrane to 1,650 ^{0}F.

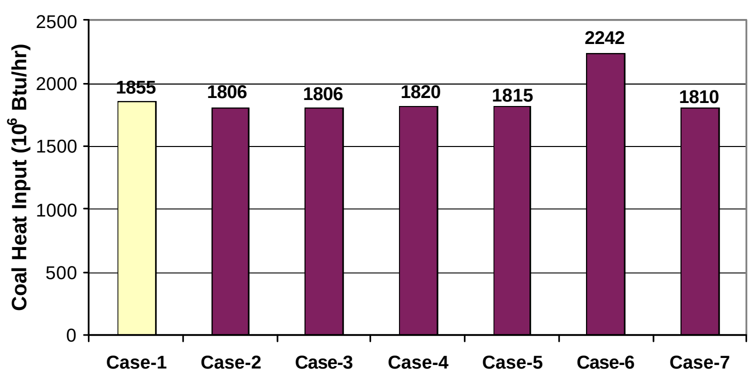


Figure 5.1. 2: Combustion Cases Coal Heat Input Comparison

Steam Cycle Efficiency:

Figure 5.1.3 compares steam cycle efficiency for the seven combustion cases. Case-1, the air-fired Base Case, has a slightly higher steam cycle efficiency than any of the other cases primarily due to the fact that there is no low level heat recovery (LLHR) by feedwater streams in parallel with the extraction feedwater heaters.

All other combustion cases in this study except Cases 5 and 7 include LLHR in parallel with the extraction feedwater heaters system to some degree. Cases 2, 3, and 4, all include various amounts of feedwater heating in parallel with the low temperature extraction feedwater heaters (Heaters 1 and 2).

Cases 5 and 7 do not use LLHR systems, however, both cases include relatively small low-pressure process steam extractions (about 5 percent of main steam flow) from the steam turbine to the Boiler Island. These extractions also tend to reduce the steam cycle efficiency somewhat.

Case-6 includes the largest amount of LLHR. The Case-6 low level heat recovery systems are in parallel with all six closed extraction feedwater heaters. This is a result of the OTM system, which requires a relatively large amount of heat recovery downstream of the gas expander which expands the depleted oxygen air stream leaving the OTM.

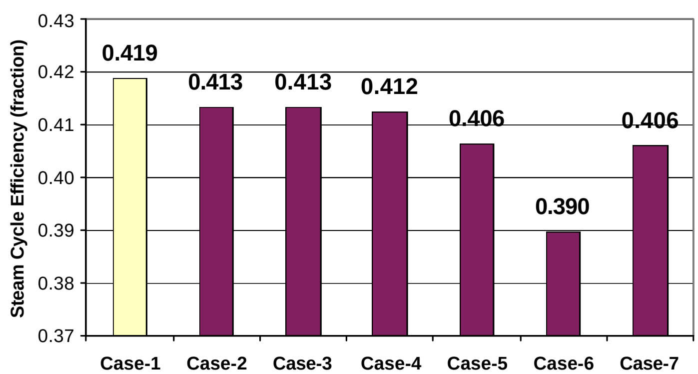


Figure 5.1. 3: Combustion Cases Plant Steam Cycle Efficiency Comparison

Gas Processing System Auxiliary Power:

Each of the CO_2 capture cases (Cases 2-7) requires some type of CO_2 compression system within their respective Gas Processing Systems (GPS). Each of these Gas Processing Systems differs slightly in its design and / or operating conditions. These differences are a result of gas analysis variations from case to case or, for Case-3, different products gas specification (Note: for Case-3 the product gas is used for sequestration only). These design and/or operating condition differences can cause subtle variations in the power requirements for the respective Gas Processing Systems. These power requirements varied from about 127-154 kWh/ton of CO_2 captured as shown below in Figure 5.1.4.

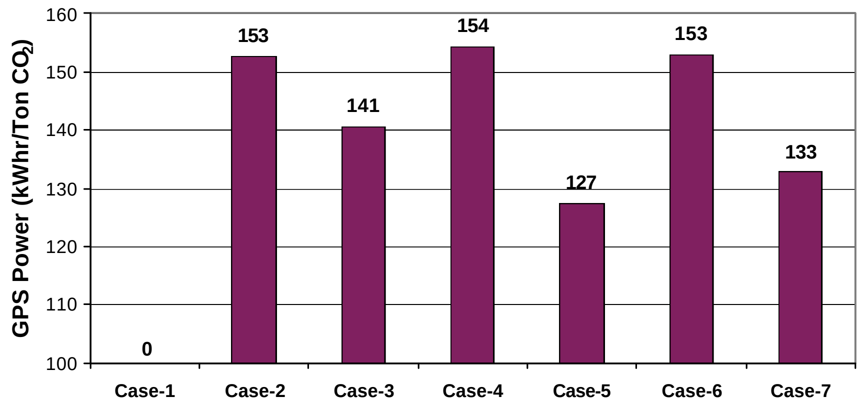


Figure 5.1. 4: Combustion Cases Gas Processing System Normalized Auxiliary Power Comparison

Total Plant Auxiliary Power:

Figure 5.1.5 compares total plant auxiliary power among the seven combustion cases. There are three main components that comprise the total plant auxiliary power. These are (1) the Gas Processing System (discussed above), (2) the oxygen supply system (ASU, OTM or air fired), and (3) the traditional power plant auxiliaries associated with the draft system, cooling water system, material handling, etc.

Case-1, the air-fired Base Case without CO_2 recovery, requires much less auxiliary power than the other cases, since it does not require an ASU or a Gas Processing System to compress the CO_2. This case requires slightly less than 8 percent of the generator output for auxiliary power.

Cases 2, 3, and 4 are all oxygen-fired systems; and, therefore, the cryogenic based ASU system adds a significant load to the plant auxiliary power requirement (about 231 kWh/ton of oxygen supplied in these cases or about 18 percent of the steam turbine generator output). Case 6 is also oxygen fired but utilizes an advanced OTM system. This OTM system which includes an air compressor, high temperature air heater, OTM, and gas expander actually generates a small amount of power. This system can best be described as a modified Brayton Cycle operating at relatively low temperature (~1,650 ^{o}F). Cases 5 and 7 are both air fired systems and, therefore, do not require any ASU systems and require significantly less auxiliary power.

The other component of auxiliary power is that which is attributable to the traditional power plant part of these systems. This includes equipment for solids handling (coal, limestone, and ash), air and gas handling, water pumping for the steam cycle and cooling water systems, as well as other miscellaneous systems within the traditional power plant. Total power requirements for these systems range from about 5 to 9 percent of steam turbine generator output for these cases as shown in Table 5.1.1.

Figure 5.1. 5: Combustion Cases Total Plant Auxiliary Power Comparison

Net Plant Power Output:

Figure 5.1.6 compares the resulting net power output (MWe) among the seven combustion cases. Cases 2, 3, and 4, which utilize the cryogenic ASU systems, show the lowest net plant outputs. Case-6, which uses the OTM system to produce oxygen, has the highest net plant output of all the cases. It should be pointed out, however, that this case also has a significantly higher coal heat input (by about 20 percent) than the other cases as a result of the OTM high temperature air heating which is done in the CMB boiler.

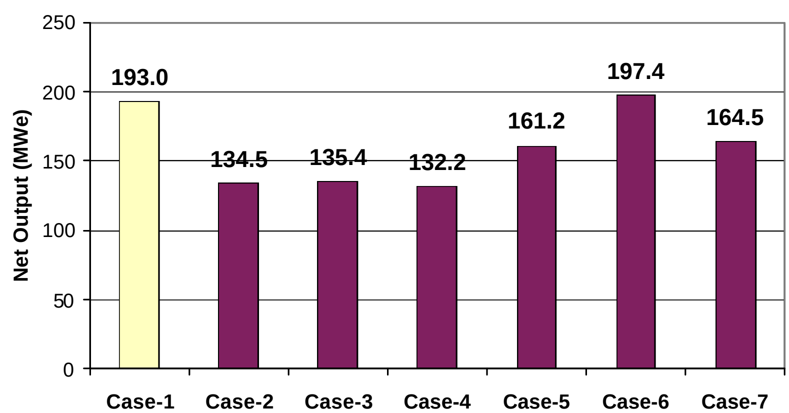


Figure 5.1. 6: Combustion Cases Net Plant Output Comparison

Plant Thermal Efficiency:

Figure 5.1.7 compares overall plant thermal efficiency (HHV basis) among the seven combustion cases. These efficiency results reflect the combined impact of boiler efficiency, steam cycle efficiency, and plant auxiliary power on overall plant thermal efficiency. As shown previously, the differences in plant auxiliary power represent the dominant factor for differences in overall plant thermal efficiency for the cases studied.

The resulting energy penalties associated with Cases 2-7 compared to Case-1 fall into two groups. The first group, which includes Cases 5, 6, and 7, shows an energy penalty of about 12 percent. The second group, which includes Cases 2, 3, and 4, shows an energy penalty of about 30 percent.

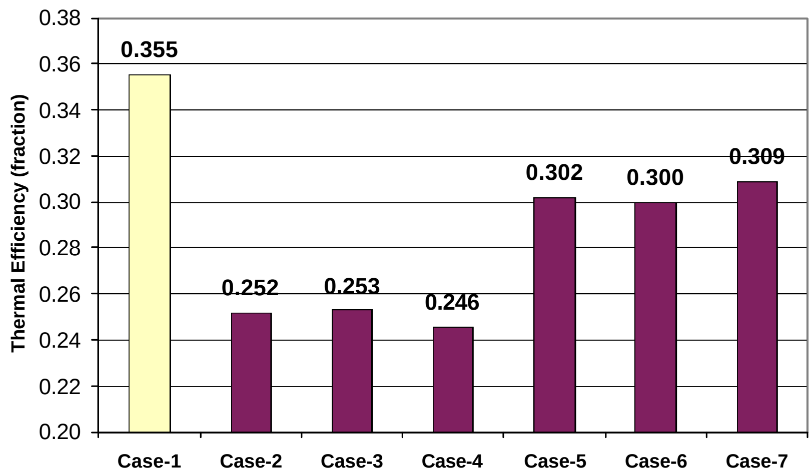


Figure 5.1. 7: Combustion Cases Plant Thermal Efficiency Comparison

Plant CO_2 Emissions:

Figure 5.1.8 compares overall plant CO_2 emissions on a normalized basis (lbm/kWh) among the seven combustion cases. Also shown in this figure are the quantities of captured CO_2 (normalized basis). The Base Case (Case-1) emits about 2.0 lbm/kWh of CO_2 as is typical for coal fired power plants with subcritical steam cycles. The remaining cases, all of which include CO_2 capture systems, show normalized CO_2 emissions ranging from about 0.01 to 0.21 lbm/kWh of CO_2. All these systems capture more than 93 percent of the CO_2 produced by the respective plants.

The upper bars (lighter shade) and the upper set of data labels (bold) shown on the figure indicate the normalized quantities of CO_2 captured. The captured quantities of CO_2 range from about 2.2 to 2.8 lbm/kWh for the six CO_2 capture cases. The lower bars (darker shade) and the lower set of data labels show the normalized CO_2 emitted. The emitted quantities of CO_2 range from about 0.01 to 0.21 lbm/kWh for the six CO_2 capture cases. The sum of these two quantities (captured + emitted) of course represents the quantity of CO_2 produced (e.g., the Case-2 power plant produces 2.66 + 0.19 = 2.85 lbm/kWh of CO_2 on a normalized basis).

Figure 5.1.9 compares avoided CO_2 emissions on a normalized basis (lbm/kWh) among the seven combustion cases. The avoided CO_2 emissions are calculated relative to the

Base Case (Case-1) shown in this figure. The avoided quantities of CO_2 range from about 1.8 to 2.0 lbm/kWh for the six CO_2 capture cases.

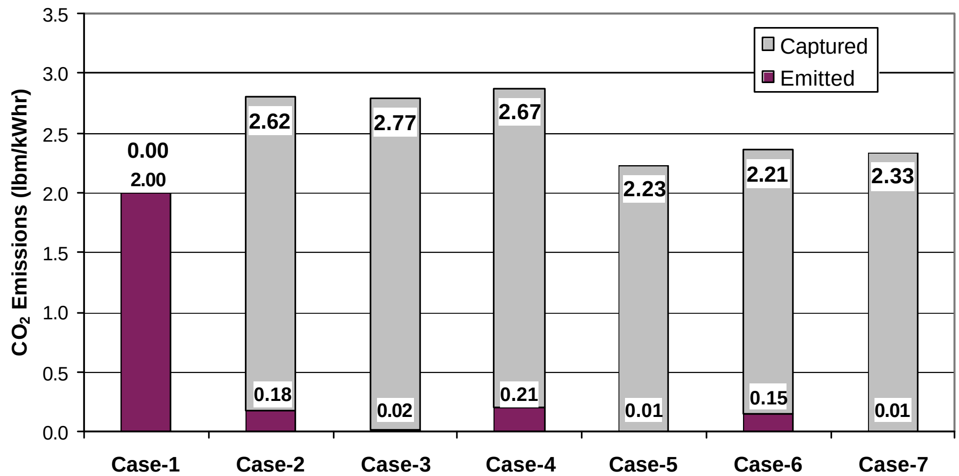


Figure 5.1. 8: Combustion Cases Plant CO_2 Emission Comparison

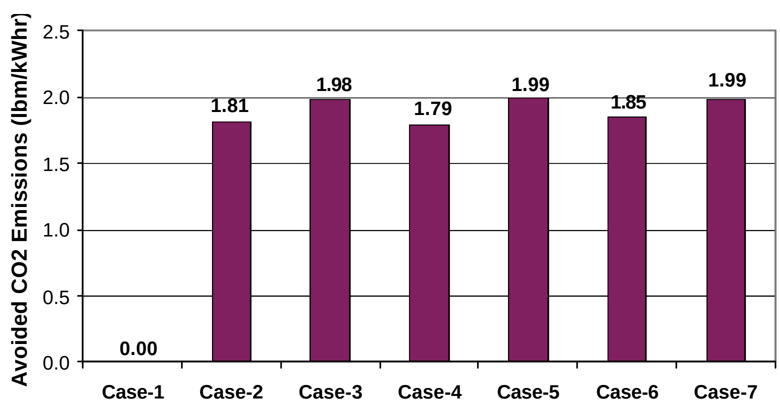


Figure 5.1. 9: Combustion Cases Avoided CO_2 Emission Comparison (Relative to Case-1)

5.1.2. Combustion Cases: Costs and Economics Comparison

This section summarizes and compares overall system costs (both investment and O&M), and economic results for the seven combustion cases (Cases 1-7).

<u>Investment Costs:</u>
The plant investment costs for the combustion cases are shown in Table 5.1.2 and Figure 5.1.10. The plant investment cost for the Base Case (Case-1) without CO_2 capture was 1,304 \$/kW. The plant investment cost range for the remaining cases (Cases 2-7) with

CO_2 capture was from about 1,660 to 2,550 $/kW. Case 7 (Chemical Looping Combustion) was found to be the lowest cost of the capture cases (1,663 $/kW) followed closely by Case-5 (Regenerative Carbonate Process) at 1,677 $/kW. Cases 2, 3, and 4, all variants of the cryogenic based oxygen fired process, were found to have significantly higher costs (2,370 – 2,550 $/kW). Case-3, which used a simplified Gas Processing System (only drying and compression) such that the flue gas was produced for sequestration only, cost 2,369 $/kW. This represents a savings of about 74 $/kW or about 3 percent as compared to Case-2 which purified the flue gas to meet the product gas specification (refer to Table 2.0.1). Case 6 (oxygen fired via an advanced OTM system) was slightly less costly (2,375 $/kW) than the comparable cryogenic Case-4. The savings was about 7 percent or 178$/kW for Case 6 as compared to Case-4.

Table 5.1. 2: Plant Investment Costs for Combustion Cases

Study Case	Net Plant Output, kW	Total Investment Cost, EPC Basis	
		$x1000	$/kW
Case 1, Air-fired CFB w/o CO_2 Capture	193,037	251,804	1,304
Case 2, O_2-Fired CFB w/ASU & CO_2 Capture	134,514	328,589	2,443
Case 3, O_2-Fired CFB w/ASU & Flue Gas Sequestration	135,351	320,638	2,369
Case 4, O_2-Fired CMB w/ASU & CO_2 Capture	132,168	337,402	2,553
Case 5, Air-Fired CFB w/Carbonate Reg. Process & CO_2 Capture	161,184	270,232	1,677
Case 6, O_2-Fired CMB w/OTM & CO_2 Capture	197,435	468,919	2,375
Case 7, CMB Chemical Looping Combustion w/CO_2 Capture	164,484	273,568	1,663

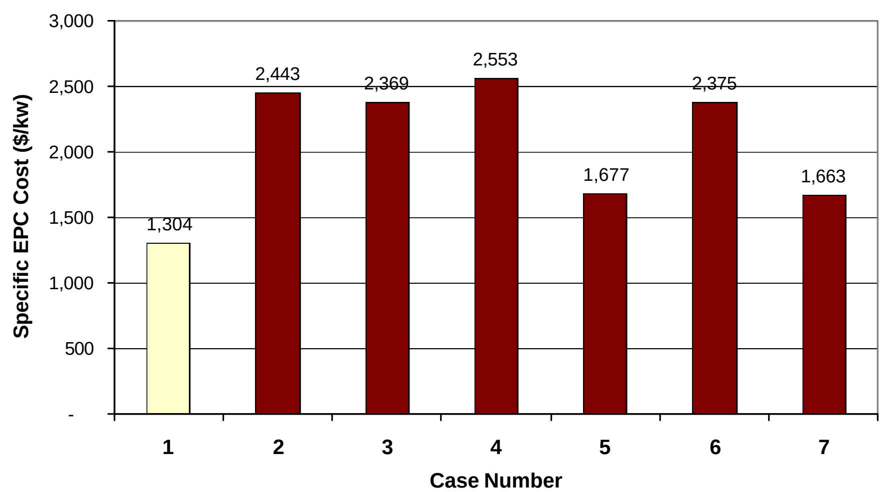


Figure 5.1. 10: Plant Investment Costs for Combustion Cases

Operating and maintenance (O&M) costs were calculated for all systems for the combustion cases as shown in Table 5.1.3. Total O&M costs for the capture cases ranged from about 1.2 to 1.8 cents/kWh while the Base Case (Case-1) was about 0.8 cents/kWh.

Table 5.1. 3: Operating and Maintenance Costs for Combustion Cases

Study Case	Net Plant Output, kW	Operating & Maintenance (O&M) Costs					Total O&M, cents/kWh
		Fixed		Variable @ 80% Capacity		Total, $	
		$	$/kW	$	$/kWh		
Case 1, Air-fired CFB w/o CO_2 Capture	193,037	5,657,635	29.31	5,587,188	0.0041	11,244,823	0.83
Case 2, O_2-Fired CFB w/ASU & CO_2 Capture	134,514	7,853,885	58.39	8,820,048	0.0094	16,673,933	1.77
Case 3, O_2-Fired CFB w/ASU & Flue Gas Sequestration	135,351	8,060,787	59.55	8,653,810	0.0091	16,714,598	1.76
Case 4, O_2-Fired CMB w/ASU & CO_2 Capture	132,168	7,899,070	59.77	8,889,066	0.0096	16,788,137	1.81
Case 5, Air-Fired CFB w/Carbonate Reg. Process & CO_2 Capture	161,184	5,799,465	35.98	8,264,460	0.0073	14,063,925	1.25
Case 6, O_2-Fired CMB w/OTM & CO_2 Capture	197,435	6,537,784	33.11	10,133,605	0.0073	16,671,389	1.20
Case 7, CMB Chemical Looping Combustion w/CO_2 Capture	164,484	5,797,471	35.25	8,014,747	0.0070	13,812,218	1.20

Economic Evaluations:

Table 5.1.4 and figure 5.1.11 summarize the levelized economic analysis results for the seven combustion cases.

Table 5.1. 4:Cost of Electricity for Combustion Cases

Study Case	Levelized Cost of Electricity (c/kWh)					Incremental COE (c/kWh)
	Financial	Fixed O&M	Variable O&M	Fuel	Total	
Without CO2 Capture						
Case 1, Air-fired CFB w/o CO2 Capture	2.49	0.42	0.41	1.20	4.53	
With CO2 Capture						
Case 2, O2-Fired CFB w/ASU & CO2 Capture	4.73	0.85	0.95	1.72	8.25	3.72
Case 3, O2-Fired CFB w/ASU & Flue Gas Sequestration	4.53	0.85	0.91	1.69	7.98	3.45
Case 4, O2-Fired CMB w/ASU & CO2 Capture	4.86	0.85	0.96	1.74	8.41	3.88
Case 5, Air-Fired CFB w/Carbonate Reg. Process & CO2 Capture	3.29	0.51	0.73	1.41	5.95	1.42
Case 6, O2-Fired CMB w/OTM & CO2 Capture	4.43	0.47	0.73	1.42	7.05	2.53
Case 7, CMB Chemical Looping Combustion w/CO2 Capture	3.26	0.50	0.70	1.38	5.84	1.32

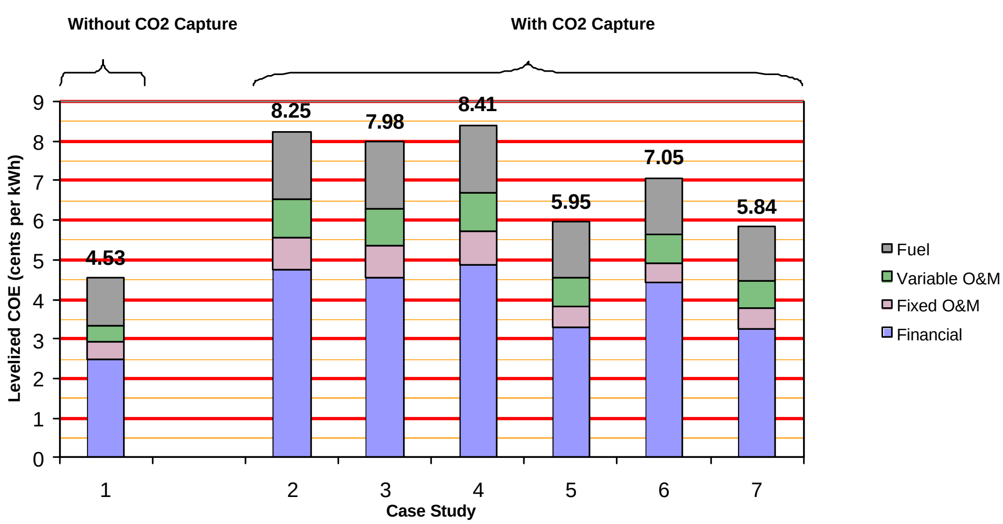


Figure 5.1. 11: Cost of Electricity for Combustion Cases

Case-7 (Chemical Looping Combustion) was found to be the best alternative of the six capture concepts studied based on levelized Cost of Electricity (COE) evaluation criteria (5.84 cents/kWh). Case-5 (High Temperature Carbonate Regeneration) is only slightly worse (by about 2 percent) than Case-7. Case-7 was found to have an incremental COE value of 1.31 cents/kWh as compared to Case-1 (about a 29 percent increase). The cryogenic ASU cases (Cases 2, 3, and 4) were significantly higher than Case-7 (by about 40 percent). Case-6, which used an OTM for oxygen production, was in between, about 20 percent higher than Case-7 and about 20 percent better than Cases 2, 3 and 4 which used cryogenic ASU's.

Case-3 which uses a simplified Gas Processing System (no purification) and produces a product suitable for sequestration only showed a relatively insignificant improvement in COE of about 3 percent as compared to Case-2 where product purification was used.

5.2. Gasification Cases

This section summarizes overall system performance, CO_2 emissions, costs (investment and O&M), and economic results for the four IGCC and two advanced Chemical Looping cases (Texaco based: Cases 8-11and Alstom Chemical Looping: Cases 12-13).

5.2.1. Gasification Cases: Performance and CO_2 Emissions Comparison

Table 5.2.1 shows a fairly detailed comparison of plant performance and CO_2 emissions for both the Base Cases (Cases 8, 10 and 12) that employ no CO_2 recovery systems and the respective CO_2 Recovery Concepts (Cases 9, 11 and 13).

By way of review, the primary differences between the gasification cases are described as follows.

- Cases 8 and 9 are both based on the built and operating Tampa IGCC demonstration project process with gasifiers operating at about 450 psig and include radiant syngas coolers. Case-9 includes CO_2 capture and Case-8 does not.

- Cases 10 and 11 are both based on current commercially offered but not yet built and operating designs with gasifiers at 950 psig with syngas expanders and quench cooling. Case-11 includes CO_2 capture and Case-10 does not.

- Cases 12 and 13 are based on an advanced Chemical Looping gasification process Alstom is developing. Case-13 includes CO_2 capture and Case-12 does not.

- All Cases (8-13) use a single train GE-7FA gas turbine, HRSG, and an 1,800 psig, 1,000 ^{o}F, 1,000 ^{o}F, 3.0 in. Hga steam cycle.

Additionally, selected results from Table 5.2.1 are illustrated and compared in Figures 5.2.1-5.2.6. The comparisons shown in the figures are Gas Processing Systems Auxiliary Power, Total Plant Auxiliary Power, Plant Net Output, Plant Thermal Efficiency, and Plant CO_2 Emissions.

Table 5.2. 1: Gasification Plants Performance Summary and Comparisons

	(Units)	Texaco Built and Operating IGCC w/o CO_2 Removal (Case 8)	Texaco Built and Operating IGCC w/ CO_2 Removal (Case 9)	Texaco Commercially Offered IGCC w/o CO_2 Removal (Case 10)	Texaco Commercially Offered IGCC w/ CO_2 Removal (Case 11)	Chemical Looping Gasification w/o CO_2 Removal (Case 12)	Chemical Looping Gasification w/ CO_2 Removal (Case 13)
Power Generator Outputs							
Gas Turbine Power	(kW)	187,150	187,150	187,150	187,150	197,000	197,000
Sweet Gas Expander Power	(kW)	0	0	6,650	6,570	0	0
Steam Turbine Power	(kW)	113,717	112,318	97,924	96,550	104,990	117,379
Gross Plant Power	(kW)	300,867	299,468	291,724	290,270	301,990	314,379
Key Auxiliary Power Listing							
ASU Auxiliaries	(kW)	20,680	22,911	20,080	23,302	0	0
Fuel Compressor	(kW)					29,200	13,080
Oxygen Compressor	(kW)	9,150	10,137	10,570	10,990	0	0
CO_2 Compressor	(kW)	0	27,105	0	25,644	0	35,469
Balance of Auxiliaries	(kW)	7,950	8,800	25,780	29,330	7,644	8,999
Total Auxilary Power	(kW)	37,780	68,953	56,430	89,266	36,844	57,548
Auxiliary Power, % of Gross	(kW)	12.6	23.0	19.3	30.8	12.2	18.3
Net Plant Power	(kW)	263,087	230,515	235,294	201,004	265,146	256,830
Coal Feed Rate	(lbm/hr)	215,454	238,694	210,010	225,822	197,428	213,582
Gasifier Oxygen (95% pure)	(lbm/hr)	183,333	204,167	174,309	187,431	0	0
Thermal Input (HHV)	(kW-thermal)	699,073	774,479	681,410	732,714	640,756	696,012
Net Plant Thermal Efficiency (HHV)	(percent)	37.6	29.8	34.5	27.4	41.4	36.9
Net Plant Heat Rate (HHV)	(Btu/kWhr)	9,069	11,467	9,884	12,441	8,248	9,249
CO_2 Emissions							
CO_2 Produced	(lbm/hr)	477,093	528,791	464,940	500,275	454,321	492,600
CO_2 Captured	(lbm/hr)	0	476,042	0	450,379	0	486,572
CO_2 Fraction Captured	(frac)	0.00	0.90	0.00	0.90	0.00	0.99
Specific CO_2 Captured	(lbm/kWhr)	0.00	2.07	0.00	2.24	0.00	1.89
CO_2 Emitted	(lbm/hr)	477,093	52,749	464,940	49,896	454,321	6,028
Specific CO_2 Emissions	(lbm/kWhr)	1.81	0.23	1.98	0.25	1.71	0.02
Normalized Specific CO_2 Emissions (Relative to Base)	(frac)	1.00	0.13	1.00	0.13	1.00	0.01
Avoided CO_2 Emissions (Compared to Base Case)	(lbm/kWhr)	0.00	1.58	0.00	1.73	0.00	1.69

Gas Processing System Auxiliary Power:

Each of Cases 9, 11 and 13 requires a CO_2 compression system within its Gas Processing System (GPS) to provide the product stream at 2,000 psig.

The Case 9 and 11 Gas Processing Systems are very similar in their design and/or operating conditions. Hence, their auxiliary power consumption's were the same at about 114 kWh/ton-CO_2 captured. Both the Case 9 and Case 11 gas compression systems receive CO_2 from the CO_2 capture system at two pressures. About 90 percent of the CO_2 captured is flashed off from the solvent at 50 psia with the remainder at 14.7 psia. The stream at 14.7 psia is boosted to 50 psia and combined with the first CO_2 stream for final compression.

The Case 13 Gas Processing System compresses the entire CO_2 product stream starting at 14.7 psia and therefore requires about 28 percent more compression power than Cases 9 and 11 (146 kWh/ton vs. 114 kWh/ton) as shown in Figure 5.2.1.

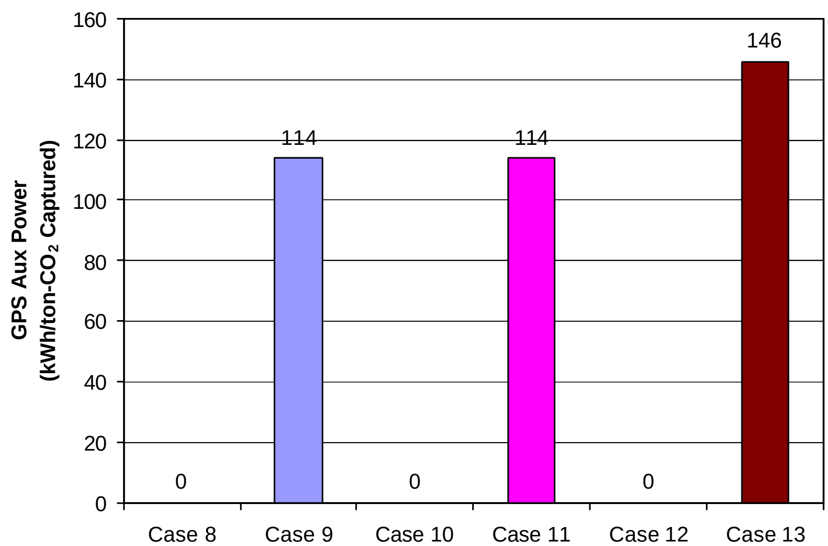


Figure 5.2. 1: Gasification Plants Gas Processing System Normalized Auxiliary Power Comparison

Total Plant Auxiliary Power:

Figure 5.2.2 compares total plant auxiliary power among the four IGCC and two advanced Chemical Looping cases. There are four main components that comprise the total auxiliary power. These are (1) the Gas Processing System; (2) the Air Separation Unit (Cases 8-11), (3) the Fuel Compressor (Cases 12 and 13) and (4) the traditional auxiliary power plant auxiliary associated with the steam cycle, material handling, etc.

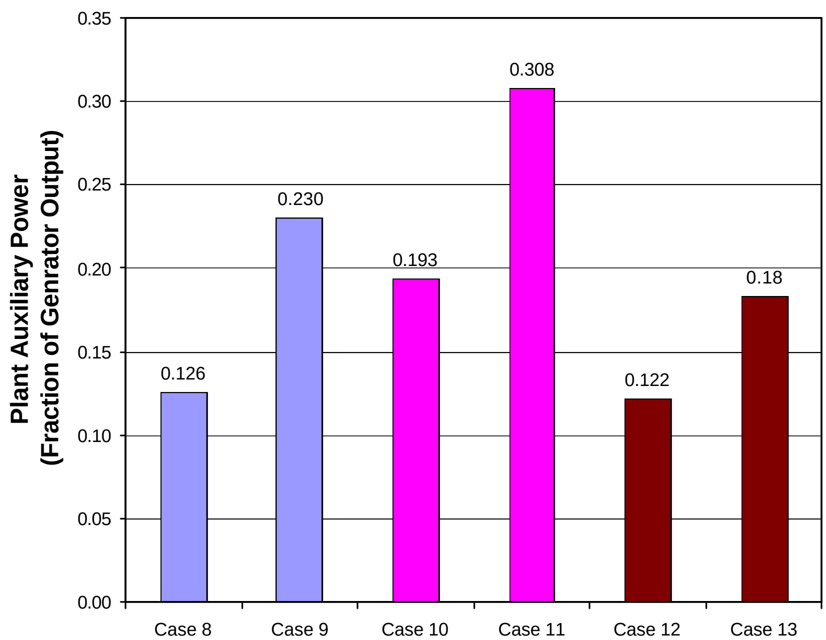


Figure 5.2. 2: Gasification Plants Total Plant Auxiliary Power Comparison

Base Cases 8 and 10 (Built and Operating IGCC and Commercially Offered IGCC, without CO_2 Capture) each requires an Air Separation Unit (ASU), but not a Gas Processing System to compress the CO_2. The total Auxiliary Power for these cases require about 13 and 19 percent of the Gross Power outputs, respectively. Both cases' ASU require roughly similar amounts of auxiliary power. Hence, the basic differences in the overall auxiliary power requirements between them come from their gasification islands: Case 8 uses a radiant syngas cooler gasifier, and Case 10 uses a quench cooler. Cases 9 and 11 (Built and Operating IGCC and Commercially Offered IGCC, w/CO_2 Capture) each requires an Air Separation Unit (ASU) and a Gas Processing System to compress the CO_2. These cases require much more auxiliary power than their respective base cases, 23 and 31 percent of Gross Power. Again, the difference between them is basically from their gasification islands.

Cases 12 and 13 both utilize air fired advanced Chemical Looping systems and therefore do not require an ASU; however, they do require Fuel Compressors since the gasifiers operate at near atmospheric pressure. The total Auxiliary Power for these cases requires about 12 and 18 percent of the Gross Power outputs, respectively.

Plant Net Output:

Figure 5.2.3 compares the resulting net power outputs among the four IGCC and two advanced Chemical Looping cases. The Gross power outputs for Cases 8, 9, 10, and 11 are about 301, 299, 292, and 290 MW, respectively, and for Cases 12 and 13 is 302 and 314 MW as shown in Table 5.2.1. The respective net outputs are shown in Figure 5.2.3. Clearly, Cases 9, 11 and 13 incur more degradation than their Base Case counterparts, due to the heavy demands of auxiliary power for gas processing and CO_2 compression.

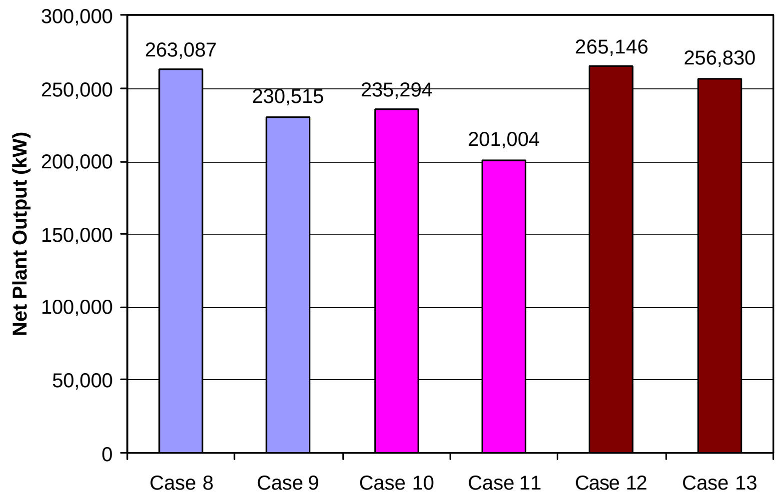


Figure 5.2. 3: Gasification Plants Net Plant Output Comparison

Plant Thermal Efficiency:

Figure 5.2.4 compares the overall plant thermal efficiency (HHV basis) among the four IGCC and two advanced Chemical Looping cases. The efficiency differences among these cases are a reflection of the combination of oxygen supply process, gasification process, CO_2 capture process, and auxiliary power requirements presented in Figure 5.2.3. The chemical looping cases, which avoid the energy intensive cryogenic ASU process, are significantly more efficient both with and without CO_2 capture.

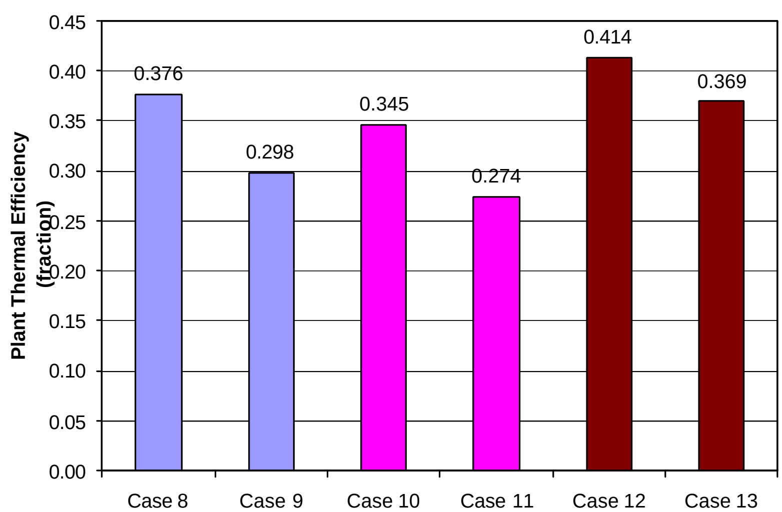


Figure 5.2. 4: Gasification Plants Thermal Efficiency Comparison

The energy penalty due to gas processing and CO_2 compression for Case 9 (Compared to Base Case 8) is 20.7 percent (from 0.376 to 0.298 fractional percent). The corresponding energy penalty for Case 11 (Compared to Base Case 10) is 20.6 percent (from 0.345 to 0.274 fractional percent). Similarly, the energy penalty due to gas processing and CO_2 compression for Case 13 (Compared to Base Case 12) is 10.8 percent (from 0.414 to 0.369 fractional percent). The magnitudes of these energy penalties are consistent with those found by EPRI and Parsons Energy and Chemical Group, Inc. (Holt, 2000 and 2003).

Plant CO_2 Emissions:

Figure 5.2.5 compares overall CO_2 emissions on a normalized basis (lbm/kWh) among the four IGCC and two advanced Chemical Looping cases. Base Cases 8, 10 and 12, with no CO_2 capture, show respective CO_2 emissions of 1.81, 1.98 and 1.71 lbm/kWh. These emissions are consistent with those obtained by other Researchers (e.g., Griffin, et al., 2001; Bozzuto, et al., 2001; Holt, 2000; Herzog, 2000) for coal firing. Cases 9, 11 and 13 were designed to capture a minimum of 90 percent of the CO_2 produced [e.g., for Case 9, CO_2 captured = 2.07/(2.07+0.23) = 90 percent), 2.07 and 0.23 lbm/kWh are the CO_2 captured and emitted, respectively].

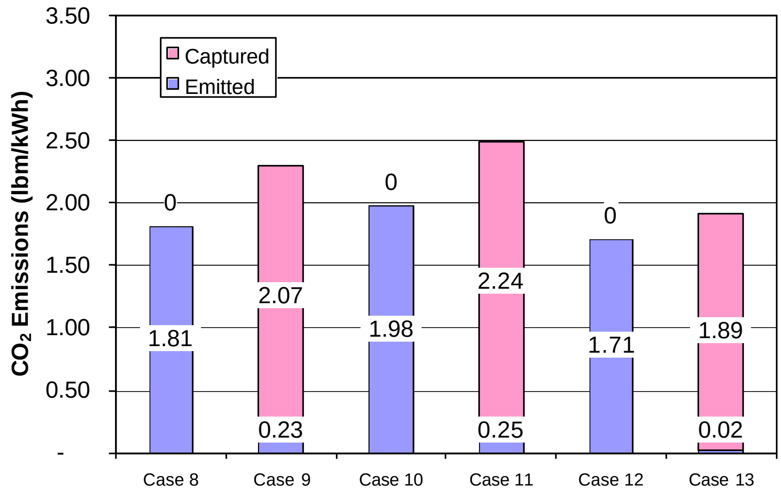


Figure 5.2. 5: Gasification Plants CO_2 Emissions Comparison

Figure 5.2.6 compares avoided CO_2 emissions on a normalized basis (i.e., in lbm/kWh) among the two IGCC and one advanced Chemical Looping cases, which entailed capturing the CO_2. The avoided quantities of CO_2 were 1.64, 1.78 and 1.69 lbm/kWh for Cases 9, 11 and 13, respectively.

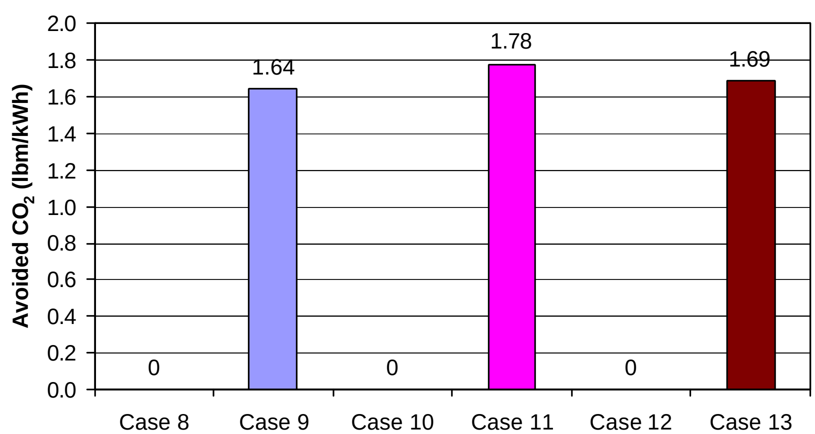


Figure 5.2. 6: Gasification Plants Avoided CO_2 Emissions Comparison

5.2.2. Gasification Cases: Costs and Economics Comparison

The plant investment costs for the four IGCC and two advanced Chemical Looping cases are shown in Table 5.2.2 and Figure 5.2.7. The plant investment cost (EPC basis) for the Texaco Base Cases (Cases 8 and 10) without CO_2 capture was 1,565 and 1,461 $/kW. The plant investment costs for the corresponding cases (Cases 9 and 11) with CO_2 capture was 2,179to 2,052 $/kW respectively. Case 13 (Chemical Looping Gasification) was found to be the lowest cost of the capture cases (1,383 $/kW) as compared to Case 12 without capture at 1,120 $/kW.

Table 5.2. 2: Investment Costs for Gasification Cases

Study Case	Net Plant Output, kW	Total Investment Cost, EPC Basis	
		$x1000	$/kW
Without CO2 Capture			
Case 8, Built & Operating IGCC w/o CO2 Capture	263,087	411,731	1,565
Case 10, Commercially Offered IGCC w/o CO2 Capture	235,294	341,468	1,451
Case 12, Chemical Lopping Gasification w/o CO2 Capture	265,146	296,991	1,120
With CO2 Capture			
Case 9, Built & Operating IGCC w/ CO2 Capture	230,515	502,330	2,179
Case 11, Commercially Offered IGCC w/CO2 Capture	201,004	412,377	2,052
Case 13, Chemical Lopping Gasification w/ CO2 Capture	256,830	355,132	1,383

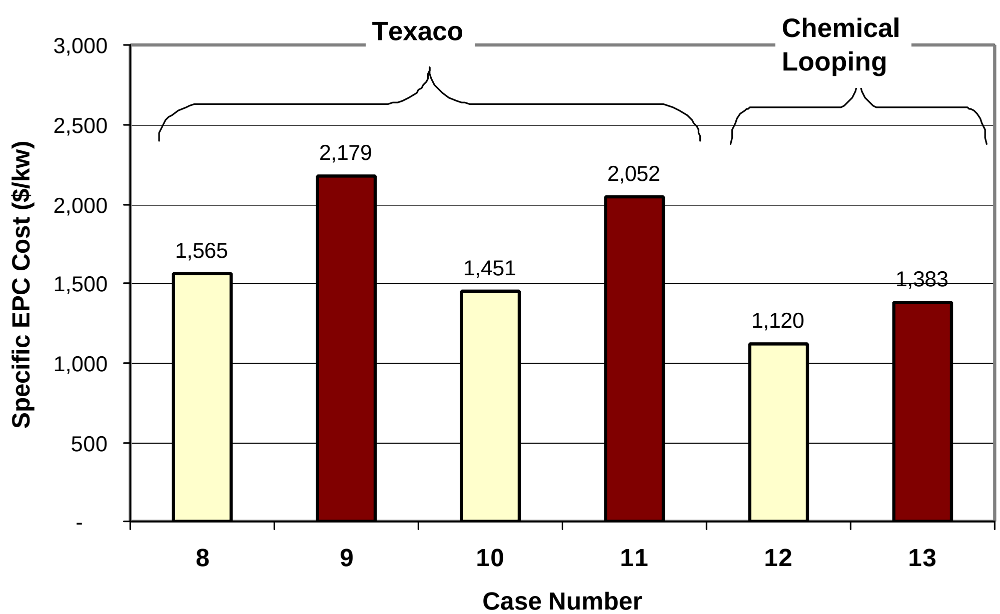


Figure 5.2. 7: Investment Costs for Gasification Cases

Total O&M costs for the gasification cases are shown in Table 5.2.3. The Base Cases (8, 10, and 12) ranged from about 0.7 to 1.0 cents/kWh while the capture cases (9, 11, and 13) ranged from about 1.0 to 1.4 cents/kWh.

Table 5.2. 3: Operating and Maintenance Costs for Gasification Cases

Study Case	Net Plant Output, kW	Operating & Maintenance (O&M) Costs					Total O&M, cents/kWh
		Fixed		Variable @ 80% Capacity		Total, $	
		$	$/kW	$	$/kWh		
Case 8, Built & Operating IGCC w/o CO_2 Capture	263,087	10,180,299	38.70	7,745,766	0.0042	17,926,065	0.97
Case 9, Built & Operating IGCC w/ CO_2 Capture	230,515	12,138,670	52.66	9,201,958	0.0057	21,340,627	1.32
Case 10, Commercially Offered IGCC w/o CO_2 Capture	235,294	9,343,766	39.71	6,899,778	0.0042	16,243,544	0.99
Case 11, Commercially Offered IGCC w/CO_2 Capture	201,004	11,067,713	55.06	9,110,706	0.0065	20,178,419	1.43
Case 12, Chemical Lopping Gasification w/o CO_2 Capture	265,146	6,487,709	24.47	5,989,858	0.0032	12,477,567	0.67
Case 13, Chemical Lopping Gasification w/ CO_2 Capture	256,830	7,915,922	30.82	9,888,018	0.0055	17,803,941	0.99

Table 5.2.4 and Figure 5.2.8 summarize the levelized economic analysis results for the four IGCC and two advanced Chemical Looping cases. In this figure the Base Cases without CO_2 capture are shown on the left and the capture cases on the right side of the figure.

Table 5.2. 4: Cost of Electricity for Gasification Cases

Study Case	Levelized Cost of Electricity (c/kWh)					Incremental COE (c/kWh)
	Financial	Fixed O&M	Variable O&M	Fuel	Total	
Without CO2 Capture						
Case 8, Built & Operating IGCC w/o CO2 Capture	3.20	0.55	0.42	1.13	5.30	
Case 10, Commercially Offered IGCC w/o CO2 Capture	3.00	0.57	0.42	1.24	5.22	
Case 12, Chemical Lopping Gasification w/o CO2 Capture	2.34	0.47	0.44	1.03	4.28	
With CO2 Capture						
Case 9, Built & Operating IGCC w/ CO2 Capture	4.40	0.75	0.57	1.43	7.15	1.85
Case 11, Commercially Offered IGCC w/CO2 Capture	4.19	0.79	0.65	1.56	7.18	1.95
Case 13, Chemical Lopping Gasification w/ CO2 Capture	2.85	0.55	0.66	1.16	5.22	0.93

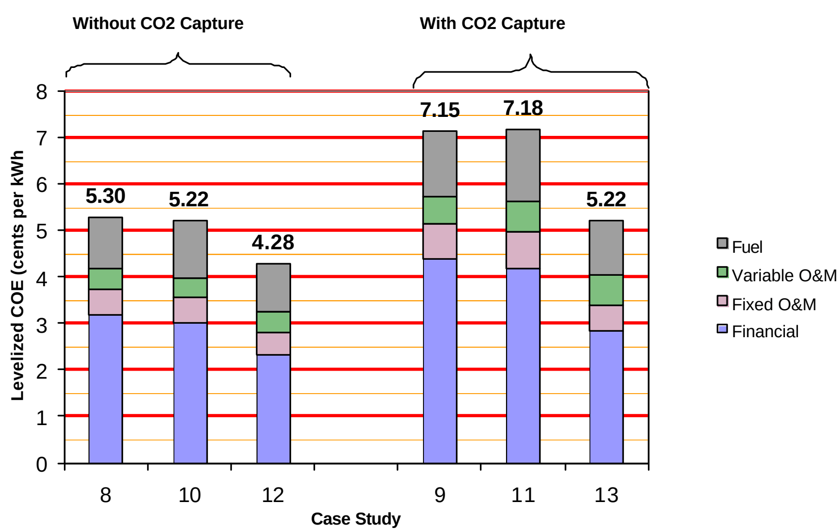


Figure 5.2. 8: Cost of Electricity for Gasification Cases

Case-13 (advanced Chemical Looping gasification with CO_2 capture) was found to be clearly the best alternative of the two IGCC and one advanced Chemical Looping CO_2 capture concepts studied based on levelized COE evaluation criterion (5.22 cents/kWh).

This case was found to be about 27 percent lower with respect to COE than the Texaco based IGCC cases. This case was found to have an incremental COE value of 0.93 cents/kWh as compared to Case 12 (Chemical Looping Gasification without CO_2 capture).

Case-13 was also found to be about 11 percent better, with respect to COE, than the best of the combustion cases (Case-7; Chemical Looping Combustion; 5.84 cents/kWh).

5.3. Comparison of Combustion Cases and Gasification Cases

This section summarizes in a comparative manner some of the more important plant performance, CO_2 emission, investment cost, and economic results obtained from Combustion and Gasification cases. These comparative results are briefly discussed below.

Gas Processing System Auxiliary Power:

The auxiliary power consumed by the Gas Processing Systems (GPS), for both Combustion and Gasification cases, was one of the more important parameters used to calculate overall system performance. This was already shown in Tables 5.1.1 and 5.2.1 for the pertinent Combustion and Gasification cases respectively. Figure 5.3.1 compares GPS specific power consumption for Combustion and Gasification cases. The GPS auxiliary power for combustion cases ranged from 127 to 154 kWh/ton-CO_2 captured. Both Texaco based IGCC Cases consumed 114 kWh/ton-CO_2 captured. The Texaco based IGCC cases used less auxiliary power, because they process the majority of their gas from 50 psig, as compared to the combustion cases and the Chemical Looping Gasification case, which process all of their gases from near atmospheric pressure.

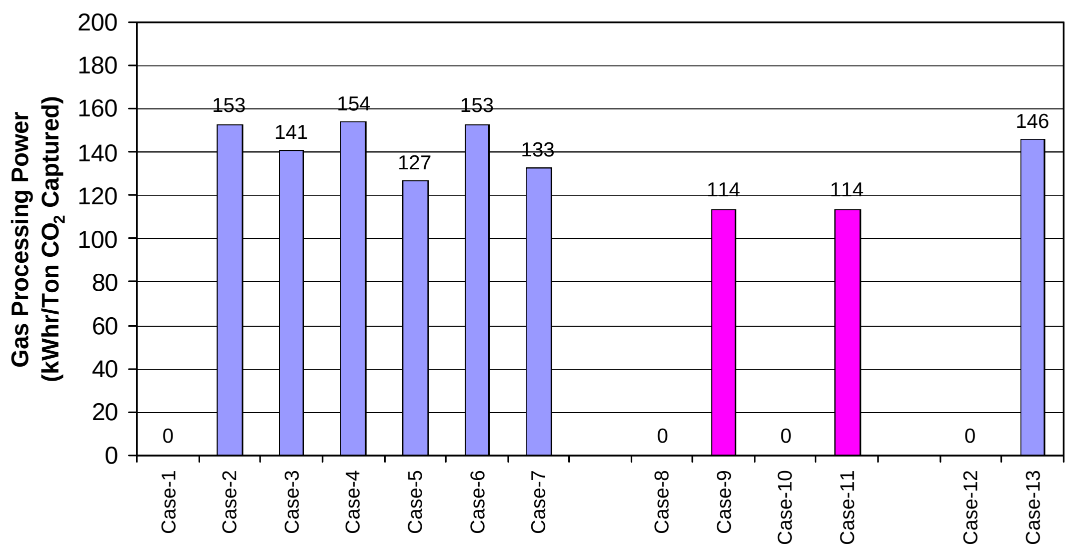


Figure 5.3. 1: Gas Processing Power Comparison between Combustion and Gasification Cases

Net Plant Heat Rate:

Figure 5.3.2 shows a comparison of net plant heat rate (HHV basis) between the Combustion and Gasification Cases, both with and without CO_2 capture.

For the cases without CO_2 capture, the Base combustion case (Case-1; 9,611 Btu/kWh) falls between the two Texaco based Base IGCC cases (Cases 8 and 10; 9,069 and 9,884 Btu/kWh respectively) and Case 12 (8,248 Btu/kWh) is clearly the most efficient case studied.

For the cases with CO_2 capture the advanced combustion cases (Cases 5, 6, and 7) are slightly more energy efficient than the Texaco Based IGCC cases (Cases 9 and 11) but not nearly as efficient as Case-13 (9,249 Btu/kWh), the Chemical Looping Gasification case.

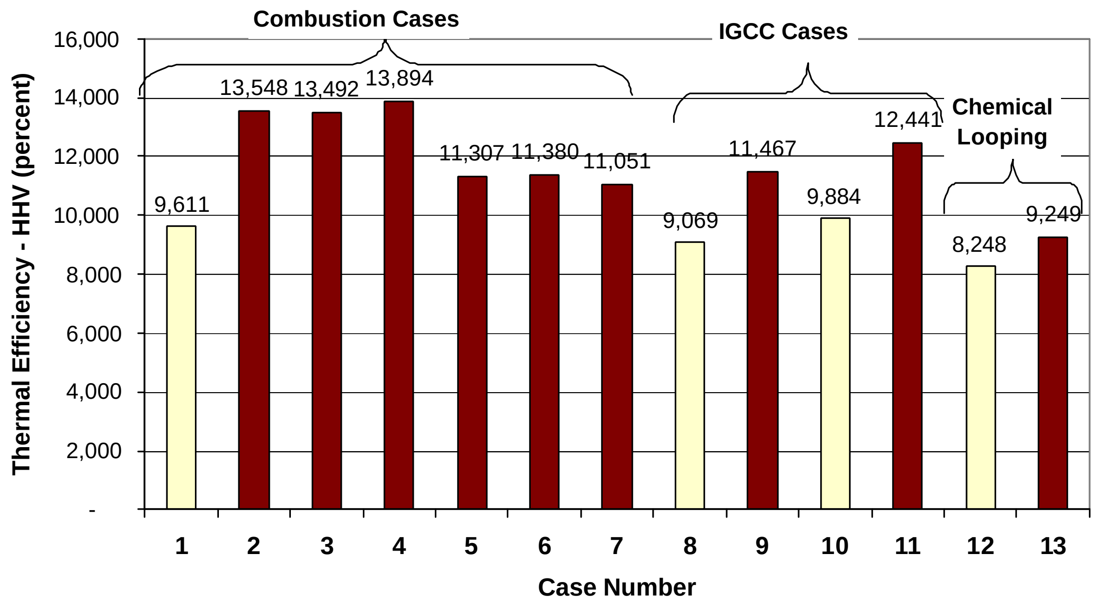


Figure 5.3. 2: Net Plant Heat Rate (HHV) Comparison between Combustion and Gasification Cases

CO_2 Emissions:

Figure 5.3.3 shows a comparison of CO_2 emissions between the Combustion and Gasification Cases, both with and without CO_2 capture while Figure 5.3.4 shows CO_2 avoided. Avoided emissions were calculated with respect to the appropriate Base Case. All cases captured a minimum of 90 percent of the CO_2 emitted.

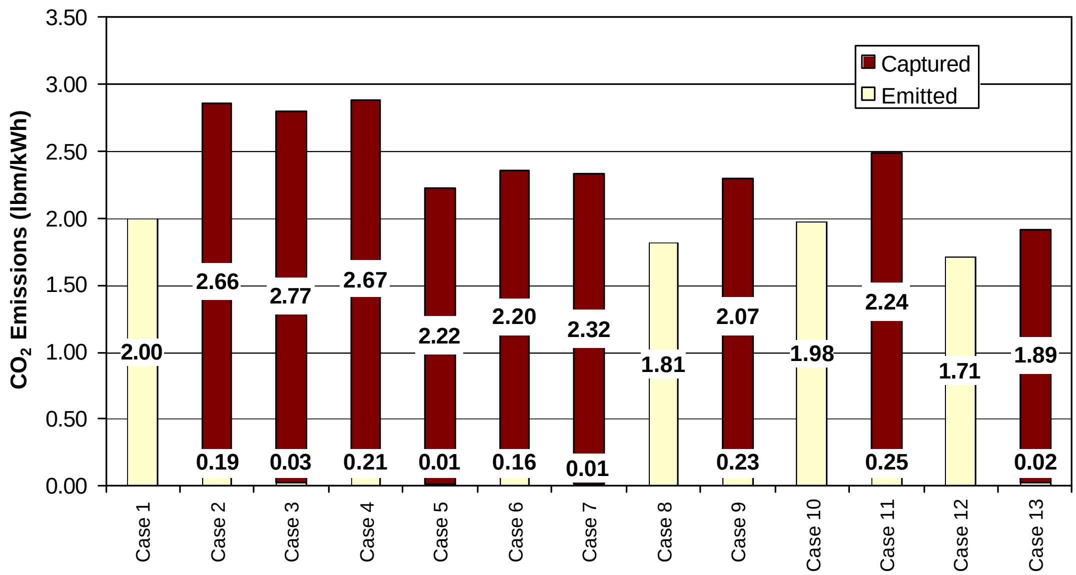


Figure 5.3. 3: CO_2 Emissions Comparison between Combustion and Gasification Cases

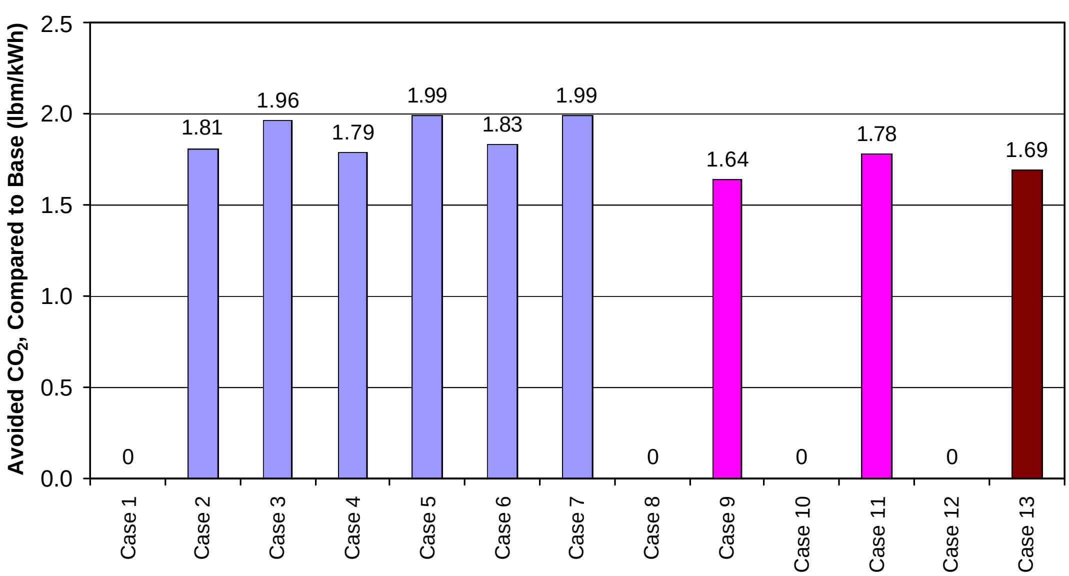


Figure 5.3. 4: Avoided CO_2 Emissions Comparison between Combustion and Gasification Cases

Costs and Economics:

Figure 5.3.5 shows a comparison of overall plant EPC costs between the Combustion, IGCC, and advanced Chemical Looping cases both with and without CO_2 capture. The yellow shaded bars are "Base Cases" without CO_2 capture and the brown shaded bars are with capture.

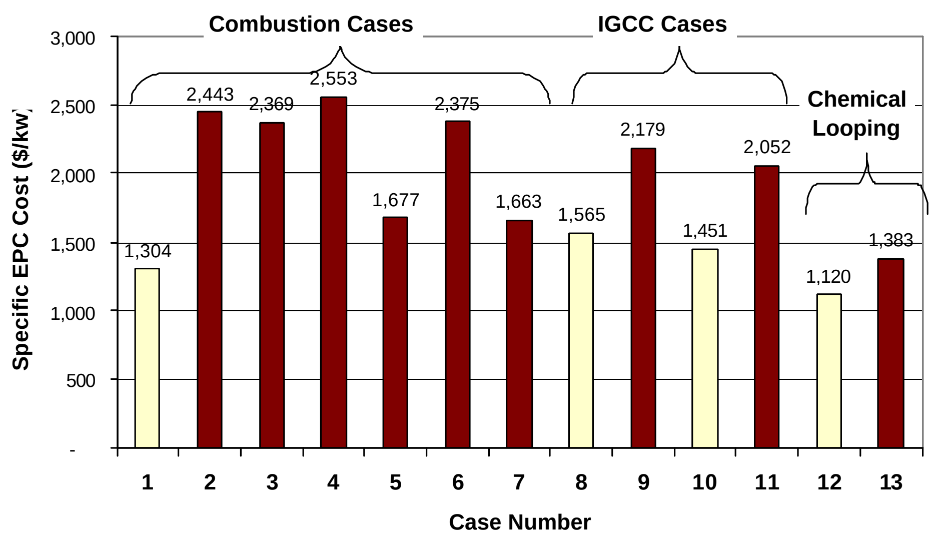


Figure 5.3. 5: Overall Plant EPC Cost Comparison between Combustion and Gasification Cases

Figure 5.3.6 shown below provides a comparison of levelized Cost of Electricity (COE). Cases on the left side of Figure 5.3.6 are "Base Cases" without CO_2 capture and those on the right are with capture. Case 13 was calculated to be the best of the capture cases.

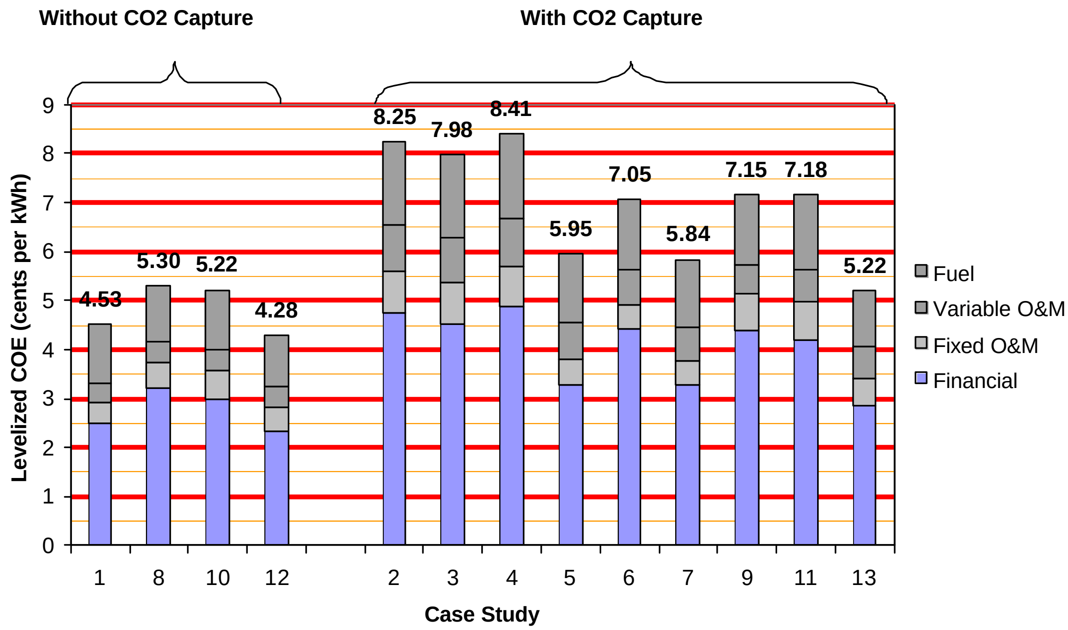


Figure 5.3. 6: Cost of Electricity Comparison between Combustion and Gasification Cases

Figure 5.3.7 shown below provides a comparison of CO_2 mitigation costs. Case 13 (advanced Chemical Looping) was the lowest at 11 \$/Ton of CO_2 avoided. Case 7 and Case 5 were the next best cases at 13 and 14 \$/Ton of CO_2 avoided respectively.

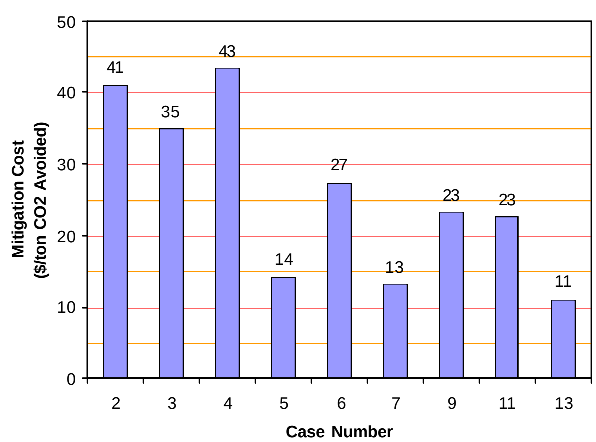


Figure 5.3. 7: CO_2 Mitigation Cost Comparison between Combustion and Gasification Cases

Conclusions:

In a carbon constrained region the following conclusions can be drawn with respect to investment costs and levelized costs of electricity (COE):

The advanced Chemical Looping gasification plant concept (Case-13) is the least costly of all the concepts considered: its EPC (engineered, procured, and constructed) capital cost and levelized cost of electricity are 1,380 $/kW and 5.2 cents/kWh, respectively. This cost of electricity for advanced Chemical Looping case is nearly equivalent to new coal-fired plants offered today without CO_2 capture (i.e., Case 1 @ 4.5 cents/kWh).

The Carbonate Regeneration Process (Case-5) and Chemical Looping Combustion CMB (Case-7) are less capital-intensive than the built and operating (Case-9) or commercially offered (Case-11) IGCC's (e.g., 1,680-1,660 vs. 2,180-2,050 $/kW). Hence, their levelized costs of electricity are correspondingly lower higher (e.g., 5.9 - 5.8 vs. 7.2 cents/kWh).

The cryogenic oxygen-fired CFB or CMB plants (Cases 2, 3, and 4) are more capital intensive than the built and operating or commercially offered IGCC's (Cases 9 and 11), because they incur high energy penalties by comparison (e.g., 2,350-2,440 vs. 2,180-2,050 $/kW). Consequently, their levelized costs of electricity are, correspondingly, higher (e.g., 8.3 - 8.4 vs. 7.2 cents/kWh)

The capital investment of an oxy-fuel fired plant designed to produce "for Sequestration only flue gas" (Case-3) is approximately 3 percent lower than that of an oxy-fuel fired CFB (Case-2) designed to produce a CO_2 product suitable for EOR application (2,370 vs. 2,440 $/kW). Consequently, its levelized cost of electricity is, correspondingly, 3 percent lower (8.0 vs. 8.3 cents/kWh)

In a region without carbon constraints the following conclusions can be drawn with respect to investment costs and levelized costs of electricity (COE):

The Texaco IGCC power plant technology (Cases 8 and 10) is about 11-20 percent more capital-intensive than the air-fired CFB (Case-1) technology (e.g., 150 - 250 $/kW). This is due to the fact that an IGCC power plant is very complex, comprised of many specialized components operating at elevated pressures.

The air-fired CFB plant (Case-1) produces electricity about 9-15 percent cheaper in terms of COE than the Texaco IGCC plants (e.g., 4.5 vs. 5.2 - 5.3 cents/kWh)

The advanced Chemical Looping gasification plant concept (Case-12) shows the potential to provide electricity at a lower cost than both air-fired CFB and traditional IGCC plants (e.g., 4.3 vs. 4.5 & 5.3 cents/kWh). It is noted that the advanced Chemical Looping plant concept is in a very early stage of development. Hence, the COE estimates for this concept are preliminary.

5.4. Comparisons With Other Work

This section compares the results from this study to selected results from similar work from the literature. Table 5.4.1 summarizes in a comparative manner the performance, CO_2 emissions, and cost information obtained from this study and that from the literature. This information covers the following power plant types: Integrated Gasification Combined Cycle (IGCC), Pulverized Coal (PC) Fired, Circulating Fluidized Bed (CFB) Boiler-Based Combustion Concepts, and Natural Gas Combined Cycle (NGCC). The CO_2 capture technologies used were Selexol, MDEA, and oxy-fuel Firing. The PC plants evaluated were mostly sub-critical, although there were a couple of instances where they were super-critical, as shown in Table 5.4.1. Details are given in the literature cited in Table 5.4.1. Selected data in Table 5.4.1 are plotted in Figures 5.4.1-5.4.6.

Table 5.4. 1: Performance and Cost Comparison between This Work and Literature Results

Technology	Reference Plants	CO_2 Capture Technology	NPHR, Btu/kWh, LHV	Net Plant Output Mwe	Avoided CO_2 Emissions, kg/kWh	CO_2 Emissions, kg/kWh	Capital Cost, $kW	Plant Thermal Eff., % LHV
			All Plants					
Reference Plants without CO_2 Capture								
IGCC Plants	Argonne, Doctor, et al., 1997	N/A	8,938	414	N/A	0.790	1,332	38.2
	Milan, Chiesa et al., 1998		7,817	404		0.709	1,536	43.7
	SFA Pacific, Simbeck, 1998		7,210	400		0.674	1,300	47.3
	Utrecht, Hendricks, 1994		7,826	600		0.760	1,265	43.6
	EPRI, Condrelli, 1991		9,280	432		0.868	1,600	36.8
	IEA, Stork Engineering Consultancy, 1999		7,369	408		0.710	1,471	46.3
	Herzog, 1999		8,124	500		0.752	1,401	42.0
	EPRI & Parsons, 2000, H-class GT		7,611	424.5		0.718	1,420	44.8
	This Work, Case 8, F-class GT		8,720	263		0.844	1,565	39.1
	This Work, Case 10, F-class GT		9,504	235		0.921	1,451	35.9
PC Plants	SFA Pacific, Simbeck, 1998)	N/A	7,680	400	N/A	0.717	1,300	44.4
	Utrecht, Hendricks, 1994		8,322	600		0.800	1,150	41.0
	EPRI, Smelser et., 1991; Booras & Smelser, 1991		9,440	513		0.909	1,129	36.2
	IEA, Stork Engineering Consultancy, 1999		7,482	501		0.722	1,022	45.6
	Herzog, 1999		8,462	500		0.789	1,150	40.3
	EPRI & Parsons, SC 2000		8,097	462		0.734	1,281	42.2
	EPRI & Parsons, USC, 2000		7,677	506		0.746	1,301	42.7
	Bozzuto, et al. 2001- Existing Unit		9,309	434		0.906		36.7
CFB Plant	This Work, Case 1	N/A	9,241	193	N/A	0.907	1,304	36.9
NGCC Plants	Milan, Chiesa et al., 1998	N/A	6,400	373	N/A	0.374	531	53.3
	SFA Pacific, Simbeck, 1998)		5,688	400		0.300	485	60.0
	IEA, Stork Engineering Consultancy, 1999		6,071	790		0.370	414	56.2
	Trondheim, Bolland and Saether, 1992		6,536	721		0.400	754	52.2
	Herzog, 1999		6,308	500		0.337	525	54.1
	EPRI & Parsons, 2000, F-class GT		6,136	509		0.364	549	55.6
		CO_2 Capture Plants						
IGCC Plants	Argonne, Doctor, et al., 1997	Selexol	9,791	378	0.614	0.176	1,687	34.9
	Milan, Chiesa et al., 1998		9,140	346	0.639	0.070	1,913	37.3
	SFA Pacific, Simbeck, 1998)		9,173	314	0.586	0.088	1,767	37.2
	Utrecht, Hendricks, 1994		9,399	500	0.720	0.040	1,799	36.3
	EPRI, Smelser et., 1991; Booras & Smelser, 1991		11,528	347	0.858	0.011	2,152	29.6
	IEA, Stork Engineering Consultancy, 1999		8,932	382	0.576	0.134	2,204	38.2
	EPRI & Parsons, 2000, H-class GT		8,871	404	0.664	0.088	1,909	38.5
	Herzog, 1999		9,639	421	0.645	0.074	1,844	35.4
	This Work, Case 9, F-class GT		11,026	231	0.744	0.104	2,179	31.0
	This Work, Case 11, F-class GT		11,963	201	0.807	0.113	2,052	28.5
PC Plants	SFA Pacific, Simbeck, 1998)	Amine	9,130	337	0.589	0.128	2,022	37.4
	Utrecht, Hendricks, 1994		10,832	462	0.700	0.100	2,073	31.5
	EPRI, Smelser et., 1991; Booras & Smelser, 1991		14,331	338	0.771	0.138	2,484	23.8
	IEA, Stork Engineering Consultancy, 1999		10,339	362	0.574	0.148	1,856	33.0
	Herzog, 1999		10,581	400	0.774	0.015	2,090	32.3
	Bozzuto, et al. 2001 - Retrofit	MEA	15,872	255	0.847	0.059	1,602	21.5
	Bozzuto, et al. 2001 - Retrofit	Oxyfuel	14,500	273	0.818	0.088	1,042	23.5
	Bozzuto, et al. 2001 - Retrofit	MEA/MDEA	14,395	336	0.781	0.124	2,197	23.7
	EPRI & Parsons, SC, 2000	Amine	11,362	329	0.626	0.108	2,219	30.0
	EPRI & Parsons, USC, 2000		10,576	367	0.686	0.060	2,175	31.0
CFB Plants	This Work, Case 2	Oxyfuel	13,231	135	0.821	0.086	2,481	25.8
	This Work, Case 3		12,973	135	0.890	0.014	2,369	26.3
	This Work, Case 4		13,360	132	0.812	0.095	2,553	25.5
	This Work, Case 5	High Temp. Carbonate Reg.	10,872	161	0.903	0.005	1,677	31.4
	This Work, Case 6	Oxyfuel	10,942	197	0.830	0.073	2,375	31.2
	This Work, Case 7	Chemical Looping	10,626	164	0.903	0.005	1,663	32.1
NGCC Plants	Milan, Chiesa et al., 1998	Amine	7,097	337	0.337	0.037	807	48.1
	SFA Pacific, Simbeck, 1998)		6,433	354	0.244	0.056	1,135	53.1
	IEA, Stork Engineering Consultancy, 1999		7,229	663	0.276	0.061	786	47.2
	Trondheim, Bolland and Saether, 1992		7,667	615	0.318	0.046	1,317	44.5
	Herzog, 1999		7,293	432	0.295	0.042	1,013	46.8
	EPRI & Parsons, 2000, F-class GT		7,836	399	0.600	0.045	1,099	43.6

Net Plant Heat Rates:

Figure 5.4.1 shows a comparison of net plant heat rates (expressed on Lower Heating Value, LHV, basis) between the Combustion and Gasification cases of this study. Figure 5.4.2 shows a comparison of NPHR's for the IGCC, Super-Critical and Ultra-Super Critical PC, and NGCC cases studied by EPRI & Parsons (Holt, 2000). These results indicate the following:

- The Net Plant Heat Rate (NPHR) of the Base Case, air-fired CFB (Case 1) for this work, is 9,241 Btu/kWh (LHV basis). The corresponding values for the combustion-based cases with CO_2 capture (Cases 2-7) ranged from 13,231 to 10,626 Btu/kWh. Hence, Chemical Looping (Case 7) is the most efficient of these CO_2 capture options. Case-5 (air fired high temperature carbonate regeneration) and Case-6 (OTM O_2-fired CMB) are nearly as efficient as Case-7. The cryogenic based O_2 fired cases (2, 3, 4) are significantly less efficient.

- The Net Plant Heat Rate (NPHR) of the Base Case, Built & Operating IGCC (Case 8) is 8,720 Btu/kWh. The corresponding value for this IGCC, when integrated with a water gas shift reactor to capture CO_2 (Cases 9) is 11,026 Btu/kWh. The performance of commercially offered IGCC with and without CO_2 capture (Cases 10 and 11) are slightly higher due primarily to quench cooling at 9,504 and 11,963 Btu/kWh, respectively.

- EPRI's and Parsons' IGCC results with and without CO_2 capture show NPHR's of 7,611and 8,871 Btu/kWh, respectively. The Parsons IGCC is more efficient than the one used in Cases 8 and 9 of this work due to several factors. The most important factor is due to differing gas turbines, however, many other factors also contribute as described below.

  The gas turbines used in this work (cases 8-11) were all F-Class units, whereas the EPRI & Parsons IGCC cases used H-Class units. Different ambient conditions, 80 ^oF dry bulb for this work and 63 ^oF dry bulb for EPRI & Parsons, also contributes to the difference. Different coal analyses and coal gasification processes (E-Gas^{TM} for Parsons and Texaco for this work) as well as differing condenser pressures (3.0 in. Hga for this work and 2.0 in. Hga for EPRI & Parsons) also contributed to IGCC case performance differences.

Taking all these differences into account, the results for the IGCC cases appear very consistent, as would be expected, since Parsons provided the process simulations for both studies.

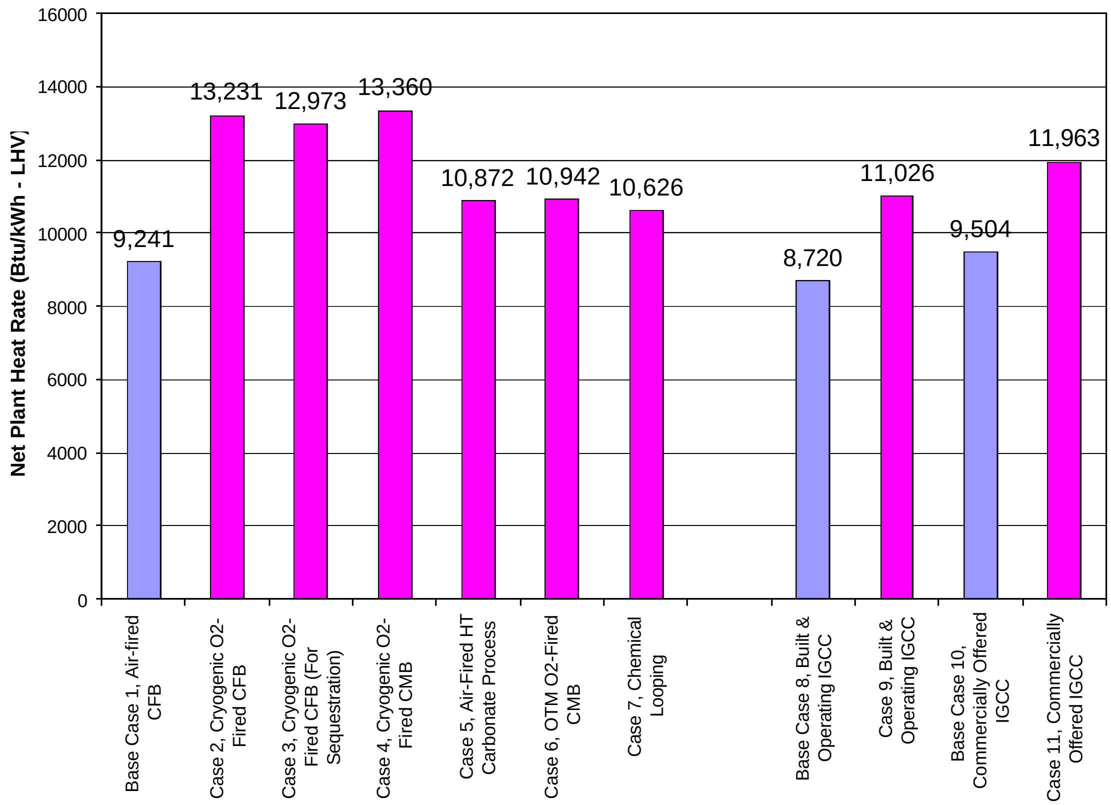


Figure 5.4. 1: Net Plant Heat Rates (LHV basis) from This Work

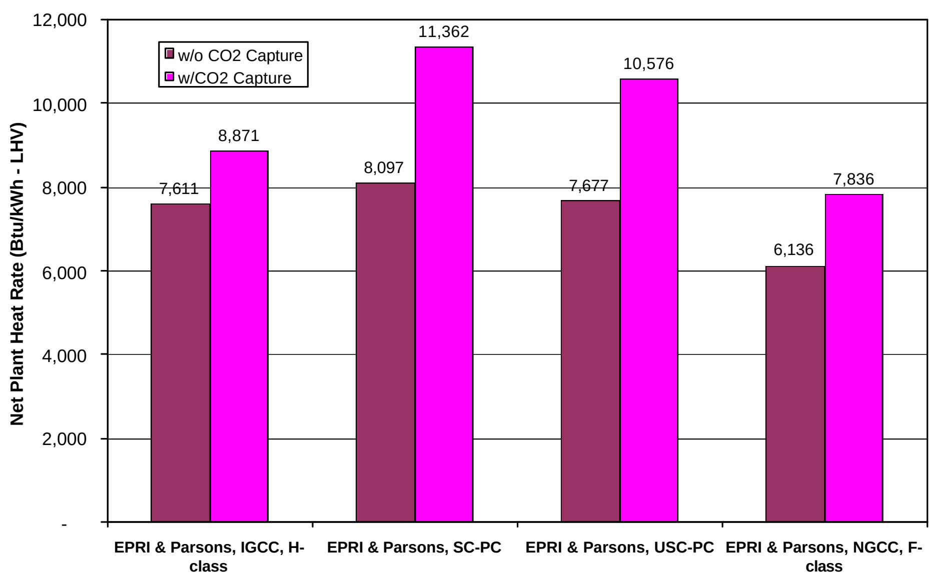


Figure 5.4. 2: Net Plant Heat Rates (LHV basis) from EPRI & Parsons (Holt, 2000)

Figures 5.4.3 and 5.4.4 show comparisons of energy penalty and CO_2 emissions respectively plotted against NPHR. As expected, energy penalties for the advanced combustion cases (Cases 5, 6, and 7) of this work, with NPHR's slightly below 11,000 Btu/kWh are significantly lower than other PC or CFB based cases. All CO_2 emissions correlate quite well with NPHR as would be expected.

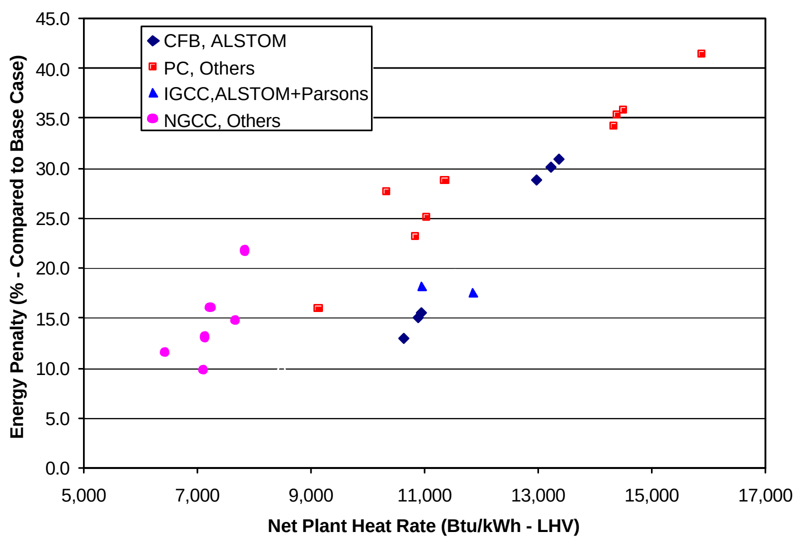


Figure 5.4. 3: Energy Penalty Comparison with Other Work

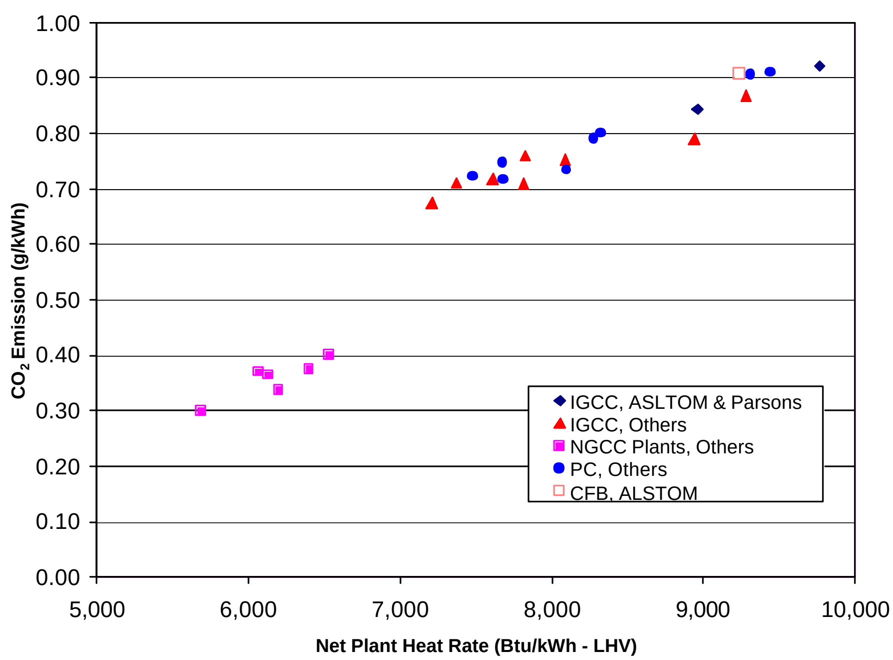


Figure 5.4. 4: CO_2 Emissions Comparison with Other Work

Figure 5.4.5 shows a comparison of plant investment costs between this work and literature for IGCC plants without CO_2 capture. Similarly, Figure 5.4.6 shows a comparison of plant investment costs between this work and literature for IGCC plants with CO_2 capture.

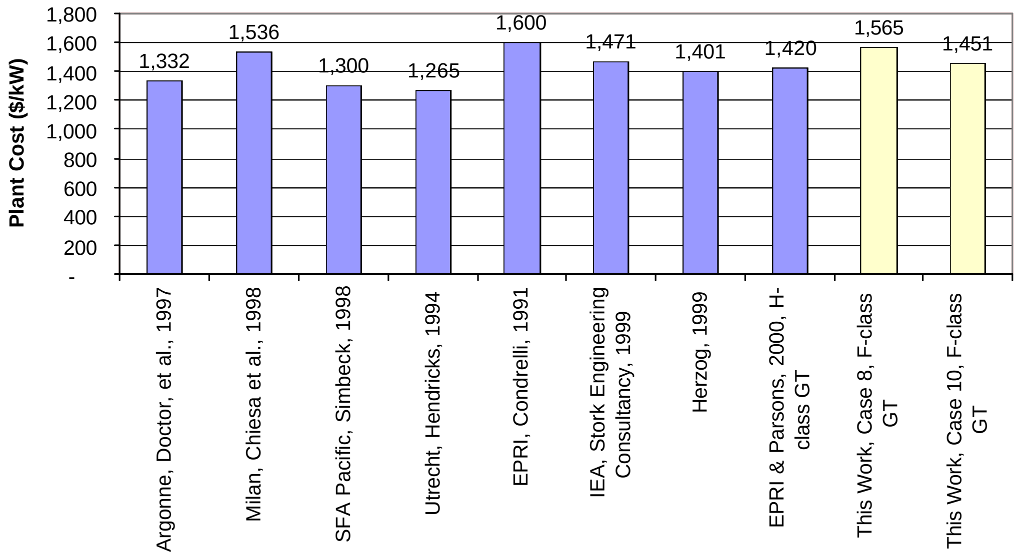


Figure 5.4. 5: IGCC's w/o CO_2 Capture - Comparison with Other Work

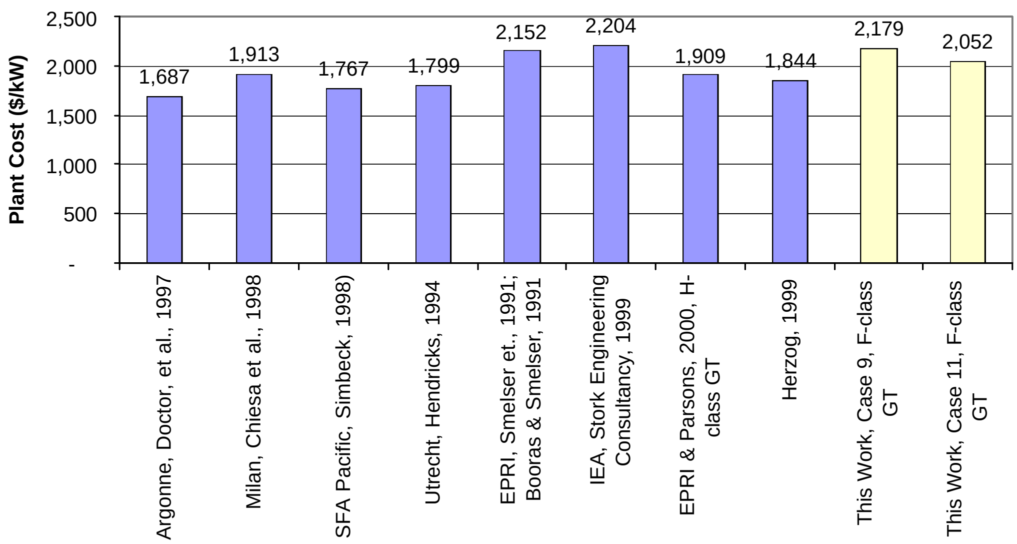


Figure 5.4. 6: IGCC's with CO_2 Capture - Comparison with Other Work

6. COMMERCIAL APPLICATION OF O_2-FIRED CIRCULATING FLUIDIZED BED BOILER (CFB) TECHNOLOGY

6.1. Background

ALSTOM Power Inc. (ALSTOM) held, on November 19, 2002, a project review meeting at the DOE NETL's (DOE's) offices in Pittsburgh, PA. This review consisted of a Phase I progress report and a recommendation, based on Phase I results up to that point, on how to proceed in the future. ALSTOM recommended that the DOE continue to develop technologies with short and long-term commercialization potentials.

For long-term ALSTOM recommended a hybrid combustion-gasification chemical looping concept with CO_2 capture, due to its potentially ultra-clean, low cost and high efficiency coal power generation (Case-7 Combustion, Case 13 Gasification). The development of this technology, which has a long-term commercialization potential, has been addressed by ALSTOM through a proposal responding, in February 2003, to the DOE NETL's Solicitation No. DE-PS26-02NT41613-01 (Hybrid Combustion-Gasification Chemical Looping Coal Power Technology Development).

For short-term, ALSTOM recommended the development of Oxygen-Fired CFB technology (i.e., Case 2) for capturing CO_2 from coal or delayed petroleum coke for Enhanced Oil Recovery (EOR) application. In this technology, a modified CFB (circulating fluid bed) type boiler is fired with pure oxygen plus recirculated flue gas (mainly CO_2) instead of atmospheric air, as is current practice. This results in a flue gas stream with high CO_2 concentration (~90 percent by volume, as compared to ~15 percent for air-firing). It follows that the CO_2 can be separated from this flue gas stream relatively easily for use or sequestration. The results from Case 2 lead to the conclusion that that further work is justified. In summary this recommendation was made for the following reasons:

- Case 2, which would be a green field plant, is the most near-term solution (~5-year horizon), as it uses <u>enabling</u> technologies, which are readily available commercially, for example:
  - Oxygen production by cryogenic air separation
  - CO_2 capture, purification, liquefaction, and compression

- Preliminary economic analysis looked viable for commercial EOR application
  - CO_2 sale for oil field stimulation
  - N_2 sale for oil field pressurization

- Oxy-fuel firing in CFB's offers added emissions benefits over air firing, as had been preliminarily shown through combustion testing in an ALSTOM's bench-scale fluidized bed combustor:
  - CO_2 in the flue gas is highly concentrated (~90 percent vs.~15 percent), thus making the processing of this stream to achieve the required CO_2 purity for EOR application relatively cheaper.
  - Typically low NOx emissions in combustion-staged air-fired CFB's are further reduced due primarily to elimination of thermal NOx.
  - SO2 emissions reductions of up to 90 percent with sorbent utilization should not be negatively impacted. Furthermore, ALSTOM has a commercial product called "Flash Drier Absorbent (FDA)," which has been successfully demonstrated in the MTF to reduce SO2 emissions by as much as 99 percent.

- Unburned carbon (UBC) loss should not be negatively impacted.

The DOE concurred with ALSTOM's recommendation of developing Oxygen-Fired CFB technology for capturing CO_2 from coal or delayed petroleum coke for Enhanced Oil Recovery (EOR) application and, hence, authorized, as a first step, the implementation of Phase II of the current project. Phase II, which started on May 16, 2003, entails testing of two coals and one delayed petroleum coke in ALSTOM's pilot-scale Multiuse Test Furnace (MTF) at firing rates in the 4.0 – 9.5 MM-Btu/hr, and using this information to refine the design, performance, and economic analyses of the commercial-scale oxy-fuel fired CFB concept of Case 2.

This section addresses three areas of particular importance to the oxy-fuel fired CFB technology for commercial EOR application: (1) Potential early EOR site for applying this technology; (2) Technical and economic potentials of the technology; and (3) timeline vision for commercialization of the technology.

6.2. Potential Early EOR Site

Lost Hills oil field, located about 45 miles northwest of Bakersfield, CA, (Figure 6.2.1) was discovered in 1910 (Walker and Perri, 2002). There are 2.2 billion barrels of oil in place in the Belridge Diatomite in Lost Hills. While this diatomite has high oil saturation and high porosity, it has low permeability, which has led to low primary oil recovery. Consequently, to date only 112 million barrels of oil have been produced (i.e., 5 percent of the original oil) from the Belridge diatomite.

Chevron initiated a pilot diatomite water injection flood project in 1992, which has increased the oil production rate by about 50 percent (i.e., from about 6,400 to 10,400 Bbl/day). In 2000, Chevron initiated a pilot CO_2 flooding of Lost Hill diatomite, because CO_2 injectivity is two to three times more effective than water or steam, due mainly to two favorable mechanisms: (1) it reduces the reservoir's oil viscosity; and (2) it increases fluid expansion. This pilot work is being carried out under the auspices of the Department of Energy to address two main economic uncertainties associated with CO_2 injection, namely, oil response and CO_2 utilization required for such response.

Figure 6.2. 1: Location Map of Major Oil Fields in Southern San Joaquin Valley, CA. Lost Hills is Highlighted (from Walker and Perri, 2002)

This proposed project, referred to as the Bakersfield Project, is located on/or near AERA Energy LLC oil production property (a Limited Liability Corporation for oil production in California for Exxon/Mobil and Shell Oil companies). It presents an opportunity whereby relatively cheap delayed petroleum coke from oil refineries owned by AERA LLC partners could be used to generate CO_2 for EOR.

Plasma, Inc. (PLASMA), a project developer with expertise in oil/gas production projects (Schuller 2002), has provided ALSTOM with information from the CO_2 Users Coop Group and financial data on CO_2 use for EOR, particularly with respect to the Bakersfield, CA, project. Furthermore, PLASMA has requested, on behalf of their clients, that ALSTOM propose CFB boiler designs that are in sufficient detail to get material take-off cost estimates. These designs would be done for a 20,300 Ton/day steam production and ~210 MWe power generation from 3,000 tons/day delayed petroleum coke. These designs would also include marketable CO_2 recovery using a "zero-emission" boiler design from the same 3,000 tons/day coke with oxygen and CO_2 recycle as the combustion medium. At the completion of this work PLASMA should be able to determine the boiler system required and to finalize the emission profile needed for permit application purposes.

Based on favorable results from a bench-scale FBC study, performed under the auspices of Plasma, Inc., ALSTOM recommended that work on the next phase of Plasma's overall plan proceed, namely, to do design studies of commercial air-fired and O_2/CO_2-fired CFB boilers in sufficient detail to permit material take-off cost estimates to be made. The total capital investment cost estimates would be ±10 percent for the air-fired CFB's and ±25 percent for the oxy-fuel fired CFB's. The following three tasks were recommended:

- The first task would entail conducting a design study of two similar Greenfield CFB boilers (both boilers are conventional air-fired CFB boilers, one is for producing steam for use in enhanced oil recovery (EOR) at AERA's oil production property and the other is for producing electricity for sale). Each CFB boiler would have a total firing capacity of approximately 1,500 T/D of Mobil Torrance (or equivalent) delayed petroleum coke.

- The second task would deal with a retrofit design study of the CFB boilers developed in Task 1 for conversion to oxy-fuel firing.

6.3. Performance and Economic Analyses

The performance and economic analyses obtained from this evaluation for EOR application are presented below.

6.3.1. Performance Analysis

Table 6.3.1 summarizes the results of the Net Plant Heat Rates and Net Plant Outputs obtained from Case 1 (Air-Fired CFB without CO_2 Capture), and Case 2a (Air-Fired CFB Retrofit to O_2-Firing w/ CO_2 Capture).

Figure 6.3.1 shows a comparison of Net Plant Heat Rates between Cases 1 and 2a. The Net Plant Heat Rates of the air fired CFB and the retrofitted CFB are 9,611 and 14,660 Btu/kWh, corresponding to net plant efficiencies of 35.5 and 23.3 percent. The main reason for the high energy penalty (34 percent) associated with Case 2a, compared to Case 1, is the integration into the power plant of both the Air Separation Unit (ASU) to provide combustion oxygen, and the Gas Processing System (GPS) to capture, clean-up, compress, and liquefy the CO_2 product. Both these systems require large quantities of auxiliary power.

Table 6.3. 1: Summary of Results

Study Case		Fuel-Type and Cost		Net Plant Heat Rate, Btu/kW	Net Plant Output, kW	Total EPC Investment Costs, $/kW	Levelized Cost of Electricity (Cents/kWh)					
#	Description	Type	Cost, $/MMBtu				Total O&M	Financial	Fuel	CO_2 Credit	N_2 Credit	Total
1	Air-Fired CFB w/Pet. Coke, and w/o CO_2 Capture	Delayed Petroleum Coke	0.65	9,611	193,037	1,304	0.83	2.49	0.62	0.00	0.00	3.95
2a	O_2-Fired CFB Retrofit w/Pet. Coke, and w/ Captured CO_2 & N_2 Credits	Delayed Petroleum Coke	0.65	14,660	128,075	2,766	1.88	5.25	0.95	-2.41	-1.72	3.95

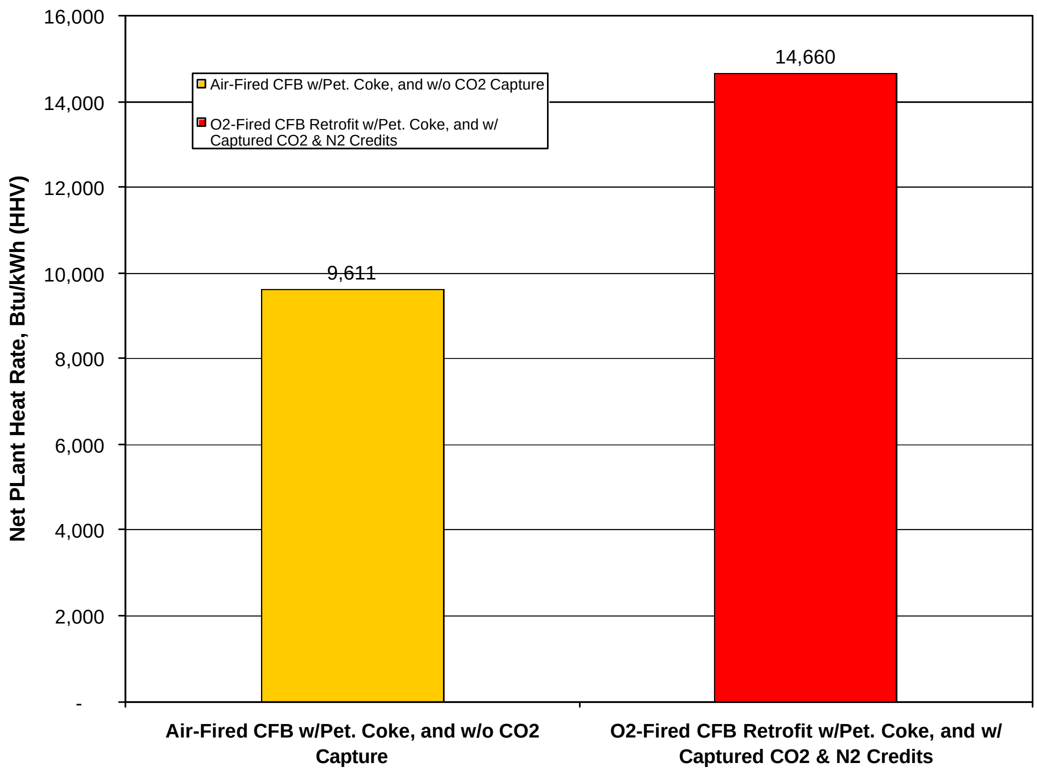


Figure 6.3. 1: Net Plant Heat Rate Comparison between Air-Fired and Oxy-fuel Fired CFB Plants

6.3.2. Economic Analysis

The total EPC (Engineered, Procured, and Constructed) plant costs for Cases 1 and 2a are 1,304 and 2,776 $/kW, respectively as shown in Table 6.3.1. The high EPC costs of

Case 2a are, as stated above, a direct reflection of integration into the power plant of the ASU and GPS.

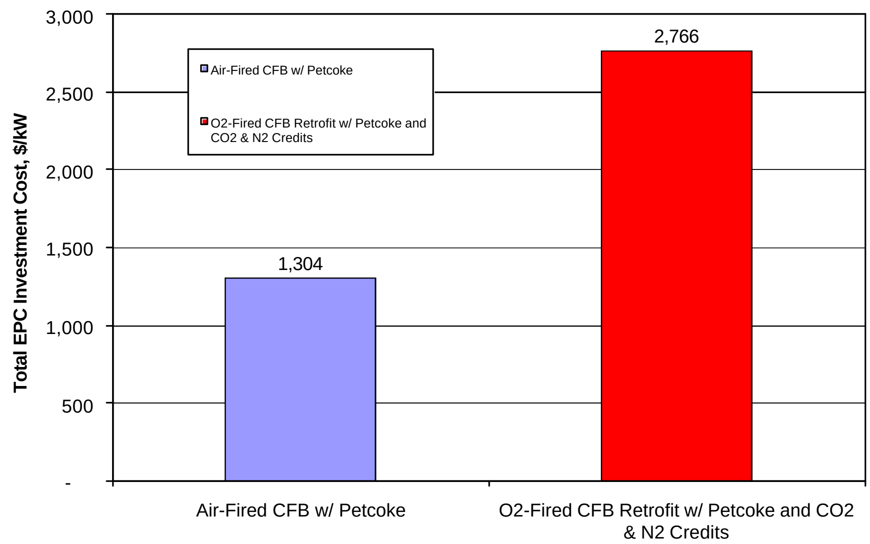


Figure 6.3. 2: Total EPC Investment Costs of Air-Fired CFB and Retrofit of this CFB to O_2 Firing

In conducting the economic analyses of Cases 1 and 2a, the same basic financial assumptions used for all other cases were used here (refer to Table 4.1.1). One difference, however, is the fact that delayed petroleum coke is the fuel of choice in this analysis. The cost of this delayed petroleum coke was assumed to be $\$0.65/10^6$-Btu (Schuller, 2003), as opposed to the base value of $\$1.25/10^6$-Btu used for the coal.

The investment costs of the air-fired CFB and retrofit plant are 1,304 and 2,776 $/kW, respectively (Figure 6.3.2). Levelized costs of electricity for these two cases are presented in Figure 6.3.3. For the retrofit case, credits of 2.4 and 1.7 cents/kWh were calculated for CO_2 and N_2 by-products, respectively. These numbers are equivalent to specific product costs of 17 $/Ton of CO_2 and 4 $/Ton of N_2. It should be noted that CO_2 is the product at the tail end of the gas processing system, and N_2 is a by-product of O_2 production in the air separation unit.

In the analysis of Case-2a, the N_2 credit (4 $/Ton of N_2) was a calculated value. The N_2 value calculated was the cost required to make the calculated COE for Case 2a equivalent to the Case-1 (air-fired electricity production only) COE (i.e., 3.95 cents/kWh). In other words, the breakeven N_2 value was calculated. As stated above, the required N_2 credit was about 4 $/Ton of N_2. By way of comparison, the current value of N_2 is about 11 $/Ton (Schuller, 2003).

The fact that delayed petroleum coke was used as a fuel as opposed to coal in this analysis has very little impact on the overall result. In fact if coal were used instead of delayed petroleum coke, the calculated breakeven N_2 value would have been about 5 $/Ton of N_2 instead of 4 $/Ton.

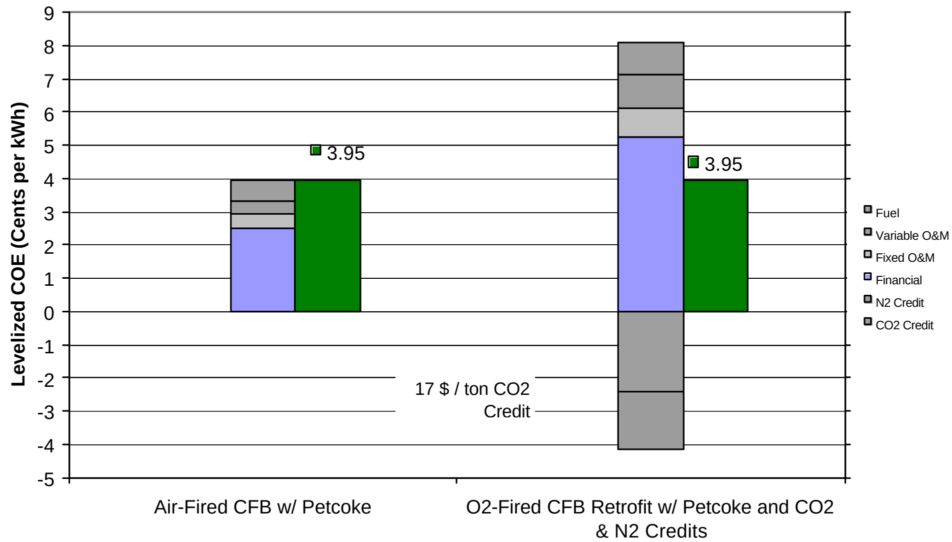


Figure 6.3. 3: Levelized Costs of Electricity for Base Case and EOR Application with Credits

6.4. Vision of Commercial Development Pathway

Figure 6.4.1 depicts a vision of a development pathway for a commercial oxy-fuel fired CFB with CO_2 capture, covering a time period of about five years. A logical first step is the implementation of Budget Period II work scope of the current project, between mid 2003 and end 2004. At the end of this project, there would be sufficient maturity to prepare proposals to demonstrate this technology at a commercial-scale (50-100 MWe range), envisioning a timeline between 2006 and 2008. If such demonstration should prove successful, then it would be feasible to start commercial plant offerings anytime thereafter (i.e., after year 2008).

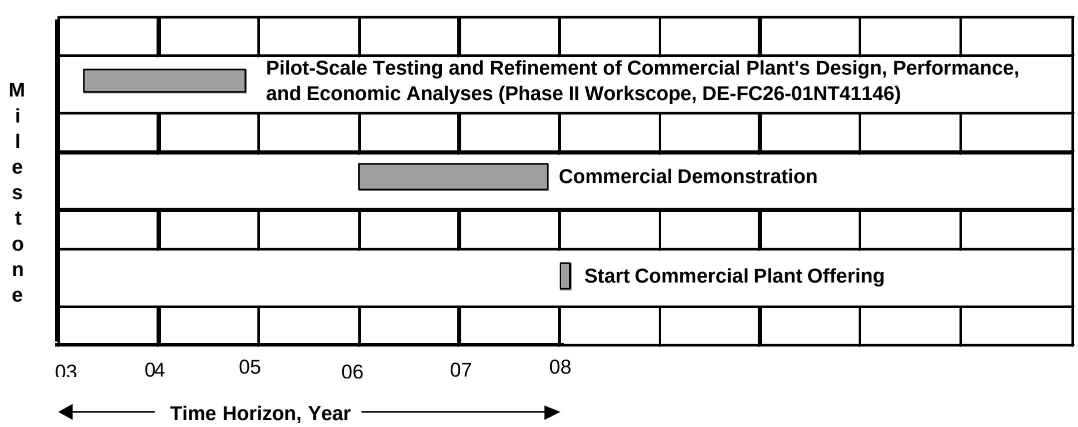


Figure 6.4. 1: Commercialization Pathway

6.5. Concluding Remarks

Results from this evaluation indicate that oxy-fuel fired CFB's can be competitive with the "business as usual" air-fired CFB for electricity production scenario, in a carbon-constrained world, under appropriate niche conditions, e.g., enhanced oil recovery (EOR), with the sale of process by-products (CO_2 and N_2). Importantly, this EOR option is short-ranged, with an implementation timetable of approximately five years.

7. SUMMARY, CONCLUSIONS & RECOMMENDATIONS FOR FUTURE WORK

7.1. Summary

ALSTOM Power Inc.'s Power Plant Laboratories (ALSTOM) teamed with Parsons Energy and Chemical Group Inc., ABB Lummus Global Inc., Praxair Inc., and the US Department of Energy's National Energy Technology Laboratory (DOE NETL), to conduct a comprehensive study evaluating the technical feasibility and economics of alternate CO_2 capture technologies applied to Greenfield US coal-fired electric generation power plants.

The key goals of the study were to evaluate the impacts resulting from the addition of CO_2 capture systems to a variety of newly constructed coal power plants. Major impacts considered were on plant output, efficiency, CO_2 emissions, investment costs, cost of electricity, and CO_2 mitigation costs. An objective of the proposed project was to determine if carbon dioxide could be recovered at an avoided cost of $10/ton (or less).

Thirteen (13) separate but related cases, representing various levels of technology development, were evaluated in a directly comparable manner in this study. The first seven cases represent coal combustion cases in Circulating Fluidized Bed (CFB) equipment. The next four cases represent Integrated Gasification Combined Cycle (IGCC) based systems (Texaco processes). The final two cases are based on an advanced Chemical Looping Gasification process being developed by ALSTOM. These final two cases were developed outside the project funding and are included in these results to show the comparison to advanced cycles that might be "purpose built" for CO_2 capture. These thirteen cases are briefly summarized as follows:

<u>Seven CFB Combustion-Based Cases</u>

- One CFB case, representing a built and operating, air-fired CFB plant without CO_2 capture to provide a reference or "Base Case" for comparison with the other six cases which include CO_2 capture.
- Six CFB-based cases, representing various oxy-fuel fired and other novel concepts, all designed to capture at least 90 percent of the CO_2 in the flue gas.
- These seven combustion cases used 1,800 psia / 1,000 ^oF / 1,000 ^oF steam cycles.

<u>Four IGCC Cases</u>

- Two reference or "Base Cases" representing a built and operating IGCC and a commercially offered, but not yet built, IGCC, both without CO_2 capture to provide comparison data for the other two IGCC cases with CO_2 capture.
- Two cases, one representing a built and operating IGCC and the other representing a commercially offered IGCC, both integrated with a water-gas shift reactor and CO_2 capture equipment to capture 90 percent of the CO_2 in the flue gas.
- These four IGCC cases all used Texaco gasifiers with F-Class gas turbines and 1,800 psia / 1,000 ^oF / 1,000 ^oF steam cycles.

<u>Two Advanced Chemical Looping Cases</u>

- One "Base Case" without CO_2 capture to provide a reference point for the other case.
- One case designed to capture more than 90 percent of the CO_2 in the flue gas.
- These two advanced Chemical Looping cases used F-Class gas turbines with 1,800 psia / 1,000 ^oF / 1,000 ^oF steam cycles.

The steam cycle represents a common thread among all the cases. It is nearly identical for all the combustion cases differing only in the arrangement of the low level heat recovery system or small process steam extractions in some cases (Case-5 and Case-7). The steam turbine for the combustion cases is an 1,800 psig, 1,000 ^oF / 1,000 ^oF, single reheat machine with a main steam flow of 1,400,555 lbm/hr and a condenser pressure of 3.0 inches of mercury, absolute (Hga). The cold reheat flow is 1,305,632 lbm/hr. The main steam flow is identical for all the combustion cases. The reheat steam flow is also identical for all the combustion cases except for a slight increase in Case-6. Six extraction feedwater heaters are used, and the final feedwater temperature is 470 ^oF.

The steam cycles utilized for the four IGCC cases and two advanced Chemical Looping cases used the same steam conditions as for the combustion cases but with somewhat different steam flows as required by the respective gasifier, gas turbine and heat recovery arrangements. These steam cycles are well within today's technology for steam, and no attempt was made to optimize the cycle efficiencies using supercritical or advanced supercritical steam cycles.

The scope of work for each power plant/concept, with or without CO_2 capture, consisted of the following major analyses: process and equipment design, plant performance, plant cost estimation, and levelized economics. The Dakota Gasification Company's CO_2 specification (DGC WebPages, 2001) for Enhanced Oil Recovery (EOR) was used as the basis for the design of the CO_2 capture systems. One of the CO_2 capture systems (Case-3) was designed for essentially "zero CO_2 emissions into the atmosphere." That is, the entire dry flue gas stream (including SO_2, excess O_2, NOx, etc.) was compressed and liquefied in preparation for "sequestration only." As such, this so-called "dirty" flue gas did not have to meet the stringent pipeline quality specifications stipulated by the Dakota Gasification Company. However, the actual costs of sequestering the CO_2 were beyond the scope of this work.

Table 7.1.1 summarizes the results obtained from the thirteen cases studied herein.

Table 7.1. 1: Summary of Plant Performance and Economic Analyses

Study Case	Net Plant Efficiency, % HHV	Net Plant Output, kW	Total Investment Cost, EPC Basis		Total O&M, cents/kWh	Total COE (c/kWh)
			$x1000	$/kW		
Without CO2 Capture						
Case 1, Air-fired CFB w/o CO2 Capture	35.5	193,037	251,804	1,304	0.83	4.53
Case 8, Built & Operating IGCC w/o CO2 Capture	37.6	263,087	411,731	1,565	0.97	5.30
Case 10, Commercially Offered IGCC w/o CO2 Capture	34.5	235,294	341,468	1,451	0.99	5.22
Case 12, Chemical Lopping Gasification w/o CO2 Capture	41.4	265,146	296,991	1,120	0.67	4.28
With CO2 Capture						
Case 2, O2-Fired CFB w/ASU & CO2 Capture	25.2	134,514	328,589	2,443	1.77	8.25
Case 3, O2-Fired CFB w/ASU & Flue Gas Sequestration	25.3	135,351	320,638	2,369	1.76	7.98
Case 4, O2-Fired CMB w/ASU & CO2 Capture	24.6	132,168	337,402	2,553	1.81	8.41
Case 5, Air-Fired CFB w/Carbonate Reg. Process & CO2 Capture	30.2	161,184	270,232	1,677	1.25	5.95
Case 6, O2-Fired CMB w/OTM & CO2 Capture	30.0	197,435	468,919	2,375	1.20	7.05
Case 7, CMB Chemical Looping Combustion w/CO2 Capture	30.9	164,484	273,568	1,663	1.20	5.84
Case 9, Built & Operating IGCC w/ CO2 Capture	29.8	230,515	502,330	2,179	1.32	7.15
Case 11, Commercially Offered IGCC w/CO2 Capture	27.4	201,004	412,377	2,052	1.43	7.18
Case 13, Chemical Lopping Gasification w/ CO2 Capture	36.9	256,830	355,132	1,383	0.99	5.22

7.2. Conclusions

- There are a number of potentially viable approaches to CO_2 capture for sequestration in solid fuel-fired power plants.

- In the long term, the potential for advanced Chemical Looping gasification appears to come closest to approaching the target economic values:
  - COE = 5.2 cents/kWh vs. today's COE = 4.5 cents/kWh
  - Avoided cost of CO_2 = $11/ton of CO_2 vs. $3/ton of CO_2 or $10/ton of carbon

- Nearer term, chemical looping combustion alternatives appear to be economically superior to IGCC for CO_2 capture.

- Oxygen-fired CFB/CMB alternatives can be competitive with "business as usual" electricity production under appropriate niche conditions (EOR) and are available in the near term.

- Oxygen-fired CFB/CMB with flue gas sequestration provides the only true "Zero Gaseous Emissions" power plant studied herein.

- Oxygen-fired CFB/CMB can provide near term demonstration of circulating moving bed technology, which is an enabling technology for chemical looping.

- In a <u>region without carbon constraints</u>, where only power generation is considered, the following techno-economic conclusions can be drawn based on the results from this study:

  - With respect to efficiency:
    - The Texaco IGCC power plant technology is more efficient than the air-fired CFB technology with subcritical steam conditions (37.6 vs. 35.5 percent, HHV). This is due principally to the fact that the IGCC technology takes advantage of both elevated pressure and combined cycle operation principles.
    - The advanced Chemical Looping gasification concept, being developed by ALSTOM, is more efficient than the Texaco IGCC (41.4 vs. 37.6 percent, HHV). This is due to the fact that advanced Chemical Looping takes advantage of: (1) Combined cycle operation principles; and (2) Separation of oxygen from air via the chemical looping process rather than oxygen from a cryogenic air separation unit.
    - The Texaco IGCC efficiency values reported in this study are lower than values reported by Parsons (Holt, 2000) primarily because Parsons' study used H-class gas turbines. Additionally, Parsons used ambient conditions of (63 ^oF and 14.4 psia), whereas ALSTOM used ABMA ambient conditions (80 ^oF and 14.7 psia). Also, Parsons and ALSTOM used condenser pressures of 2.0, and 3.0 in. Hga, respectively.

  - With respect to investment costs and levelized costs of electricity (COE):
    - The Texaco IGCC power plant technology is about 20 percent more capital-intensive than the air-fired CFB technology (> 250 $/kW). This is due to the fact that an IGCC power plant syngas cleanup system is very complex, comprised of many specialized components operating at elevated pressures.
    - The air-fired CFB plant produces electricity about 15 percent cheaper in terms of COE than the Texaco IGCC plant (4.5 vs. 5.3 cents/kWh)

- ➢ The advanced Chemical Looping gasification plant concept shows the potential to provide electricity at a lower cost than both air-fired CFB and IGCC plants (4.3 vs. 4.5 and 5.3 cents/kWh). It is noted that the advanced Chemical Looping plant concept is in a very early stage of development. Hence, the COE estimates for this concept are preliminary.

- ❑ In <u>a carbon-constrained region</u>, where both power generation and carbon capture are considered, the following techno-economic conclusions can be drawn based on the results from this study:

  - ♦ With respect to efficiency:
    - ➢ All options reduce power plant efficiencies compared to baseline plants without CO_2 capture.
    - ➢ The advanced Chemical Looping gasification plant concept is the most efficient of all cases studied herein (36.9 percent vs. 24.6 - 30.9 percent, HHV). Additionally, this plant produces Hydrogen as its fuel gas.
    - ➢ The efficiencies of the advanced oxygen-fired combustion cases (CMB w/OTM, CFB w/Carbonate Regeneration Process, and Chemical Looping Combustion (CLC) CMB) fall in the range that is marginally higher than that of the Texaco IGCC plant (30 – 31 percent vs. 29.8 percent, HHV).
    - ➢ The Texaco IGCC power plant technology is more efficient than cryogenic oxygen-fired CFB power plant technology (29.8 percent vs. 24.6-25.3 percent, HHV). These results are equivalent to energy penalties, compared to their respective reference plants, of 21 percent for IGCC and 28 - 31 percent for CFB plants. This is principally due to the fact that the CO_2 product is compressed from 50 psia to 2,000 psig for the IGCC and from atmospheric pressure to 2,000 psig for the CFB's. Also, the CFB plants require more oxygen per unit of coal fired than the Texaco IGCC plants, resulting in additional ASU power requirements.
    - ➢ The use of oxy-fuel firing to produce a "for sequestration only" flue gas yields only marginal benefit from a plant performance efficiency standpoint (25.3 vs. 25.2 percent, HHV). This is due to the fact that the compression step, which is the most energy-intensive in flue gas processing, cannot be avoided. However, this plant provides the only true "zero emissions" plant.

  - ♦ With respect to investment costs and levelized costs of electricity (COE):
    - ➢ The advanced Chemical Looping gasification plant concept is the least costly of all the concepts considered, its EPC (engineered, procured, and constructed) capital cost and levelized cost of electricity are 1,380 $/kW and 5.2 cents/kWh, respectively. This cost of electricity for advanced Chemical Looping is nearly equivalent to new coal-fired plants offered today without CO_2 capture (Case 1 @ 4.5 cents/kWh).
    - ➢ The Carbonate Regeneration Process and CLC CMB are less capital-intensive than the built and operating or commercially offered IGCC's (1,680 - 1,660 vs. 2,180 - 2,050 $/kW). Hence, their levelized costs of electricity are correspondingly lower (5.9 - 5.8 vs. 7.2 cents/kWh).
    - ➢ The cryogenic oxygen-fired CFB or CMB plants are more capital intensive than the built and operating or commercially offered IGCC's, because they incur high energy penalties by comparison (2,350 - 2,550 vs. 2,180 -2,050 $/kW). Consequently, their levelized costs of electricity are correspondingly higher (8.3 - 8.4 vs. 7.2 cents/kWh)
    - ➢ The capital investment of an oxy-fuel fired plant designed to produce "for sequestration only flue gas" is approximately 3 percent lower than that of an oxy-fuel fired CFB designed to produce a CO_2 product suitable for EOR application (2,370 vs. 2,440 $/kW). Consequently, its levelized cost of electricity is correspondingly about 3 percent lower (8.0 vs. 8.3 cents/kWh)

- Figure 7.2.1 is a plot of COE vs. Capacity Factor for all the technologies evaluated. This plot was obtained by keeping all the economic assumptions given in Table 4.1.1 at their "base" values, and varying only the capacity factor. Overall, these results indicate the following:

  - One of the lessons learned from this figure is that with CO_2 capture, the cycle advantages of IGCC over oxygen fired combustion systems may overcome IGCC's disadvantages in terms of capital cost and availability found in a no CO_2 capture comparison and may well provide marketplace incentive to move this technology into the mainstream market.

  - While IGCC with CO_2 capture offers potential advantages over oxygen based (utilizing cryogenic ASU's) combustion systems with CO_2 capture, the advantage of chemical looping over IGCC is even greater. Taken as a group, a possible road map emerges which shows the oxygen based combustion system as a short-term solution and a strong economic incentive for the development of Chemical Looping. Each step has significant advantages over the prior step.

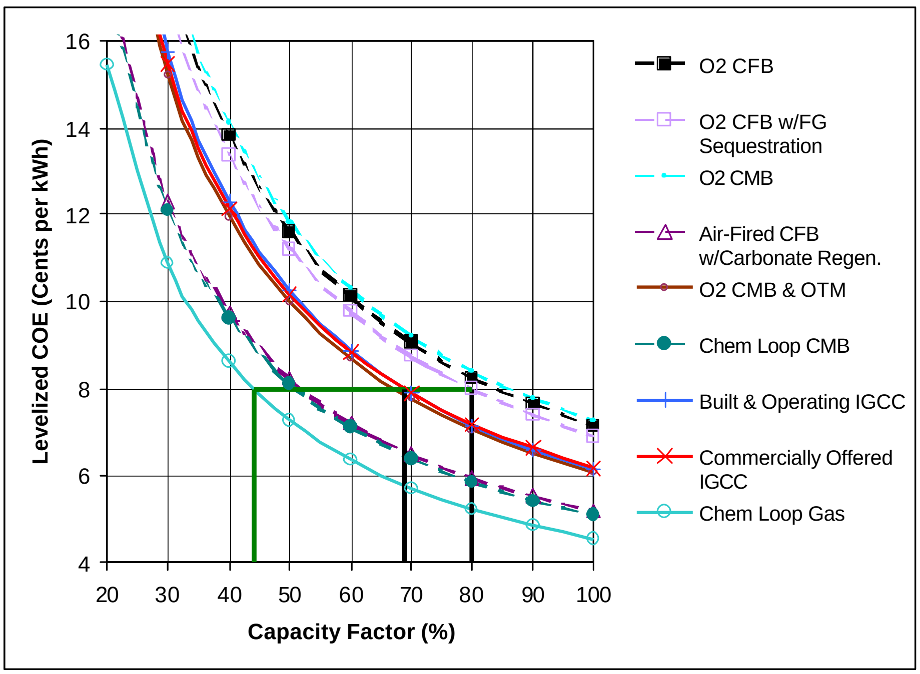


Figure 7.2. 1: Variation of Levelized Cost of Electricity with Capacity Factor for Various Plant Technologies with CO_2 Capture

  - The chart above demonstrates the importance of availability for an operating plant. In the example shown, an O_2-fired CFB with flue gas sequestration and 80 percent availability has the same COE as an IGCC plant with CO_2 capture and 70 percent availability. Industry practice for CFB's has been to have one planned two-week outage every twelve to eighteen months. Current requirements for IGCC plants suggest two planned three-week outages per year. The differences in planned outage times equalizes the COE of these two cases.

- ♦ Another observation is that a Chemical Looping Gasification Plant would have the same COE at 43 percent availability. Such a figure provides considerable latitude in the initial stages of product introduction, where most start-up problems occur. This reduces the commercialization risk of the Chemical Looping Gasification technology.

7.3. Recommendations for Future Work

It is recommended that the Department of Energy's National Energy Technology Laboratory pursue a strategy that supports technologies with short-range and long-range commercialization potential that include the following.

- Continue the development of the circulating moving bed (CMB), O_2-fired CFB, Chemical Looping Combustion and Chemical Looping Gasification.

- Short-Range Technology: Oxy-fuel Fired CFB technology. Given that it would use a combination of already available technologies (e.g., cryogenic O_2 production and gas processing system), it could be commercially deployed within a five-year time horizon. The moving bed portion of the CMB technology is also utilized in the O_2-fired CFB. As shown in Section 6, this technology would be suitable for enhanced oil recovery (EOR) application, with the Bakersfield Project in California being a potential first application site. This technology also would be applicable for enhanced gas recovery (EGR) via coal bed methane. An additional advantage of this technology is that efficiency is maintained over a wide plant size range (50 MWe and larger).

  The DOE's authorization of the present project's continuation -- Phase II, pilot-scale testing of two coals and one delayed petroleum coke, followed by a refinement of the oxy-fuel plant's design, performance and economic analyses -- represents a first step toward this development. This technology is the only one capable of "zero gaseous emissions" at the present time.

- Long-Range Technology: Advanced Chemical Looping technology is a technology that shows such promise that ALSTOM has already begun the design of a small "Proof of Concept" pilot-scale facility. Additionally, ALSTOM has responded to a DOE NETL RFP to conduct an extensive test program in this facility (DE-PS26-02NT41613-01). The Circulating Moving Bed (CMB^{TM}) technology, being developed by ALSTOM with DOE NETL's financial support (Jukkola, et al., 2003), is short-range, enabling technology that represents a stepping-stone towards the development of the Chemical Looping technology.

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9. APPENDICES

This section provides four appendices that provide detailed information for each of the thirteen plants studied. The basic information included in the each appendix is listed below.

- **Appendix I** - Equipment Lists
- **Appendix II -** Drawings
- **Appendix III -** Detailed Investment Cost and Operating & Maintenance Cost Breakdowns
- **Appendix IV** - Economic Sensitivity Study Results

9.1. Appendix I - Equipment lists

Appendix I provides complete equipment lists for each of the thirteen plants studied. The lists are presented consecutively, starting with Case-1 and ending with Case-13. The lists are grouped into four separate areas: Boiler or Gasifier Island, Air Separation, Gas Processing, and Balance of Plant, which includes the power block. Some of the cases do not have equipment in all four areas.

9.1.1. Case-1 Equipment List

9.1.1.1. Boiler Island Equipment

Fuel Feeding System:
- Day Silo
- Fuel Silo Isolation Valves
- Fuel Feeders
- Feeder Isolation Valves
- Piping to Furnace
Limestone Feeding System:
- Day Silo
- Limestone Silo Isolation Valves
- Rotary Feeder
- Blower
- Piping from Blower to Furnace Injection Points
Furnace Loop Equipment:
- Drum Including Internals, Nozzles, Lugging, Hanger Rods
- Downcomer System
- Connecting Tubes/Piping
- Furnace Tube Panels/Headers
- Furnace Evaporator Pendants/Headers/Piping
- Furnace Grate and Plenum Including Air Nozzles
- Ash Drain Valve(s)
- Start-up Burner System (Including Burners, Piping, Ducts, and Local Control Equipment)
- Ductwork – Furnace to Recycle Particle Separators
- Refractory-Lined Recycle Particle Separator – Complete
- Ductwork – Recycle Particle Separator to Backpass Inlet
- Backpass Enclosure
- Metal/Fabric Expansion Joints
- Seal pots and Seal pot Grate – Including Air Nozzles and Plenum
- Buckstay System:
Furnace
Backpass
Backpass & FBHE Equipment:
- Connecting Tubes/Piping
- Backpass Tube Panels/Headers
- Backpass Heat Absorbing Surface:
Horizontal Economizer
Horizontal Superheater/Reheater
- Superheater/Reheater Desuperheaters
- Desuperheater Block Valves
- Desuperheater Piping
- Economizer Piping to Drum
- Superheater Interconnecting Piping
- Feedwater Stop, Feedwater Check
- Safety Valves/Discharge Piping/Silencers
- Electro. Relief Valve/Silencer and Discharge Piping
Trim Valves:

- Double Valving
Drum Level Gauge and Indicators
Sootblowing System:
- Economizer
- Superheater/Reheater
- Airheater
- Sootblower Control System
Air System:
- Primary Air Fan w/Drive (by others)
- Secondary Air Fan w/Drive (by others)
- Fluidizing Air Blower w/Drive (by others)
- Fan and Blower Inlet Silencers (by others)
- Tubular Air Heater
- Ductwork - Fan Outlet(s) to Airheater Inlet(s)
- Ductwork - Blower Outlets to Seal pots
- Steam Coil Air Preheater
- Air Duct Expansion Joints
Combustion Gas System:
- Ductwork and Expansion Joints - Economizer Outlet to Airheater
- Ductwork - Airheater Outlet (including airheater plenum & hoppers)
- Ductwork Outlet to I.D. Fan Inlet
- I.D. Fan w/Drive
- Ductwork – I.D. Fan Outlet to Stack Flange Connection
Ash Handling System:
- Bed Ash Drains and Ash Coolers
Structural:
- Structural Steel including platforms, walkways, stairways, and ladders
- Boiler Internal Grid Steel
- Boiler Island Elevator
- Pressure Part Support Steel
- Boiler Building Siding, Weather Enclosure, HVAC
Instrumentation and Controls:
- Burner Management (FBSS) Logic
- CFB Field Instruments
- Controller Drives
Refractories:
- Material for All Internal Refractory Linings for Furnished Process and Boiler Equipment
Insulation and Lagging:
- Material for Insulation and Lagging for Heat Conservation and Personnel Protection for furnished equipment
Painting:
- Shop Prime Paint Coating for Seller furnished Equipment
Miscellaneous:
- Operator Training Program
- Maintenance Training Program

- Instruction Manuals
- Spare Parts for commissioning
- Technical Representation during start-up and testing
- Field Erection of Equipment Scope
- Freight to Site

9.1.1.2. Case-1 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-1 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor 1	54" belt	450 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	2
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	200 tph	2
3	Conveyor 4	48" belt	200 tph	1
4	Transfer Tower	N/A	200 tph	1
5	Tripper	N/A	200 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	500 ton	2
7	Feeder	Gravimetric	100 tph	2

ACCOUNT 2B LIMESTONE PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Bin Activator		20 tph	1
2	Weigh Feeder	Gravimetric	20 tph	1
3	Storage Silo	Cylindrical	1,000 ton	1
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Per Alstom Steam Cycle

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 1,000 gpm	2
14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	15,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Design Condition	Qty
1	Power Boiler	AFBC	Alstom Design	
2	Primary Air Fan	Centrifugal	1,165,000 pph, 281,000 acfm, 98" wg, 5,000 hp	1
3	Secondary Air Fan	Centrifugal	677,000 pph, 163,000 acfm, 78" wg, 2,500 hp	1
4	ID Fan	Centrifugal	2,168,000 pph, 724,000 acfm, 39" wg 4,200 hp	1
5	Fluidizing Air Blower	Centrifugal	16,000 acfm/25 psig 1,800 hp	2

ACCOUNT 5 FLUE GAS CLEANUP

ACCOUNT 5A PARTICULATE CONTROL

Equipment No.	Description	Type	Design Condition	Qty
1	Bag Filter	Pulse-jet cleaned	550,000 acfm, total removal efficiency = 99.9%+	1

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Not applicable.

ACCOUNT 7 DUCTING AND STACK

Equipment No.	Description	Type	Design Condition	Qty
1	Boiler Stack	Concrete with FRP liner	291°F 600,000 acfm	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	209 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	60,000 gpm @ 80 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

In boiler scope of supply.

ACCOUNT 10B FLY ASH HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Baghouse Hoppers (part of baghouse scope of supply)			12
2	Air Heater Hopper (part of boiler scope of supply)			1
3	Air Blower		1,800 cfm	1
4	Fly Ash Silo	Reinforced concrete	400 tons	1
5	Slide Gate Valves			2
6	Wet Unloader		30 tph	1
7	Telescoping Unloading Chute			1

9.1.2. Case-2 Equipment List

9.1.2.1. Case-2 Boiler Island Equipment

Fuel Feeding System:
- Day Silo
- Fuel Silo Isolation Valves
- Fuel Feeders
- Feeder Isolation Valves
- Piping to Furnace
Limestone Feeding System:
- Day Silo
- Limestone Silo Isolation Valves
- Rotary Feeder
- Blower
- Piping from Blower to Furnace Injection Points
Furnace Loop Equipment:
- Drum Including Internals, Nozzles, Lugging, Hanger Rods
- Downcomer System
- Connecting Tubes/Piping
- Furnace Tube Panels/Headers
- Furnace Evaporator Pendants/Headers/Piping
- Furnace Grate and Plenum Including Air Nozzles
- Ash Drain Valve(s)
- Start-up Burner System (Including Burners, Piping, Ducts, and Local Control Equipment)
- Ductwork – Furnace to Recycle Particle Separators
- Refractory-Lined Recycle Particle Separator – Complete
- Ductwork – Recycle Particle Separator to Backpass Inlet
- Backpass Enclosure
- Metal/Fabric Expansion Joints
- Seal pots and Seal pot Grate – Including Air Nozzles and Plenum
- Buckstay System:
Furnace
Backpass
Backpass & MBHE Equipment:
- Connecting Tubes/Piping
- Backpass Tube Panels/Headers
- Backpass Heat Absorbing Surface:
Horizontal Economizer
- MBHE Heat Absorbing Surface:
Horizontal Superheaters
Horizontal Reheater
Horizontal Evaporator
- Superheater/Reheater Desuperheaters
- Desuperheater Block Valves
- Desuperheater Piping
- Economizer Piping to Drum
- Superheater Interconnecting Piping
- Feedwater Stop, Feedwater Check valves

- Safety Valves/Discharge Piping/Silencers
- Electro. Relief Valve/Silencer and Discharge Piping
Trim Valves:
- Double Valving
Drum Level Gauge and Indicators
Sootblowing System:
- Economizer
- Superheater/Reheater
- Oxygen Heater
- Sootblower Control System
Oxygen System:
- Gas Recirculation Fan w/Drive (by others)
- Fluidizing Gas Blower w/Drive (by others)
- Fan and Blower Inlet Silencers (by others)
- Tubular Oxygen Heater
- Ductwork – GR Fan Outlet(s) to Oxygen Heater Inlet(s)
- Ductwork – FG Blower Outlets to Seal pots
- Steam Coil Oxygen Preheater
- Oxygen Duct Expansion Joints
Combustion Gas System:
- Ductwork and Expansion Joints - Separator Outlet to Oxygen Heater
- Ductwork – Oxygen Heater Outlet (including O_2 heater plenum & hoppers)
- Ductwork O_2 heater Outlet to Baghouse Inlet
- Ductwork Baghouse Outlet to PFWH Inlet
- PFWH economizer
- Ductwork PFWH Outlet to Gas Cooler Inlet
- Gas Cooler (by others)
- Ductwork Gas Cooler Outlet to ID Fan Inlet
- I.D. Fan w/Drive (by others)
- Ductwork ID Fan Outlet to GR Fan and FG Fan Inlet
Ash Handling System:
- Bed Ash Drains and Ash Coolers
Structural:
- Structural Steel including platforms, walkways, stairways, and ladders
- Boiler Internal Grid Steel
- Boiler Island Elevator
- Pressure Part Support Steel
- Boiler Building Siding, Weather Enclosure, HVAC
Instrumentation and Controls:
- Burner Management (FBSS) Logic
- CFB Field Instruments
- Controller Drives
Refractories:
- Material for All Internal Refractory Linings for Furnished Process and Boiler Equipment
Insulation and Lagging:

- Material for Insulation and Lagging for Heat Conservation and Personnel Protection for furnished equipment
Painting:
- Shop Prime Paint Coating for Seller furnished Equipment
Miscellaneous:
- Operator Training Program
- Maintenance Training Program
- Instruction Manuals
- Spare Parts for commissioning
- Technical Representation during start-up and testing
- Field Erection of Equipment Scope
- Freight to Site

9.1.2.2. Case-2 Gas Processing System Equipment

Tag No.	Service	Sizing Parameters	MOC
DA	Columns and Towers		
DA-101	Direct Contact Flue Gas Cooler	18' 6" ID x 34' S/S, DP 10 psig, 3 psi vacuum	CS w/ SS liner
DA-102	CO2 Rectifier	4'/ 12' ID x 30'/ 12' S/S, DP 425 psig	LTCS
E	Heat Transfer Equipment		
EA	Shell & Tube Exchangers		
EA-101	Flue Gas Compressor 1 Stage Trim Cooler	7.8 MMBTU/h, DP S/T, 85 psig/ 125 psig	CS/SS
EA-102	Flue Gas Compressor 2 Stage Trim Cooler	4.1 MMBTU/HR, DP S/T, 175 psig/ 125 psig	CS/SS
EA-103	Flue Gas Compressor 3 Stage Trim Cooler	4.2 MMBTU/HR, DP S/T, 425 psgi/ 125 psig	CS/SS
EA-104	CO2 Condenser	61.4 MMBTU/HR, DP S/T, 300 psig/ 425 psig	LTCS/ LTCS
EA-201	Refrig condenser	69.5 MMBTU/HR, DP S/T, 300 psig/ 125 psig	CS/CS
EA-202	Refrig Subcooler	20 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/ LTCS
EA-107	CO2 Rectifier Condenser	5.0 MMBTU/HR, DP S/T, 425 psig/ 425 psig	SS/SS
EA -108	Rectifier Ovhd Interchanger	1.6 MMBTU/HR, DP S/T, 425 psig/ 425 psig	LTCS/ SS
EB	Plate Exchangers		
EB-101	Water Cooler	Total 55 MMBTU/HR, DP P/U, 65 psig/ 125 psig	CS
EC	Air Coolers		
EC-101	Flue Gas Compressor 1 Stage Aftercooler	24.5 MMBTU/HR, DP 85 psig	SS
EC-102	Flue Gas Compressor 2 Stage Aftercooler	23.7 MMBTU/HR, DP 175 psig	SS
EC-103	Flue Gas Compressor 3 Stage Aftercooler	21.0 MMBTU/HR, DP 425 psig	SS
FH	Heaters		
FH-101	Dryer Regeneration Gas Heater	Gas fired, 5.75 MMBTU/HR fired duty	
FA	Drums and Vessels		
FA-101	Flue Gas Compressor 2nd Stage Suction Drum	13' - 6" ID x 18' S/S, DP 85 psig	CS w/ SS liner
FA-102	Flue Gas Compressor 3rd Stage Suction Drum	10' - 0" ID x 14' S/S DP 175 psig	CS w/ SS liner
FA-103	Flue Gas Compressor Third Stage Discharge K/O Drum	8' ID x 12' S/S, DP 425 psig	CS w/ SS liner
FA-201	Refrig Surge Drum	12' ID x 30' S/S DP 300 psig	CS
FA-202	Refrig Suction Srubber	16' ID x 24' S/S, DP 300 psig	ITCS
FD	Filters and Dryers		
FD-101	Water Filter	5 units, 805 gpm each, DP 100 psig	SS
FD-102	Flue Gas Filter	Total 2665 ACFM, DP 425 psig	CS
FF	Dryers (Dessicant Type)		
FF-101	Flue Gas Dryer	Two Vessels 10' -0 " ID x 20' S/S DP 425 psig DT 500 F	CS
GA	Pumps Centrifugal		
GA-101	Water Pump	4180 gpm, DP 40 psi	CI w/ SS impeller
GA-103	CO2 Pipeline pump	745 gpm, DP 1655 psi	ITCS
GB	Compressors & Blowers		
GB-101	Flue Gas Compressor	Motor Drive 3 stages Includes Lube/Seal Oil Systems, 19760 kW	SS
GB-102	Propane Refrig Compressor	Motor Drive, Includes lube oil/ seal oil system, 8700 kW	ITCS

9.1.2.3. Case-2 Air Separation Unit Equipment

Rotating Equipment

Main Air Compressor (Qty 1)

One centrifugal compressor meets the entire range of plant air. The compressor is a 3-stage high efficiency integral gear centrifugal compressor. Included with the compressor are adjustable inlet guide vanes, coupling with guard, lube oil system and two intercoolers. Aftercooling of the MAC will be accomplished in a Direct Contact Aftercooler. The compressor is driven by a synchronous electric motor which is field mounted on its' own foundation.

Delivered Air Flow:	511,000 Nm^3/h (19,40,000 cfh-ntp)
Suction Temperature:	27°C (80°F)
Discharge Pressure:	6 bar(a) (87 psia)

Upper Column Turbine Skid (UCT) (Qty 1)

A Cryogenic expansion turbine provides refrigeration for producing liquid products and heat leak for the distillation process. The Turbine is sized for plant specific requirements. Lube oil is provided by an integral lube oil skid.

Delivered Flow:	22,560 Nm^3/h (857,300 cfh-ntp)
Isothermal Efficiency:	90%
Inlet Temperature:	-88°C (-127°F)
Exhaust Pressure:	1.4 bar(a) (21 psia)

Process Equipment

Air Suction Filter House (ASFH) (Qty 1)

A pulse jet type filter house will be implemented for this case. The filter will be built in 3 modules.

Overall Efficiency:	100% retention of 3 micron particles
Design Flow	511,000 Nm^3/h (19,400,000 cfh-ntp)

Direct Contact Aftercooler (DCA) (Qty 1)

The heat of compression from the MAC is removed through a two-stage Direct Contact Aftercooler (DCA). The DCA is a packed column where water is put in direct contact with compressed air from the MAC. The 1st stage of the DCA is cooled by water from the plant cooling water system. The air exiting this first stage is cooled to within 1°C of the cooling water inlet temperature. The 2nd stage of the DCA is fed by a closed chilled water loop. A Mechanical Chiller provides the refrigeration to chill this stage's water loop. The air exiting the 2nd stage is designed to be at 15°C or less to feed the Prepurifier system. An integral Moisture Separator is provided to remove 99.9 % of free water droplet 3 microns and larger.

Design Discharge Air Temp.:	10.0°C (50°F) Process Air to TSA PP
1st Stage Packing Height:	2.4 m (9.5 ft)
1st Stage Water Flow:	18,900 l/min (5,000 gpm)
2nd Stage Packing Height:	3.2 m (9.5 ft)
2nd Stage Water Flow:	8,710 l/min (2,300 gpm)

Mechanical Chiller (Qty 4)

An R-134A mechanical chiller provides refrigerant to cool the 2nd stage DCA chilled water. The mechanical chiller cools down the water to within the desired process temperature. The chiller consists of one full sized, centrifugal refrigerant compressor, and shell and tube heat exchangers for the evaporator and condenser services.

Tons @ 100% Load	450 (1,800 Total)
Water Design Temperature:	8.9ºC (48ºF)
Evaporator Water Flow:	8,710 l/min (2,300 gpm)

DCA Chilled Water Pumps (Qty 2)

	Chilled Water Pump	1st Stage DCA Pump
Pump Flow Range:	8,710 l/min (2,300 gpm)	18,930 l/min (5,000 gpm)
Design TDH:	20 m (65 ft)	39 m (127 ft)

TSA Prepurifier Vessels (Qty 2)

The air purification system is designed to remove water and CO_2 from the feed air stream going to the column or other warm end piping in order to prevent fouling heat exchangers from CO_2 buildup in the main condenser. The system is designed as a horizontal two-bed system with each vessel containing a bed of molecular sieve. While one vessel is removing water and CO_2 from the feed air stream, the other bed is being regenerated at low pressure by hot N_2 from a Regeneration Heater. Water, CO_2, and other hydrocarbons are desorbed from the sieve and vented to atmosphere.

Design Inlet Air Temperature:		10.0°C (50°F) {Process Air from DCA}
Adsorbents:	Sieve:	4x8 13X APG II Molecular Sieve 86,200 kg (190,000 lbs) Each
	Alumina:	D-201 Alumina 29,500 kg (65,000 lbs) Each
Est Vessel Size:		4.9 m Diam. x 13.1 m L (16 ft. Diam. x 43 ft. L) (Seam to Seam)

TSA Prepurifier Dust Filter (Qty 2)

Following adsorption, the air passes through one full-size Dust Filter to remove any particles of molecular sieve. The filter design provides positive gasket sealing to prevent by-pass of unfiltered fluids.

Filter Efficiency:	99% retention of 1 micron particles 100% retention of 3 micron particles

TSA Prepurifier Natural Gas Regeneration Heater

One 100% Natural Gas Regeneration/Thaw heater is used to heat the Regeneration N_2 and Thaw Air. The unit packaged and mounted on a single skid. The burners are fully modulating, with combustion air blower and motor. A packaged control system is included for control and safety monitoring.

Design Regeneration Flow:	75,200 Nm^3/h (2,855,000 cfh-ntp)

Design Heat Duty:	7,115 kW (24,300,000 Btu/hr)
Inlet Temp	29 °C (85 °F)
Outlet Temp	232 °C (450 °F)
Peak Fuel Consumption	966 Nm^3/h (34,100 scfh)

Silencers

All silencers provide a 35 dBA insertion loss. 50 dBA attenuation is also available.

	MAC Vent (Qty 1)	**Waste Nitrogen Vent (Qty 1)**
Inlet:	457mm (24 in) dia	508 mm (20 in.) Diam
Outlet:	2,742 mm (96 in) diam	2,437 mm (96 in.) Diam
Length:	3,046 mm (120 in)	3,046 mm (120 in.)

	Prepurifier Vent (Qty 1)	**Product Oxygen Vent (Qty 1)**
Inlet:	254 mm (10 in.) Diam	686 mm (27 in.) Diam (Reduced)
Outlet:	660 mm (26 in.) Diam	1,524 mm (60 in.) Diam
Length:	1,803 mm (71 in.)	4,242 mm (167 in.)

Cold Box Equipment
Primary Heat Exchanger (PHX) (Qty 1)

Oxygen Boiler

Main Condenser

Lower Column

Upper Column

Additional Equipment and Services

• Local Instruments & Controls	Praxair
• Switchgear & MCC	Praxair
• Process Analyzers	Praxair
• Cooling System	Client
• Project Management & Engineering	Praxair
• Construction Management	Praxair
• Construction	Local Contractors
• Commissioning & Startup	Praxair with Client support
• Land/Site	Client
• Control Room/Administration Offices/Warehouse/Maintenance Shop, etc.	Client
• Start-Up Utilities	Client

9.1.2.4. Case-2 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-2 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor 1	54" belt	450 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	2
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	200 tph	2
3	Conveyor 4	48" belt	200 tph	1
4	Transfer Tower	N/A	200 tph	1
5	Tripper	N/A	200 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	500 ton	2
7	Feeder	Gravimetric	100 tph	2

ACCOUNT 2B LIMESTONE FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Storage Silo	Cylindrical	1,000 Ton	1
2	Weigh Feeder	Gravimetric	20 tph	1
3	Bin Activator		20 tph	
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Per Alstom steam cycle.

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 2,500 gpm	2
14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	100,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Design Condition	Qty
1	Power Boiler	AFBC	Alstom Design	
2	Gas Recirculation Fan	Centrifugal	217,000 lbm/hr 35,753 acfm 97.5 in. w.g 800 hp	1
3	Fluidizing Gas Blower	Centrifugal	44,825 lbm/hr 7,385 acfm 11.7 w.g 500 hp	2
4	ID Fan	Centrifugal	764,579 lbm/hr 123,330 acfm 38.4 in. w.g 1,200 hp	1

ACCOUNT 5 FLUE GAS CLEANUP

ACCOUNT 5A PARTICULATE CONTROL

Equipment No.	Description	Type	Design Condition	Qty
1	Bag Filter	Pulse-jet cleaned	171,000 acfm, total removal efficiency = 99.9%+	1

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Not applicable.

ACCOUNT 7 DUCTING AND STACK

In Gas Processing scope of supply

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	209 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	1,288 MM Btu/hr 100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	60,000 gpm @ 80 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

In boiler scope of supply.

ACCOUNT 10B FLY ASH HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Baghouse Hoppers (part of baghouse scope of supply)			4
2	Air Heater Hopper (part of boiler scope of supply)			1
3	Air Blower		1,800 cfm	1
4	Fly Ash Silo	Reinforced concrete	100 ton	1
5	Slide Gate Valves			2
6	Wet Unloader		30 tph	1
7	Telescoping Unloading Chute			1

9.1.3. Case-3 Equipment List

9.1.3.1. Case-3 Boiler Island Equipment

The Case-3 Boiler Island equipment is identical to Case-2 and is not repeated here. Refer to section 9.1.2.1 for the relevant Boiler Island equipment list.

9.1.3.2. Case-3 Gas Processing System Equipment

Tag No.	Service	Sizing Parameters	MOC
DA	Columns and Towers		
DA-101	Direct Contact Flue Gas Cooler	18' 6" ID x 34' S/S, DP 10 psig, 3 psi vacuum	CS w/ SS liner
E	Heat Transfer Equipment		
EA	Shell & Tube Exchangers		
EA-101	Flue Gas Compressor 1 Stage Trim Cooler	7.8 MMBTU/h, DP S/T, 85 psig/ 125 psig	CS/SS
EA-102	Flue Gas Compressor 2 Stage Trim Cooler	4.1 MMBTU/HR, DP S/T, 175 psig/ 125 psig	CS/SS
EA-103	Flue Gas Compressor 3 Stage Trim Cooler	4.2 MMBTU/HR, DP S/T, 425 psgi/ 125 psig	CS/SS
EA-104	Flue Gas Compressor 4 Stage Trim Cooler	3.6 MMBTU/HR, DP S/T, 125 psig/ 900 psig	CS
EA-105	Flue Gas Compressor 5 Stage Trim Cooler	9.2 MMBTU/HR, DP S/T, 125 psig/ 2200 psig	CS
EB	Plate Exchangers		
EB-101	Water Cooler	Total 55 MMBTU/HR, DP P/U, 65 psig/ 125 psig	CS
EC	Air Coolers		
EC-101	Flue Gas Compressor 1 Stage Aftercooler	24.5 MMBTU/HR, DP 85 psig	SS
EC-102	Flue Gas Compressor 2 Stage Aftercooler	23.7 MMBTU/HR, DP 175 psig	SS
EC-103	Flue Gas Compressor 3 Stage Aftercooler	21.0 MMBTU/HR, DP 425 psig	SS
EC-104	Flue Gas Compressor 4 Stage Aftercooler	17.7 MMBTU/HR, DP 900 psig	CS
EC-105	Flue Gas Compressor 5 Stage Aftercooler	28.8 MMBTU/ HR DP 2200 psig	CS
FA	Drums and Vessels		
FA-101	Flue Gas Compressor 2nd Stage Suction Drum	13' - 6" ID x 18' S/S, DP 85 psig	CS w/ SS liner
FA-102	Flue Gas Compressor 3rd Stage Suction Drum	10' - 0" ID x 14' S/S DP 175 psig	CS w/ SS liner
FA-103	Flue Gas Compressor Third Stage Discharge K/O Drum	8'- 0" ID x 12' S/S, DP 425 psig	CS w/ SS liner
FA-104	Flue Gas Compressor Fifth Stage Suction K/O Drum	6' -0" ID x 12' S/S, DP 900 psig	CS
FD	Filters and Dryers		
FD-101	Water Filter	5 units, 805 gpm each, DP 100 psig	SS
FD-102	Flue Gas Filter	Total 2665 ACFM, DP 425 psig	CS
	Packages		
PA-101	CO2 Dryer Package	4 90" x 20' driers DP 425 psig, 400 hp sundyne compressor, 7.9 MMBTU/h heater, 6.8 MMBTU/h air cooler. 36" x 8' KO drum	CS w/ SS liner
GA	Pumps Centrifugal		
GA-101	Water Pump	4180 gpm, DP 40 psi	CI w/ SS impeller
GB	Compressors & Blowers		
GB-101	Flue Gas Compressor	Motor Drive 5 stages Includes Lube/Seal Oil Systems, 35860 hp	CS casing w/ SS wheels

9.1.3.3. Case-3 Air Separation Unit Equipment

The Case-3 Air Separation Unit equipment is identical to Case-2 and is not repeated here. Refer to section 9.1.2.3 for the relevant ASU equipment list.

9.1.3.4. Case-3 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-3 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

The Case-3 Balance of Plant Equipment is nearly identical to Case-2 except for the heat removed in the Cooling Water System.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor 1	54" belt	450 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	2
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	200 tph	2
3	Conveyor 4	48" belt	200 tph	1
4	Transfer Tower	N/A	200 tph	1
5	Tripper	N/A	200 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	500 ton	2
7	Feeder	Gravimetric	100 tph	2

ACCOUNT 2B LIMESTONE FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Storage Silo	Cylindrical	1,000 Ton	1
2	Weigh Feeder	Gravimetric	20 tph	1
3	Bin Activator		20 tph	
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Per Alstom steam cycle.

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 2,500 gpm	2
14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	100,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Design Condition	Qty
1	Power Boiler	AFBC	Alstom Design	
2	Gas Recirculation Fan	Centrifugal	217,000 lbm/hr 35,753 acfm 97.5 in. w.g 800 hp	1
3	Fluidizing Gas Blower	Centrifugal	44,825 lbm/hr 7,385 acfm 11.7 w.g 500 hp	2
4	ID Fan	Centrifugal	764,579 lbm/hr 123,330 acfm 38.4 in. w.g 1,200 hp	1

ACCOUNT 5 FLUE GAS CLEANUP

ACCOUNT 5A PARTICULATE CONTROL

Equipment No.	Description	Type	Design Condition	Qty
1	Bag Filter	Pulse-jet cleaned	171,000 acfm, total removal efficiency = 99.9%+	1

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Not applicable.

ACCOUNT 7 DUCTING AND STACK

In Gas Processing scope of supply

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	209 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	1,186 MM Btu/hr 100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	60,000 gpm @ 80 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

In boiler scope of supply.

ACCOUNT 10B FLY ASH HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Baghouse Hoppers (part of baghouse scope of supply)			4
2	Air Heater Hopper (part of boiler scope of supply)			1
3	Air Blower		1,800 cfm	1
4	Fly Ash Silo	Reinforced concrete	100 ton	1
5	Slide Gate Valves			2
6	Wet Unloader		30 tph	1
7	Telescoping Unloading Chute			1

9.1.4. Case-4 Equipment List

9.1.4.1. Case-4 Boiler Island Equipment

Fuel Feeding System:
- Day Silo
- Fuel Silo Isolation Valves
- Fuel Conveyors & Feeders
- Feeder Isolation Valves
- Piping to Furnace
Limestone Feeding System:
- Day Silo
- Limestone Silo Isolation Valves
- Rotary Feeder
- Blower
- Piping from Blower to Furnace Injection Points
Furnace Loop Equipment:
- Furnace Grate and Plenum Including Air Nozzles & Bauxite Drain Tubes
- Ash Drain Valve(s)
- Start-up Burner System (Including Burners, Piping, Ducts, and Local Control Equipment)
- Ductwork – Furnace to Recycle Particle Separators
- Particle Separator – Complete
- Ductwork – Recycle Particle Separator to Oxygen Heater Inlet
- Metal/Fabric Expansion Joints
- Seal pots and Seal pot Grate – Including Air Nozzles and Plenum
- Buckstay System
MBHE Equipment:
- Drum Including Internals, Nozzles, Lugging, Hanger Rods
- Downcomer System & Circulation Pumps/Drives
- Connecting Tubes/Piping
- Tube Panels/Headers
- Buckstay System
- MBHE Heat Absorbing Surface:
Horizontal Superheaters
Horizontal Reheater
Horizontal Evaporator
Horizontal Economizer
- Superheater/Reheater Desuperheaters
- Desuperheater Block Valves
- Desuperheater Piping
- Economizer Piping to Drum
- Superheater Interconnecting Piping
- Feedwater Stop, Feedwater Check valves
- Safety Valves/Discharge Piping/Silencers
- Electro. Relief Valve/Silencer and Discharge Piping
Bauxite Return System:
-Transport Air Blower/Drive (by others)
-Transport Air Heater
-Bauxite Removal Cyclones

-Ductwork
TA Blower to TA Heater
TA Heater to MBHE Bauxite Outlet
Bauxite transport pipes from MBHE outlet to Cyclone separator
Cyclone gas outlet to TA Heater
TA Heater gas vent
Trim Valves:
- Double Valving
Drum Level Gauge and Indicators
Sootblowing System:
- Oxygen heater
- Sootblower Control System
Oxygen System:
- Gas Recirculation Fan w/Drive (by others)
- Fluidizing Gas Blower w/Drive (by others)
- Fan and Blower Inlet Silencers (by others)
- Tubular Oxygen Heater
- Ductwork – GR Fan Outlet(s) to Oxygen Heater Inlet(s)
- Ductwork – FG Blower Outlets to Seal pots
- Steam Coil Oxygen Preheater
- Oxygen Duct Expansion Joints
Combustion Gas System:
- Ductwork and Expansion Joints - Separator Outlet to Oxygen Heater
- Ductwork – Oxygen Heater Outlet (including O_2 heater plenum & hoppers)
- Ductwork O_2 heater Outlet to Baghouse Inlet
- Ductwork Baghouse Outlet to PFWH Inlet
- PFWH economizer
- Ductwork PFWH Outlet to Gas Cooler Inlet
- Gas Cooler (by others)
- Ductwork Gas Cooler Outlet to ID Fan Inlet
- I.D. Fan w/Drive (by others)
- Ductwork ID Fan Outlet to GR Fan and FG Fan Inlet
Ash Handling System:
- Bed Ash Drains and Ash Coolers
Structural:
- Structural Steel including platforms, walkways, stairways, and ladders
- Boiler Internal Grid Steel
- Boiler Island Elevator
- Pressure Part Support Steel
- Boiler Building Siding, Weather Enclosure, HVAC
Instrumentation and Controls:
- Burner Management (FBSS) Logic
- CFB Field Instruments
- Controller Drives
Refractories:
- Material for All Internal Refractory Linings for Furnished Process and Boiler Equipment

Insulation and Lagging:
- Material for Insulation and Lagging for Heat Conservation and Personnel Protection for furnished equipment
Painting:
- Shop Prime Paint Coating for Seller furnished Equipment
Miscellaneous:
- Operator Training Program
- Maintenance Training Program
- Instruction Manuals
- Spare Parts for commissioning
- Technical Representation during start-up and testing
- Field Erection of Equipment Scope
- Freight to Site

9.1.4.2. Case-4 Gas Processing System Equipment

Tag No.	Service	Sizing Parameters	MOC
DA	Columns and Towers		
DA-101	Direct Contact Flue Gas Cooler	18' 6" ID x 34' S/S, DP 10 psig, 3 psi vacuum	CS w/ SS liner
DA-102	CO2 Rectifier	4'/ 12' ID x 30'/ 12' S/S, DP 425 psig	LTCS
E	Heat Transfer Equipment		
EA	Shell & Tube Exchangers		
EA-101	Flue Gas Compressor 1 Stage Trim Cooler	7.5 MMBTU/h, DP S/T, 85 psig/ 125 psig	CS/SS
EA-102	Flue Gas Compressor 2 Stage Trim Cooler	4.1 MMBTU/HR, DP S/T, 175 psig/ 125 psig	CS/SS
EA-103	Flue Gas Compressor 3 Stage Trim Cooler	4.3 MMBTU/HR, DP S/T, 425 psig/ 125 psig	CS/SS
EA-104	CO2 Condenser	61.7 MMBTU/HR, DP S/T, 300 psig/ 425 psig	LTCS/ LTCS
EA-201	Refrig condenser	71.0 MMBTU/HR, DP S/T, 300 psig/ 125 psig	CS/CS
EA-202	Refrig Subcooler	20.0 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/ LTCS
EA-107	CO2 Rectifier Condenser	5.0 MMBTU/HR, DP S/T, 425 psig/ 425 psig	SS/SS
EA -108	Rectifier Ovhd Interchanger	1.6 MMBTU/HR, DP S/T, 425 psig/ 425 psig	LTCS/ SS
EB	Plate Exchangers		
EB-101	Water Cooler	Total 56 MMBTU/HR, DP P/U, 65 psig/ 125 psig	CS
EC	Air Coolers		
EC-101	Flue Gas Compressor 1 Stage Aftercooler	25.2 MMBTU/HR, DP 85 psig	SS
EC-102	Flue Gas Compressor 2 Stage Aftercooler	21.2 MMBTU/HR, DP 175 psig	SS
EC-103	Flue Gas Compressor 3 Stage Aftercooler	21.0 MMBTU/HR, DP 425 psig	SS
FH	Heaters		
FH-101	Dryer Regeneration Gas Heater	Gas fired, 5.75 MMBTU/HR fired duty	
FA	Drums and Vessels		
FA-101	Flue Gas Compressor 2nd Stage Suction Drum	13' - 6" ID x 18' S/S, DP 85 psig	CS w/ SS liner
FA-102	Flue Gas Compressor 3rd Stage Suction Drum	10'-0" ID x 14' S/S DP 175 psig	CS w/ SS liner
FA-103	Flue Gas Compressor Third Stage Discharge K/O Drum	8' ID x 12' S/S, DP 425 psig	CS w/ SS liner
FA-201	Refrig Surge Drum	12' ID x 30' S/S DP 300 psig	CS
FA-202	Refrig Suction Srubber	16' ID x 24' S/S, DP 300 psig	ITCS
FD	Filters and Dryers		
FD-101	Water Filter	5 units, 805 gpm each, DP 100 psig	SS
FD-102	Flue Gas Filter	2665 ACFM, DP 425 psig	CS
FF	Dryers (Dessicant Type)		
FF-101	Flue Gas Dryer	Two Vessels 10' - 0 " ID x 20' S/S DP 425 psig DT 500 F	CS
GA	Pumps Centrifugal		
GA-101	Water Pump	4180 gpm, DP 40 psi	CI w/ SS impeller
GA-103	CO2 Pipeline pump	745 gpm, DP 1655 psi	ITCS
GB	Compressors & Blowers		
GB-101	Flue Gas Compressor	Motor Drive 3 stages Includes Lube/Seal Oil Systems, 19715 kW	SS
GB-102	Propane Refrig Compressor	Motor Drive, Includes lube oil/ seal oil system, 9095 kW	ITCS

9.1.4.3. Case-4 Air Separation Unit Equipment

The Case-4 Air Separation Unit equipment is identical to Case-2 and is not repeated here. Refer to section 9.1.2.3 for the relevant ASU equipment list.

9.1.4.4. Case-4 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-4 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor 1	54" belt	450 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	2
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	200 tph	2
3	Conveyor 4	48" belt	200 tph	1
4	Transfer Tower	N/A	200 tph	1
5	Tripper	N/A	200 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	500 ton	2
7	Feeder	Gravimetric	100 tph	2

ACCOUNT 2B LIMESTONE FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Storage Silo	Cylindrical	1,000 Ton	1
2	Weigh Feeder	Gravimetric	20 tph	1
3	Bin Activator		20 tph	
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Per Alstom steam cycle.

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 2,500 gpm	2
14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	100,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Design Condition	Qty
1	Power Boiler	AFBC	Alstom Design	
2	Gas Recirculation Fan	Centrifugal	217,000 lbm/hr 35,753 acfm 97.5 in. w.g 800 hp	1
3	ID Fan	Centrifugal	764,579 lbm/hr 123,330 acfm 38.4 in. w.g 1,200 hp	1
4	Fluidizing Gas Blower	Centrifugal	44,825 lbm/hr 7,385 acfm 11.7 psia 500 hp	1
5	Transport Air Blower	Centrifugal	440,626 lbm/hr 96,785 acfm 210 in. w.g. 3,000 hp	1

ACCOUNT 5 FLUE GAS CLEANUP

ACCOUNT 5A PARTICULATE CONTROL

Equipment No.	Description	Type	Design Condition	Qty
1	Bag Filter	Pulse-jet cleaned	171,000 acfm, total removal efficiency = 99.9%+	1

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Not applicable.

ACCOUNT 7 DUCTING AND STACK

In Gas Processing scope of supply

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	210 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	1,294 MM Btu/hr 100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	60,000 gpm @ 60 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

In boiler scope of supply.

ACCOUNT 10B FLY ASH HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Baghouse Hoppers (part of baghouse scope of supply)			10
2	Air Heater Hopper (part of boiler scope of supply)			1
3	Air Blower		1,800 cfm	1
4	Fly Ash Silo	Reinforced concrete	400 tons	1
5	Slide Gate Valves			2
6	Wet Unloader		30 tph	1
7	Telescoping Unloading Chute			1

9.1.5. Case-5 Equipment List

9.1.5.1. Case-5 Boiler Island Equipment

Fuel Feeding System:
- Day Silos (3)
- Fuel Silo Isolation Valves
- Fuel Feeders
- Feeder Isolation Valves
- Conveyance to Furnace
Limestone Feeding System:
- Day Silo (1)
- Limestone Silo Isolation Valves
- Rotary Feeder
- Blower
- Conveyance from Blower to Fuel Feeders
Furnace Loop Equipment:
-Combustor
-DeCarbonator and Sealpot
- Ring Cone Separators
- Calciner and Sealpot
-Ash Drain Valve(s)
-Start-up Burner System (Including Burners, Piping, Ducts, and Local Control Equipment)
- Metal/Fabric Expansion Joints
MBHE Equipment:
- Connecting Tubes/Pipes
- Drum Including Internals, Nozzles, and Hanger Rods
–Downcomer System
MBHE Heat Absorbing Surface:
Horizontal Superheaters
Horizontal Reheater
Horizontal Evaporator
Horizontal Economizer
- Superheater/Reheater Desuperheaters
- Desuperheater Block Valves
- Desuperheater Piping
- Economizer Piping to Drum
- Superheater Interconnecting Piping
- Feedwater Stop, Feedwater Check valves
- Safety Valves/Discharge Piping/Silencers
- Circulation Pumps
- Trim Valves and Piping
- Drum Level Gauge and Indicators
Solids Ducts:
- Duct from DeCarbonator to Ring Cone Separator to Sealpot
- Duct from DeCarbonator Sealpot to Calciner
- Lift Duct from MBHE to DeCarbonator Sealpot

Gas Ducts:
- Flue Gas Duct from DeCarbonator to Ring Cone Separator to Air Heater
- Flue Gas Duct from Air Heater to Stack (By Others)
- CO_2 Duct from Calciner to Air Heater
- CO_2 Duct from Air Heater to Gas Processing System (By Others)
Air Ducts:
- Hot Air from Air Heater to Combustor
- PA Fan w/Drive (By Others)
- I.D. Fan w/Drive (by others)
- Expansion Joints
Ash Handling System:
- Bed Ash Drains and Ash Screw Coolers
Structural:
- Structural Steel Including Platforms, Walkways, Stairways, and Ladders
- Boiler Island Elevator
Instrumentation and Controls:
- Burner Management (FBSS) Logic
- Field Instruments
- Controller Drives
Refractories:
- Material for All Internal Refractory Linings for Furnished Process and Boiler Equipment
Insulation and Lagging:
- Material for Insulation and Lagging for Heat Conservation and Personnel Protection for Furnished Equipment, Materials Testing and Installation
Painting:
- Shop Prime Paint Coating for Seller Furnished Equipment
Miscellaneous:
- Operator Training Program
- Maintenance Training Program
- Instruction Manuals
- Spare Parts for Commissioning
- Technical Representation During Start-up and Testing
- Field Erection of Equipment Scope
- Freight to Site

9.1.5.2. Case-5 Gas Processing System Equipment

Tag no.	Description	Size Parameters	Material
EA-2301	CO2 Compr. 1st Stage Aftercooler	8 MMBTU/HR, DP S/T, 75 psig/ 125 psig	SS/CS
EA-2302	CO2 Compr. 2nd Stage Aftercooler	4.1 MMBTU/HR, DP S/T, 125 psig/ 125 psig	SS/CS
EA-2303	CO2 Compr. 3rd Stage Aftercooler	3.7 MMBTU/HR, DP S/T, 235 psig/ 125 psig	CS/CS
EA-2304	CO2 Condenser	63.8 MMBTU/hr DP S/T, 300 psig/ 235 psig	CS/CS
EA-2401	Propane Refrig Condenser	77.45 MMBTU/HR, DP S/T, 300 psig/ 125 psig	CS/CS
EA-2402	Propane Refrig Subcooler	19.7 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
EC-2301	CO2 compressor 1st stage air cooler	31.0 MMBTU/HR, DP 75 psig	SS
EC-2302	CO2 compressor 2nd stage air cooler	17 MMBTU/HR, DP 125 psig	SS
EC-2303	CO2 compressor 3rd stage air cooler	12 MMBTU/HR, DP 235 psig	SS
FA-2300	CO_2 Compressor 1st Stage Suction Drum	15'-0" ID x 20' S/S, DP 75 psig	CS/SS
FA-2301	CO_2 Compressor 2st Stage Suction Drum	12'- 0" ID x 16' S/S, DP 75 psig	CS/SS
FA-2302	CO_2 Compressor 3rd Stage Suction Drum	9'- 6" ID x 14' S/S, DP 125 psig	CS/SS
FA-2303	Liquid CO2 Surge Drum	8'- 0" ID x 25' S/S, DP 235 psig	KCS
FA-2304	CO2 Compressor 3rd stage Discharge KO Drum	7' 6" ID x 12' S/S, DP 235 psig	CS/SS
FA-2401	Propane Refrig Surge Drum	9' ID x 20' S/S, DP 300 psig	CS
FA-2402	Propane Refrig Suction Scubber	16' ID x 20' S/S, DP 300 psig	LTCS
FA-2403	Propane Refrigeration Economizer	10' ID x 20' S/S DP 300 psig	CS
GA-2301	CO2 Pipeline Pump	730 gpm, DP 1830 psi, 1040 hp	LTCS/CS
GB-2301	CO2 Compressor (Motor driven)	17800 hp	SS wheels
GB-2401	Propane Refrig Compressor	12775 hp	LTCS
PA-2351	CO2 Dryer Package	4 90" x 20' driers DP 235 psig, 800 hp sundyne compressor, 7.9 MMBTU/h heater, 9 MMBTU/h air cooler, 36" x 8' KO drum, Dust filter 3820 ACFM	CS w/ SS liner

9.1.5.3. Case-5 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-5 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor 1	54" belt	450 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	2
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	200 tph	2
3	Conveyor 4	48" belt	200 tph	1
4	Transfer Tower	N/A	200 tph	1
5	Tripper	N/A	200 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	500 ton	2
7	Feeder	Gravimetric	100 tph	2

ACCOUNT 2B LIMESTONE FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Storage Silo	Cylindrical	1,000 Ton	1
2	Weigh Feeder	Gravimetric	20 tph	1
3	Bin Activator		20 tph	
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Per Alstom steam cycle.

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 2,500 gpm	2
14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	100,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Design Condition	Qty
1	Power Boiler	AFBC	Alstom Design	
2	Primary Air Fan	Centrifugal	1,947,550 pph, 404,000 acfm, 45" wg, 3,300 hp	1
3	Transport Air Blower	Centrifugal	19,881 pph, 4,100 acfm, 315" wg, 200 hp	1
4	ID Fan	Centrifugal	1,791,436 pph, 450,000 acfm, 39" wg 12,000 hp	1
5	Fluidizing Air Blower	Centrifugal	16,000 acfm/25 psig 1,800 hp	2

ACCOUNT 5 FLUE GAS CLEANUP

ACCOUNT 5A PARTICULATE CONTROL

Not applicable – within boiler scope.

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Not applicable.

ACCOUNT 7 DUCTING AND STACK

Equipment No.	Description	Type	Design Condition	Qty
1	Boiler Stack	Concrete with Steel liner	160°F 500,000 acfm	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	203 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	1004 MM Btu/hr 100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	50,000 gpm @ 60 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

In boiler scope of supply.

ACCOUNT 10B ASH HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Air Blower		1,800 cfm	1
2	Bottom Drain Silo	Reinforced concrete	1,200 tons	1
3	Slide Gate Valves			2
4	Wet Unloader		30 tph	1
5	Telescoping Unloading Chute			1

9.1.6. Case-6 Equipment List

9.1.6.1. Case-6 Boiler Island Equipment

Fuel Feeding System:
- Day Silo
- Fuel Silo Isolation Valves
- Fuel Conveyors & Feeders
- Feeder Isolation Valves
- Piping to Furnace
Limestone Feeding System:
- Day Silo
- Limestone Silo Isolation Valves
- Rotary Feeder
- Blower
- Piping from Blower to Furnace Injection Points
Furnace Loop Equipment:
- Furnace Grate and Plenum Including Air Nozzles & Bauxite Drain Tubes
- Ash Drain Valve(s)
- Start-up Burner System (Including Burners, Piping, Ducts, and Local Control Equipment)
- Ductwork – Furnace to Recycle Particle Separators
- Particle Separator – Complete
- Ductwork – Recycle Particle Separator to Oxygen Heater Inlet
- Metal/Fabric Expansion Joints
- Seal pots and Seal pot Grate – Including Air Nozzles and Plenum
- Buckstay System
MBHE Equipment:
- Drum Including Internals, Nozzles, Lugging, Hanger Rods
- Downcomer System & Circulation Pumps/Drives
- Connecting Tubes/Piping
- Tube Panels/Headers
- Buckstay System
- MBHE Heat Absorbing Surface:
HT Air Heater
Horizontal Superheaters
Horizontal Reheater
Horizontal Evaporator
Horizontal Economizer
- Air Heater Piping to Oxygen Transport Membrane
- Superheater/Reheater Desuperheaters
- Desuperheater Block Valves
- Desuperheater Piping
- Economizer Piping to Drum
- Superheater Interconnecting Piping
- Feedwater Stop, Feedwater Check valves
- Safety Valves/Discharge Piping/Silencers
- Electro. Relief Valve/Silencer and Discharge Piping
Bauxite Return System:
-Transport Air Blower/Drive (by others)

-Tubular Transport Air Heater
-Bauxite Removal Cyclones
-Ductwork
TA Blower to TA Heater
TA Heater to MBHE Bauxite Outlet
Bauxite transport pipes from MBHE outlet to Cyclone separator
Cyclone gas outlet to TA Heater
TA Heater gas vent
Trim Valves:
- Double Valving
Drum Level Gauge and Indicators
Sootblowing System:
- Oxygen heater
- Sootblower Control System
Sweep Gas System:
- Sweep Gas Fan w/Drive (by others)
- Fan and Blower Inlet Silencers (by others)
- Steam Coil SG Preheater
- Tubular LT Sweep Gas Heater
- Ductwork – GR Fan Outlet(s) to LT SG Heater Inlet(s)
- Ductwork – LT SG Heater Outlet to HT SG Heater Inlet (cold side)
- Tubular HT SG Heater
- Ductwork – HT SG Heater Outlet (cold side) to OTM
- Oxygen Transport Membrane (by others)
- Ductwork – OTM outlet to HT SG Heater Inlet (hot side)
- Ductwork –HT SG Heater Inlet (hot side) to Combustor
- SG Duct Expansion Joints
Combustion Gas System:
- Ductwork and Expansion Joints - Cyclone Outlet to LT SG Heater Inlet
- Ductwork – LT SG Heater Outlet (including SG heater plenum & hoppers)
- Ductwork LT SG heater Outlet to Baghouse Inlet
- Ductwork Baghouse Outlet to PFWH Inlet
- HT & LT PFWH economizers
- Ductwork PFWH Outlet to Gas Cooler Inlet
- Gas Cooler (by others)
- Ductwork Gas Cooler Outlet to ID Fan Inlet
- I.D. Fan w/Drive (by others)
- Ductwork ID Fan Outlet to SG Fan and FG Fan Inlet
- Ductwork SG Fan outlet to LT SG Heater Inlet
- Fluidizing Gas Blower w/Drive (by others)
- Sweep Gas Fan w/Drive (by others)
- Ductwork – FG Blower Outlets to Seal pot
Ash Handling System:
- Bed Ash Drains and Ash Coolers
Structural:
- Structural Steel including platforms, walkways, stairways, and ladders
- Boiler Internal Grid Steel
- Boiler Island Elevator
- Pressure Part Support Steel

- Boiler Building Siding, Weather Enclosure, HVAC
Instrumentation and Controls:
- Burner Management (FBSS) Logic
- CFB Field Instruments
- Controller Drives
Refractories:
- Material for All Internal Refractory Linings for Furnished Process and Boiler Equipment
Insulation and Lagging:
- Material for Insulation and Lagging for Heat Conservation and Personnel Protection for furnished equipment
Painting:
- Shop Prime Paint Coating for Seller furnished Equipment
Miscellaneous:
- Operator Training Program
- Maintenance Training Program
- Instruction Manuals
- Spare Parts for commissioning
- Technical Representation during start-up and testing
- Field Erection of Equipment Scope
- Freight to Site

9.1.6.2. Case-6 Gas Processing System Equipment

Tag No.	Service	Sizing Parameters	MOC
DA	Columns and Towers		
DA-101	Direct Contact Flue Gas Cooler	22' 0" ID x 34' S/S, DP 10 psig, 3 psi vacuum	CS w/ SS liner
DA-102	CO2 Rectifier	4'/ 12' ID x 30'/ 13' S/S, DP 425 psig	LTCS
E	Heat Transfer Equipment		
EA	Shell & Tube Exchangers		
EA-101	Flue Gas Compressor 1 Stage Trim Cooler	9.6 MMBTU/h, DP S/T, 85 psig/ 125 psig	CS/SS
EA-102	Flue Gas Compressor 2 Stage Trim Cooler	5.25 MMBTU/HR, DP S/T, 175 psig/ 125 psig	CS/SS
EA-103	Flue Gas Compressor 3 Stage Trim Cooler	5.2 MMBTU/HR, DP S/T, 425 psgi/ 125 psig	CS/SS
EA-104	CO2 Condenser	76.1 MMBTU/HR, DP S/T, 300 psig/ 425 psig	LTCS/ LTCS
EA-201	Refrig condenser	88.6 MMBTU/HR, DP S/T, 300 psig/ 125 psig	CS/CS
EA-202	Refrig Subcooler	24.1 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/ LTCS
EA-107	CO2 Rectifier Condenser	6.1 MMBTU/HR, DP S/T, 425 psig/ 425 psig	SS/SS
EA -108	Rectifier Ovhd Interchanger	1.55 MMBTU/HR, DP S/T, 425 psig/ 425 psig	LTCS/ SS
EB	Plate Exchangers		
EB-101	Water Cooler	Total 63.7 MMBTU/HR, DP P/U, 65 psig/ 125 psig	CS
EC	Air Coolers		
EC-101	Flue Gas Compressor 1 Stage Aftercooler	31.1 MMBTU/HR, DP 85 psig	SS
EC-102	Flue Gas Compressor 2 Stage Aftercooler	25.25 MMBTU/HR, DP 175 psig	SS
EC-103	Flue Gas Compressor 3 Stage Aftercooler	25.8 MMBTU/HR, DP 425 psig	SS
FH	Heaters		
FH-101	Dryer Regeneration Gas Heater	Gas fired, 7.15 MMBTU/HR fired duty	
FA	Drums and Vessels		
FA-100	Flue Gas Compressor 1st Stage Suction Drum	17' 6" ID x 20' S/S, DP 75 psig	CS w/ SS liner
FA-101	Flue Gas Compressor 2nd Stage Suction Drum	13' - 6" ID x 18' S/S, DP 85 psig	CS w/ SS liner
FA-102	Flue Gas Compressor 3rd Stage Suction Drum	10' - 6" ID x 14' S/S DP 175 psig	CS w/ SS liner
FA-103	Flue Gas Compressor Third Stage Discharge K/O Drum	8' ID x 12' S/S, DP 425 psig	CS w/ SS liner
FA-201	Refrig Surge Drum	13' ID x 30' S/S DP 300 psig	CS
FA-202	Refrig Suction Srubber	17' ID x 24' S/S, DP 300 psig	ITCS
FD	Filters and Dryers		
FD-101	Water Filter	6 units, 790 gpm each, DP 100 psig	SS
FD-102	Flue Gas Filter	Total 3360 ACFM, DP 425 psig	CS
FF	Dryers (Dessicant Type)		
FF-101	Flue Gas Dryer	Two Vessels 11' -0 " ID x 20' S/S DP 425 psig DT 500 F	CS
GA	Pumps Centrifugal		
GA-101	Water Pump	4725 gpm, DP 40 psi	CI w/ SS impeller
GA-103	CO2 Pipeline pump	925 gpm, DP 1663 psi	ITCS
GB	Compressors & Blowers		
GB-101	Flue Gas Compressor	Motor Drive 3 stages Includes Lube/Seal Oil Systems, 22850 kW	SS
GB-102	Propane Refrig Compressor	Motor Drive, Includes lube oil/ seal oil system, 10765 kW	ITCS

9.1.6.3. Case-6 Oxygen Transport Membrane System Equipment

Oxygen Transport Membrane:
Output: ~ 4800 tons/day Oxygen Output
Oxygen Purity: ~100%
Nominal Overall Dimensions: Length = ~ 121ft, Width ~ 61ft, Height ~ 96ft

The Oxygen Transport Membrane System Support Equipment (Air Compressor, Gas Expander, Generator, Compressor Motor Drive, Heat Recovery System, etc.) is included in the Balance of Plant Equipment List shown in Section 9.1.6.4.

9.1.6.4. Case-6 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-6 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	250 ton	2
2	Feeder	Vibratory	550 tph	2
3	Conveyor 1	54" belt	550 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	550 tph	2
6	Reclaim Hopper	N/A	50 ton	2
7	Feeder	Vibratory	400 tph	2
8	Conveyor 3	48" belt	400 tph	1
9	Crusher Tower	N/A	400 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	400 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	250 tph	2
3	Conveyor 4	48" belt	250 tph	1
4	Transfer Tower	N/A	250 tph	1
5	Tripper	N/A	250 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	600 ton	2
7	Feeder	Gravimetric	150 tph	2

ACCOUNT 2B LIMESTONE FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Storage Silo	Cylindrical	1,000 Ton	1
2	Weigh Feeder	Gravimetric	25 tph	1
3	Bin Activator		25 tph	
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Per Alstom steam cycle.

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 1,000 gpm	2
14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	15,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Design Condition	Qty
1	Power Boiler	CMB	Alstom Design	
2	OTM Air Compressor	Centrifugal	1,025,000 lbm/hr 250,000 acfm 215 psia 75,000 hp	2
3	OTM Gas Expander	Turbine Generator	1,650,000 lbm/hr 122,659 kW net	1
4	Sweep Gas Fan	Centrifugal	277,975 lbm/hr 46,498 ACFM 143.8 in. w.g. 1,100 hp	1
5	ID Fan	Centrifugal	941,784 lbm/hr 148,723 acfm 29.5 in. w.g 850 hp	1
6	Fluidizing Gas Blower	Centrifugal	44,825 lbm/hr 7,385 acfm 11.7 w.g 500 hp	1
7	Transport Air Blower	Centrifugal	571,284 lbm/hr 125,300 acfm 210 in. w.g. 3,300 hp	1
8	OTM HT-PFWH	Feedwater Heater	168.2 MMBtu/h	1
9	OTM LT-PFWH	Feedwater Heater	71.6 MMBtu/h	1

ACCOUNT 5 FLUE GAS CLEANUP

ACCOUNT 5A PARTICULATE CONTROL

Equipment No.	Description	Type	Design Condition	Qty
1	Bag Filter	Pulse-jet cleaned	235,000 acfm, total removal efficiency = 99.9%+	1

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Not applicable.

ACCOUNT 7 DUCTING AND STACK

In Gas Processing scope of supply

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	234 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	1,614 MMBtu/h 100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	80,000 gpm @ 80 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

In boiler scope of supply.

ACCOUNT 10B FLY ASH HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Baghouse Hoppers (part of baghouse scope of supply)			10
2	Air Heater Hopper (part of boiler scope of supply)			1
3	Air Blower		1,800 cfm	1
4	Fly Ash Silo	Reinforced concrete	400 tons	1
5	Slide Gate Valves			2
6	Wet Unloader		30 tph	1
7	Telescoping Unloading Chute			1

9.1.7. Case-7 Equipment List

9.1.7.1. Case-7 Boiler Island Equipment

Fuel Feeding System:
- Day Silos (3)
- Fuel Silo Isolation Valves
- Fuel Feeders
- Feeder Isolation Valves
- Conveyance to Furnace
Limestone Feeding System:
- Day Silo (1)
- Limestone Silo Isolation Valves
- Rotary Feeder
- Blower
- Conveyance from Blower to Fuel Feeders
Furnace Loop Equipment:
-Combustor (Oxidizer)
-Reductor
-Ring Cone Separators and Sealpots for Combustor and Reductor
-Ash Drain Valve(s)
-Start-up Burner System (Including Burners, Piping, Ducts, and Local Control Equipment) for Reductor
- Metal/Fabric Expansion Joints
MBHE Equipment:
- Connecting Tubes/Pipes
- Drum Including Internals, Nozzles, and Hanger Rods
–Downcomer System
MBHE Heat Absorbing Surface:
Horizontal Superheaters
Horizontal Reheater
Horizontal Evaporator
Horizontal Economizer
- Superheater/Reheater Desuperheaters
- Desuperheater Block Valves
- Desuperheater Piping
- Economizer Piping to Drum
- Superheater Interconnecting Piping
- Feedwater Stop, Feedwater Check valves
- Safety Valves/Discharge Piping/Silencers
- Circulation Pumps
- Trim Valves and Piping
- Drum Level Gauge and Indicators
Solids Ducts:
- Transport Duct from Combustor Sealpot to Reductor
- Transport Duct form Reduced to Ring Cone Separator to MBHE Sealpot
- Return Duct form Combustor Sealpot to Reducer
- Bypass Duct from Combustor Sealpot to Combustor
- ash Transport Duct from MBHE to Combustor

Gas Ducts:
- Flue Gas Duct from Combustor to Air Heater
- Flue Gas Duct from Air Heater to Stack (By Others)
- CO_2 Duct from Reductor to Air Heater
- CO_2 Duct from Air Heater to Gas Processing System (By Others)
Air Ducts:
- Reductor Startup Duct
- Hot Air from Air Heater to Combustor
- PA Fan w/Drive (By Others)
- I.D. Fan w/Drive (by others)
- Expansion Joints
Ash Handling System:
- Bed Ash Drains and Ash Screw Coolers
Structural:
- Structural Steel Including Platforms, Walkways, Stairways, and Ladders
- Boiler Island Elevator
Instrumentation and Controls:
- Burner Management (FBSS) Logic
- Field Instruments
- Controller Drives
Refractories:
- Material for All Internal Refractory Linings for Furnished Process and Boiler Equipment
Insulation and Lagging:
- Material for Insulation and Lagging for Heat Conservation and Personnel Protection for Furnished Equipment, Materials Testing and Installation
Painting:
- Shop Prime Paint Coating for Seller Furnished Equipment
Miscellaneous:
- Operator Training Program
- Maintenance Training Program
- Instruction Manuals
- Spare Parts for Commissioning
- Technical Representation During Start-up and Testing
- Field Erection of Equipment Scope
- Freight to Site

9.1.7.2. Case-7 Gas Processing System Equipment

Number of Trains	Tag no.	Description	Size Parameters	Material
	DA-101	Direct Contact Flue Gas Cooler	18' 6" ID x 34' S/S, DP 10 psig, 3 psi vacuum	CS w/ SS liner
1	EA-2301	CO2 Compr. 1st Stage Aftercooler	6.9 MMBTU/HR, DP S/T, 75 psig/ 125 psig	SS/CS
1	EA-2302	CO2 Compr. 2nd Stage Aftercooler	4.1 MMBTU/HR, DP S/T, 150 psig/ 125 psig	SS/CS
1	EA-2303	CO2 Compr. 3rd Stage Aftercooler	3.7 MMBTU/HR, DP S/T, 350 psig/ 125 psig	CS/CS
1	EA-2304	CO2 Condenser	66.0 MMBTU/hr DP S/T, 300 psig/ 350 psig	CS/CS
1	EA-2401	Propane Refrig Condenser	77.7 MMBTU/HR, DP S/T, 300 psig/ 125 psig	CS/CS
1	EA-2402	Propane Refrig Subcooler	20.75 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
1	EC-2301	CO2 compressor 1st stage air cooler	20.7 MMBTU/HR, DP 75 psig	SS
1	EC-2302	CO2 compressor 2nd stage air cooler	17.4 MMBTU/HR, DP 150 psig	SS
1	EC-2303	CO2 compressor 3rd stage air cooler	15.8 MMBTU/HR, DP 350 psig	SS
1	EB-101	Water Cooler	Total 99 MMBTU/HR, DP P/U, 65 psig/ 125 psig	CS
1	FA-2300	CO_2 Compressor 1st Stage Suction Drum	10'-0" ID x 20' S/S, DP 75 psig	CS/SS
1	FA-2301	CO_2 Compressor 2st Stage Suction Drum	10'- 0" ID x 18' S/S, DP 75 psig	CS/SS
1	FA-2302	CO_2 Compressor 3rd Stage Suction Drum	9'- 6" ID x 14' S/S, DP 150 psig	CS/SS
1	FA-2303	Liquid CO2 Surge Drum	8'- 0" ID x 25' S/S, DP 350 psig	KCS
1	FA-2304	CO2 Compressor 3rd stage Discharge KO Drum	9' 6" ID x 12' S/S, DP 350 psig	CS/SS
1	FA-2401	Propane Refrig Surge Drum	9' ID x 20' S/S, DP 300 psig	CS
1	FA-2402	Propane Refrig Suction Scubber	16' ID x 20' S/S, DP 300 psig	LTCS
1	FA-2403	Propane Refrigeration Economizer	10' ID x 22' S/S DP 300 psig	CS
1	FD-101	Water Filter	6 units, 830 gpm each, DP 100 psig	SS
1	GA-2301	CO2 Pipeline Pump	800 gpm, DP 1725 psi, 1070 hp	LTCS/CS
1	GA-101	Water Pump	5000 gpm, DP 40 psi	CI w/ SS impeller
1	GB-2301	CO2 Compressor (Motor driven)	20925 hp	SS wheels
1	GB-2401	Propane Refrig Compressor	12700 hp	LTCS
1	PA-2351	CO2 Dryer Package	4 90" x 15' driers DP 350 psig, 350 hp sundyne compressor, 5.8 MMBTU/h heater, 6.4 MMBTU/h air cooler, 30" x 8' KO drum, Dust filter 2575 ACFM	CS w/ SS liner
1		Crane for Compr. Bldg		
		Flue gas ducting		

9.1.7.3. Case-7 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-7 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor 1	54" belt	450 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	2
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	200 tph	2
3	Conveyor 4	48" belt	200 tph	1
4	Transfer Tower	N/A	200 tph	1
5	Tripper	N/A	200 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	500 ton	2
7	Feeder	Gravimetric	100 tph	2

ACCOUNT 2B LIMESTONE FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Storage Silo	Cylindrical	1,000 Ton	1
2	Weigh Feeder	Gravimetric	20 tph	1
3	Bin Activator		20 tph	
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Per Alstom steam cycle.

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 2,500 gpm	2
14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	100,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Design Condition	Qty
1	Power Boiler	Chemical Looping	Alstom Design	
2	Primary Air Fan	Centrifugal	1,911,244 pph, 396,000 acfm, 61.1" wg, 4,100 hp	1
3	Transport Air Blower	Centrifugal	19,881 pph, 3,900 acfm, 113" wg, 75 hp	1
4	ID Fan	Centrifugal	1,636,694 pph, 438,000 acfm, 30" wg 2,100 hp	1

ACCOUNT 5 FLUE GAS CLEANUP

ACCOUNT 5A PARTICULATE CONTROL

Within boiler scope of supply.

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Not applicable.

ACCOUNT 7 DUCTING AND STACK

Equipment No.	Description	Type	Design Condition	Qty
1	Boiler Stack	Concrete with Steel liner	160°F 500,000 acfm	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	203 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	1,115 MM Btu/hr 100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	50,000 gpm @ 60 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

In boiler scope of supply.

ACCOUNT 10B ASH HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Air Blower		1,800 cfm	1
2	Bottom Drain Silo	Reinforced concrete	1,200 tons	1
3	Slide Gate Valves			2
4	Wet Unloader		30 tph	1
5	Telescoping Unloading Chute			1

9.1.8. Case-8 Equipment List

MAJOR EQUIPMENT LIST
ACCOUNT 1 COAL HANDLING

ACCOUNT 1A COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor No. 1	54" belt	900 tph	1
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor No. 2	54" belt	900 tph	1
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor No. 3	48" belt	600 tph	1
9	Crusher Tower	N/A	600 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	600 ton	1
11	Crusher	Granulator reduction	6"x0 - 3"x0	1
12	Crusher	Impactor reduction	3"x0 - 1¼"x0	1
13	As-Fired Coal Sampling System	Swing hammer		2
14	Conveyor No. 4	48" belt	600 tph	1
15	Transfer Tower	N/A	600 tph	1
16	Tripper	N/A	600 tph	1
17	Coal Silo w/ Vent Filter and Slide Gates	N/A	1,600 ton	3
18	Front–End Loader	Rubber tired	1,600 ton	3

ACCOUNT 2 COAL PREPARATION AND FEED

ACCOUNT 2A FUEL SLURRY PREPARATION AND FUEL INJECTION

Equipment No.	Description	Type	Design Condition	Qty
1	Vibratory Feeder		140 tph	3
2	Conveyor No. 1	Belt	280 tph	1
3	Conveyor No. 2	Belt	280 tph	1
4	Rod Mill Feed Hopper	Vertical, double hopper	300 tons	1
5	Vibratory Feeder		100 tph	2
6	Weight Feeder	Belt	100 tph	2
7	Rod Mill	Rotary	100 tph	2
8	Slurry Water Storage Tank with Agitator	Field erected	100,000 gal	1
9	Slurry Water Pumps	Horizontal, centrifugal	1,200 gpm	2
10	Rod Mill Product Tank with Agitator	Field erected	200,000 gal	1
11	Rod Mill Product Pumps	Horizontal, centrifugal	2,000 gpm	2
12	Slurry Storage Tank with Agitator	Field erected	350,000 gal	1
13	Centrifugal Slurry Pumps	Horizontal, centrifugal	3,000 gpm	2
14	PD Slurry Pumps	Progressing cavity	500 gpm	4
15	Slurry Blending Tank with Agitator	Field erected	100,000 gal	1
16	Slurry Blending Tank Pumps	Horizontal, centrifugal	450 gpm	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A CONDENSATE AND FEEDWATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cond. Storage Tank	Vertical, cylindrical, outdoor	50,000 gal	1
2	Condensate Pumps	Vert. canned	900 gpm @ 400 ft	2
3	Deaerator (integral with HRSG)	Horiz. spray type	820,000 lbm/h 210°F to 240°F	2
4	IP Feed Pump	Horiz. centrifugal single stage	200 gpm/1,000 ft	2
5	HP Feed Pump	Barrel type, multi-staged, centr.	900 gpm @ 5,100 ft & 50 gpm @ 1,700 ft	2

ACCOUNT 3B MISCELLANEOUS EQUIPMENT

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fab., water tube	400 psig, 650°F	1
2	Service Air Compressors	Recip., single stage, double acting, horiz.	100 psig, 450 cfm	2
3	Inst. Air Dryers	Duplex, regenerative	450 cfm	1
4	Service Water Pumps	Horiz. centrifugal, double suction	200 ft, 700 gpm	2
5	Closed Cycle Cooling Heat Exchangers	Plate and frame	50% cap. each	2
6	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	70 ft, 700 gpm	2
7	Fire Service Booster Pump	Two-stage horiz., centrifugal	250 ft, 700 gpm	1
8	Engine-Driven Fire Pump	Vert. turbine, diesel engine	350 ft, 1,000 gpm	1
9	Raw Water Pumps	SS, single suction	60 ft, 1,100 gpm	2
10	Filtered Water Pumps	SS, single suction	160 ft, 700 gpm	2
11	Filtered Water Tank	Vertical, cylindrical	50,000 gal	1
12	Makeup Demineralizer	Anion, cation, and mixed bed	700 gpm	2
13	Liquid Waste Treatment System		10 years, 25-hour storm	1

ACCOUNT 4 GASIFIER AND ACCESSORIES

ACCOUNT 4A GASIFICATION (total for plant)

Equipment No.	Description	Type	Design Condition	Qty
1	Gasifier and associated equipment	Texaco Pressurized entrained bed	2,250 ton/day/ 465 psia	2
2	Syngas Scrubber	Vertical, upflow	450,000 lbm/h	2
3	Low Temperature Gas Cooling	Syngas Coolers	300,000 scfm syngas	5
5	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	450,000 lbm/h, medium-Btu gas	1

ACCOUNT 4B AIR SEPARATION PLANT (total for plant)

Equipment No.	Description	Type	Design Condition	Qty
1	Air Compressor	Centrifugal, multi-stage	60,000 scfm, 90 psia discharge pressure	2
2	Cold Box	Vendor Design	2,200 ton/day O_2	1
3	Oxygen Compressor	Centrifugal, multi-stage	22,000 scfm, 650 psia discharge pressure	2

ACCOUNT 5 SYNGAS CLEANUP

Equipment No.	Description	Type	Design Condition	Qty
1	COS Hydrolysis Reactor	Packed bed	400 psia, 410°F	1
2	Amine Absorber	Packed bed	400,000 lbm/h	1
3	H_2S Concentrator	Packed bed		1
4	Alumina Reabsorber	Packed bed		1
5	Amine Stripper	Packed bed		1
6	Lean/Rich Exchanger	Shell & tube		1
7	Stripper Reboiler	Shell & tube		1
8	Lean Pump	Horizontal, centrifugal		1
9	Rich Pump	Horizontal, centrifugal		1
10	Sulfur Plant	Claus plant	54 long ton/day	1

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	187 MWe Gas Turbine Generator	Axial flow, single spool based on GE 7FA	900 lbm/sec airflow 2350°F rotor inlet temp.; 15.2:1 pressure ratio	1
3	Enclosure	Sound attenuating	85 dB at 3 ft	1
4	Air Inlet Filter/Silencer	Two-stage	900 lbm/sec airflow 3.0 in. H_2O pressure drop, dirty	1
5	Starting Package	Electric motor, torque converter drive, turning gear	2,000 hp, time from turning gear to full load ~30 minutes	1
6	Mechanical Package	CS oil reservoir & pumps dual vertical cartridge filters air compressor		1
7	Oil Cooler	Air-cooled, fin fan		1
8	Electrical Control Package	Distributed control system	1 sec. update time 8 MHz clock speed	1
9	Generator Glycol Cooler	Air-cooled, fin fan		1
10	Compressor Wash Skid			1
11	Fire Protection Package	Halon		1

ACCOUNT 7 WASTE HEAT BOILER, DUCTING, AND STACK (total for plant)

Equipment No.	Description	Type	Design Condition Drums	Qty
1	Heat Recovery Steam Generator	Drum, multi-pressure, with economizer section and integral deaerator	HP-2,000 psig/ 1,000°F 460,000 lbm/h IP-405 psig/ 1,000°F 362,000 lbm/h LP-50 psig/476°F 825,000 lbm/h	1
2	Stack	Carbon steel plate, type 409 stainless steel liner	213 ft high x 28 ft dia.	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	107 MWe Steam Turbine Generator	TC2F40	1,800 psig 1,000°F/1,000°F	1
2	Bearing Lube Oil Coolers	Plate and frame		2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop		1
4	Control System	Digital electro-hydraulic	1,600 psig	1
5	Generator Coolers	Plate and frame		2
6	Hydrogen Seal Oil System	Closed loop		1
7	Surface Condenser	Single pass, divided waterbox	910,000 lbm/h steam @ 3.0 in. Hga with 74°F water, 20°F temp rise	1
8	Condenser Vacuum Pumps	Rotary, water sealed	2,500/25 scfm (hogging/holding)	2

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	Circ. Water Pumps	Vert. wet pit	40,000 gpm @ 60 ft	2
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	600 MMBtu/h 52°F WB/74°F CWT/ 94° HWT	1

ACCOUNT 10 SLAG RECOVERY AND HANDLING

ACCOUNT 10A GASIFIER SLAG DEWATERING & REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Slag Quench Tank	Water bath	12 tph	2
2	Slag Crusher	Roll	12 tph	2
3	Slag Depressurizer	Proprietary	12 tph	2
4	Slag Handling Tank	Horizontal, weir	6 tph	4
5	Slag Conveyor	Drag chain	6 tph	4
6	Slag Separation Screen	Vibrating	8 tph	2
7	Coarse Slag Conveyor	Belt/bucket	8 tph	2
8	Fine Ash Storage Tank	Vertical	10,000 gallons	2
9	Fine Ash Transfer Pumps	Horizontal/centrifugal	100 gpm	4
10	Storage Bin	Vertical	1,400 tons	3
11	Unloading Equipment	Telescoping chute	25 tph	3

ACCOUNT 10B FINE (SCRUBBER) SLAG DEWATERING & REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Clarifier		1 tph	1
2	Sump Pump	Horizontal/centrifugal	50 gpm	2
3	Vacuum Filter	Drum	1 tph	1
4	Slag Transport Conveyor	Belt	1 tph	1

9.1.9. Case-9 Equipment List

MAJOR EQUIPMENT LIST
ACCOUNT 1 COAL HANDLING

ACCOUNT 1A COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	500 tph	2
3	Conveyor No. 1	54" belt	1,000 tph	1
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor No. 2	54" belt	1,000 tph	1
6	Reclaim Hopper	N/A	50 ton	2
7	Feeder	Vibratory	350 tph	2
8	Conveyor No. 3	48" belt	650 tph	1
9	Crusher Tower	N/A	650 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	650 ton	1
11	Crusher	Granulator reduction	6"x0 - 3"x0	1
12	Crusher	Impactor reduction	3"x0 - 1¼"x0	1
13	As-Fired Coal Sampling System	Swing hammer		2
14	Conveyor No. 4	48" belt	650 tph	1
15	Transfer Tower	N/A	650 tph	1
16	Tripper	N/A	650 tph	1
17	Coal Silo w/ Vent Filter and Slide Gates	N/A	1,800 ton	3
18	Front–End Loader	Rubber tired	1,600 ton	3

ACCOUNT 2 COAL PREPARATION AND FEED

ACCOUNT 2A FUEL SLURRY PREPARATION AND FUEL INJECTION

Equipment No.	Description	Type	Design Condition	Qty
1	Vibratory Feeder		150 tph	3
2	Conveyor No. 1	Belt	3000 tph	1
3	Conveyor No. 2	Belt	300 tph	1
4	Rod Mill Feed Hopper	Vertical, double hopper	350 tons	1
5	Vibratory Feeder		110 tph	2
6	Weight Feeder	Belt	110 tph	2
7	Rod Mill	Rotary	110 tph	2
8	Slurry Water Storage Tank with Agitator	Field erected	100,000 gal	1
9	Slurry Water Pumps	Horizontal, centrifugal	1,200 gpm	2
10	Rod Mill Product Tank with Agitator	Field erected	200,000 gal	1
11	Rod Mill Product Pumps	Horizontal, centrifugal	2,000 gpm	2
12	Slurry Storage Tank with Agitator	Field erected	350,000 gal	1
13	Centrifugal Slurry Pumps	Horizontal, centrifugal	3,000 gpm	2
14	PD Slurry Pumps	Progressing cavity	600 gpm	4
15	Slurry Blending Tank with Agitator	Field erected	100,000 gal	1
16	Slurry Blending Tank Pumps	Horizontal, centrifugal	500 gpm	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A CONDENSATE AND FEEDWATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cond. Storage Tank	Vertical, cylindrical, outdoor	50,000 gal	1
2	Condensate Pumps	Vert. canned	900 gpm @ 400 ft	2
3	Deaerator (integral with HRSG)	Horiz. spray type	820,000 lbm/h 210°F to 240°F	2
4	IP Feed Pump	Horiz. centrifugal single stage	200 gpm/1,000 ft	2
5	HP Feed Pump	Barrel type, multi-staged, centr.	900 gpm @ 5,100 ft & 50 gpm @ 1,700 ft	2

ACCOUNT 3B MISCELLANEOUS EQUIPMENT

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fab., water tube	400 psig, 650°F	1
2	Service Air Compressors	Recip., single stage, double acting, horiz.	100 psig, 450 cfm	2
3	Inst. Air Dryers	Duplex, regenerative	450 cfm	1
4	Service Water Pumps	Horiz. centrifugal, double suction	200 ft, 700 gpm	2
5	Closed Cycle Cooling Heat Exchangers	Plate and frame	50% cap. each	2
6	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	70 ft, 700 gpm	2
7	Fire Service Booster Pump	Two-stage horiz., centrifugal	250 ft, 700 gpm	1
8	Engine-Driven Fire Pump	Vert. turbine, diesel engine	350 ft, 1,000 gpm	1
9	Raw Water Pumps	SS, single suction	60 ft, 1,100 gpm	2
10	Filtered Water Pumps	SS, single suction	160 ft, 700 gpm	2
11	Filtered Water Tank	Vertical, cylindrical	50,000 gal	1
12	Makeup Demineralizer	Anion, cation, and mixed bed	700 gpm	2
13	Liquid Waste Treatment System		10 years, 25-hour storm	1

ACCOUNT 4 GASIFIER AND ACCESSORIES

ACCOUNT 4A GASIFICATION (total for plant)

Equipment No.	Description	Type	Design Condition	Qty
1	Gasifier and associated equipment	Texaco Pressurized entrained bed	2,250 ton/day/ 450 psia	2
2	Syngas Scrubber	Vertical, upflow	550,000 lbm/h	2
3	Low Temperature Gas Cooling	Syngas Coolers	170,000 scfm syngas	2
5	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	170,000 scfm, medium-Btu gas	1

ACCOUNT 4B AIR SEPARATION PLANT (total for plant)

Equipment No.	Description	Type	Design Condition	Qty
1	Air Compressor	Centrifugal, multi-stage	75,000 scfm, 90 psia discharge pressure	2
2	Cold Box	Vendor Design	2,450 ton/day O_2	1
3	Oxygen Compressor	Centrifugal, multi-stage	25,000 scfm, 600 psia discharge pressure	2

ACCOUNT 5 FUEL GAS SHIFT, CLEANUP, AND CO_2 PROCESSING

ACCOUNT 5A WATER-GAS SHIFT, RAW GAS COOLING, HUMIDIFICATION, ACID GAS REMOVAL AND RECOVERY

Equipment No.	Description	Type	Design Condition	Qty
1	High-Temperature Shift Reactor	Fixed bed	1,000 psia, 750ºF	1
2	Low-Temperature Shift Reactor	Fixed bed	1,000 psia, 350ºF	1
3	HP Steam Generator	Shell and tube	50 x 10^6 Btu/h @ 1,000 psia and 700ºF	1
4	IP Steam Generator	Shell and tube	30 x 10^6 Btu/h @ 1,000 psia and 500ºF	1
5	LP Steam Generator	Shell and tube	15 x 10^6 Btu/h @ 1,000 psia and 500ºF	1
6	Saturation Water Economizers	Shell and tube	50 x 10^6 Btu/h @ 1,000 psia and 500ºF	1
7	Raw Gas Coolers	Shell and tube with condensate drain	150 x 10^6 Btu/h	3
8	Raw Gas Knockout Drum	Vertical with mist eliminator	1,000 psia, 130ºF	1
9	Fuel Gas Saturator	Vertical tray tower	20 stages 750 psia, 450ºF	1
10	Saturator Water Pump	Centrifugal	1,500 gpm @ 120 ft	1
11	Fuel Gas Reheater 1	Shell and tube	41 x 10^6 Btu/h @ 690 psia, 550ºF	1
12	Fuel Gas Reheater 2	Shell and tube	39 x 10^6 Btu/h @ 690 psia, 550ºF	1
13	Double-Stage MDEA Unit	Vendor design	150,000 scfm @ 400 psia	1
14	Claus Unit	Vendor design	60 ltpd sulfur product	1
15	Hydrogenation Reactor	Vertical fixed bed	7,000 scfm @ 22 psia	1
16	Contact Cooler	Spray contact, tray wash tower	7,000 scfm @ 21 psia	1
17	TGTU Amine Unit	Proprietary amine absorber/stripper	5,100 scfm @ 20 psia	1

ACCOUNT 5B $C0_2$ COMPRESSION, DRYING, PURIFICATION, AND LIQUEFACTION

Tag No.	Service	Sizing Parameters	MOC
DA	Columns and Towers		
DA-102	CO2 Rectifier	4'/ 12' ID x 30'/ 12' S/S, DP 450 psig	LTCS
E	Heat Transfer Equipment		
EA	Shell & Tube Exchangers		
EA-101	Flue Gas Compressor 1 Stage Trim Cooler	0.35 MMBTU/h, DP S/T, 85 psig/ 125 psig	CS/SS
EA-102	Flue Gas Compressor 2 Stage Trim Cooler	3.3 MMBTU/HR, DP S/T, 175 psig/ 125 psig	CS/SS
EA-103	Flue Gas Compressor 3 Stage Trim Cooler	5.4 MMBTU/HR, DP S/T, 425 psig/ 125 psig	CS/SS
EA-104	CO2 Condenser	79.5 MMBTU/HR, DP S/T, 300 psig/ 425 psig	LTCS/ LTCS
EA-201	Refrig condenser	93.7 MMBTU/HR, DP S/T, 300 psig/ 125 psig	CS/CS
EA-202	Refrig Subcooler	19.8 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/ LTCS
EA-107	CO2 Rectifier Condenser	6.0 MMBTU/HR, DP S/T, 450 psig/ 450 psig	SS/SS
EA -108	Rectifier Ovhd Interchanger	1.45 MMBTU/HR, DP S/T, 450 psig/ 450 psig	LTCS/ SS
EB	Plate Exchangers		
EC	Air Coolers		
EC-101	Flue Gas Compressor 1 Stage Aftercooler	2.5 MMBTU/HR, DP 85 psig	SS
EC-102	Flue Gas Compressor 2 Stage Aftercooler	12.5 MMBTU/HR, DP 175 psig	SS
EC-103	Flue Gas Compressor 3 Stage Aftercooler	21.0 MMBTU/HR, DP 450 psig	SS
FH	Heaters		
FH-101	Dryer Regeneration Gas Heater	Gas fired, 5.75 MMBTU/HR fired duty	
FA	Drums and Vessels		
FA-100	Flue Gas Compressor 1st Stage Suction Drum	5' 6" ID x 10' S/S, DP 75 psig	CS w/ SS liner
FA-101	Flue Gas Compressor 2nd Stage Suction Drum	13' -0" ID x 18' S/S, DP 85 psig	CS w/ SS liner
FA-102	Flue Gas Compressor 3rd Stage Suction Drum	11'-6" ID x 16' S/S DP 175 psig	CS w/ SS liner
FA-103	Flue Gas Compressor Third Stage Discharge K/O Drum	8'- 6" ID x 14' S/S, DP 450 psig	CS w/ SS liner
FA-201	Refrig Surge Drum	12' ID x 30' S/S DP 300 psig	CS
FA-202	Refrig Suction Srubber	18' ID x 26' S/S, DP 300 psig	ITCS
FD	Filters and Dryers		
FD-102	Flue Gas Filter	2665 ACFM, DP 425 psig	CS
FF	Dryers (Dessicant Type)		
FF-101	Flue Gas Dryer	Two Vessels 10' - 0 " ID x 20' S/S DP 450 psig DT 500 F	CS
GA	Pumps Centrifugal		
GA-103	CO2 Pipeline pump	930 gpm, DP 1655 psi	ITCS
GB	Compressors & Blowers		
GB-101	Flue Gas Compressor	See Dresser Rand quote	SS
GB-102	Propane Refrig Compressor	See Dresser Rand quote	ITCS

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	187 MWe Gas Turbine Generator	Axial flow, single spool based on GE 7FA	900 lbm/sec airflow 2350°F rotor inlet temp.; 15.2:1 pressure ratio Modified for Hydrogen Combustion	1
3	Enclosure	Sound attenuating	85 dB at 3 ft	1
4	Air Inlet Filter/Silencer	Two-stage	900 lbm/sec airflow 3.0 in. H_2O pressure drop, dirty	1
5	Starting Package	Electric motor, torque converter drive, turning gear	2,000 hp, time from turning gear to full load ~30 minutes	1
7	Mechanical Package	CS oil reservoir & pumps dual vertical cartridge filters air compressor		1
8	Oil Cooler	Air-cooled, fin fan		1
9	Electrical Control Package	Distributed control system	1 sec. update time 8 MHz clock speed	1
10	Generator Glycol Cooler	Air-cooled, fin fan		1
11	Compressor Wash Skid			1
12	Fire Protection Package	Halon		1

ACCOUNT 7 WASTE HEAT BOILER, DUCTING, AND STACK (total for plant)

Equipment No.	Description	Type	Design Condition Drums	Qty
1	Heat Recovery Steam Generator	Drum, multi-pressure, with economizer section and integral deaerator	HP-2,000 psig/ 1,000°F 460,000 lbm/h IP-405 psig/ 1,000°F 362,000 lbm/h LP-50 psig/476°F 825,000 lbm/h	1
2	Stack	Carbon steel plate, type 409 stainless steel liner	213 ft high x 28 ft dia.	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	106 MWe Steam Turbine Generator	TC2F40	1,800 psig 1,000°F/1,000°F	1
2	Bearing Lube Oil Coolers	Plate and frame		2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop		1
4	Control System	Digital electro-hydraulic	1,600 psig	1
5	Generator Coolers	Plate and frame		2
6	Hydrogen Seal Oil System	Closed loop		1
7	Surface Condenser	Single pass, divided waterbox	910,000 lbm/h steam @ 3.0 in. Hga with 74°F water, 20°F temp rise	1
8	Condenser Vacuum Pumps	Rotary, water sealed	2500/25 scfm (hogging/holding)	2

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	Circ. Water Pumps	Vert. wet pit	40,000 gpm @ 60 ft	2
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	700 MMBtu/h 52°F WB/74°F CWT/ 94° HWT	1

ACCOUNT 10 SLAG RECOVERY AND HANDLING

ACCOUNT 10A GASIFIER SLAG DEWATERING & REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Slag Quench Tank	Water bath	14 tph	2
2	Slag Crusher	Roll	14 tph	2
3	Slag Depressurizer	Proprietary	14 tph	2
4	Slag Handling Tank	Horizontal, weir	6 tph	4
5	Slag Conveyor	Drag chain	6 tph	4
6	Slag Separation Screen	Vibrating	10 tph	2
7	Coarse Slag Conveyor	Belt/bucket	10 tph	2
8	Fine Ash Storage Tank	Vertical	10,000 gallons	2
9	Fine Ash Transfer Pumps	Horizontal/centrifugal	100 gpm	4
10	Storage Bin	Vertical	1,400 tons	3
11	Unloading Equipment	Telescoping chute	25 tph	3

ACCOUNT 10B FINE (SCRUBBER) SLAG DEWATERING & REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Clarifier		1 tph	1
2	Sump Pump	Horizontal/centrifugal	50 gpm	2
3	Vacuum Filter	Drum	1 tph	1
4	Slag Transport Conveyor	Belt	1 tph	1

9.1.10. Case-10 Equipment List

ACCOUNT 1 COAL HANDLING

ACCOUNT 1A COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor No. 1	54" belt	900 tph	1
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor No. 2	54" belt	900 tph	1
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor No. 3	48" belt	600 tph	1
9	Crusher Tower	N/A	600 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	600 ton	1
11	Crusher	Granulator reduction	6"x0 - 3"x0	1
12	Crusher	Impactor reduction	3"x0 - 1¼"x0	1
13	As-Fired Coal Sampling System	Swing hammer		2
14	Conveyor No. 4	48" belt	600 tph	1
15	Transfer Tower	N/A	600 tph	1
16	Tripper	N/A	600 tph	1
17	Coal Silo w/ Vent Filter and Slide Gates	N/A	1,600 ton	3
18	Front–End Loader	Rubber tired	1,600 ton	3

ACCOUNT 2 COAL PREPARATION AND FEED

ACCOUNT 2A FUEL SLURRY PREPARATION AND FUEL INJECTION

Equipment No.	Description	Type	Design Condition	Qty
1	Vibratory Feeder		140 tph	3
2	Conveyor No. 1	Belt	280 tph	1
3	Conveyor No. 2	Belt	280 tph	1
4	Rod Mill Feed Hopper	Vertical, double hopper	300 tons	1
5	Vibratory Feeder		100 tph	2
6	Weight Feeder	Belt	100 tph	2
7	Rod Mill	Rotary	100 tph	2
8	Slurry Water Storage Tank with Agitator	Field erected	100,000 gal	1
9	Slurry Water Pumps	Horizontal, centrifugal	1,200 gpm	2
10	Rod Mill Product Tank with Agitator	Field erected	200,000 gal	1
11	Rod Mill Product Pumps	Horizontal, centrifugal	2,000 gpm	2
12	Slurry Storage Tank with Agitator	Field erected	350,000 gal	1
13	Centrifugal Slurry Pumps	Horizontal, centrifugal	3,000 gpm	2
14	PD Slurry Pumps	Progressing cavity	500 gpm	4
15	Slurry Blending Tank with Agitator	Field erected	100,000 gal	1
16	Slurry Blending Tank Pumps	Horizontal, centrifugal	450 gpm	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A CONDENSATE AND FEEDWATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cond. Storage Tank	Vertical, cylindrical, outdoor	50,000 gal	1
2	Condensate Pumps	Vert. canned	900 gpm @ 400 ft	2
3	Deaerator (integral with HRSG)	Horiz. spray type	820,000 lbm/h 210°F to 240°F	2
4	IP Feed Pump	Horiz. centrifugal single stage	200 gpm/1,000 ft	2
5	HP Feed Pump	Barrel type, multi-staged, centr.	900 gpm @ 5,100 ft & 50 gpm @ 1,700 ft	2

ACCOUNT 3B MISCELLANEOUS EQUIPMENT

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fab., water tube	400 psig, 650°F	1
2	Service Air Compressors	Recip., single stage, double acting, horiz.	100 psig, 450 cfm	2
3	Inst. Air Dryers	Duplex, regenerative	450 cfm	1
4	Service Water Pumps	Horiz. centrifugal, double suction	200 ft, 700 gpm	2
5	Closed Cycle Cooling Heat Exchangers	Plate and frame	50% cap. each	2
6	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	70 ft, 700 gpm	2
7	Fire Service Booster Pump	Two-stage horiz., centrifugal	250 ft, 700 gpm	1
8	Engine-Driven Fire Pump	Vert. turbine, diesel engine	350 ft, 1,000 gpm	1
9	Raw Water Pumps	SS, single suction	60 ft, 1,100 gpm	2
10	Filtered Water Pumps	SS, single suction	160 ft, 700 gpm	2
11	Filtered Water Tank	Vertical, cylindrical	50,000 gal	1
12	Makeup Demineralizer	Anion, cation, and mixed bed	700 gpm	2
13	Liquid Waste Treatment System		10 years, 25-hour storm	1
14	Front-End Loaders	Rubber tired, bucket		3

ACCOUNT 4 GASIFIER AND ACCESSORIES

ACCOUNT 4A GASIFICATION (total for plant)

Equipment No.	Description	Type	Design Condition	Qty
1	Gasifier and associated equipment	Pressurized entrained bed	2,250 ton/day/ 965 psia	1
2	Syngas Scrubber	Vertical, upflow	460,000 lbm/h	2
3	Low Temperature Gas Cooling	Syngas Coolers	300,000 scfm syngas	5
5	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	450,000 lbm/h, medium-Btu gas	1

ACCOUNT 4B AIR SEPARATION PLANT (total for plant)

Equipment No.	Description	Type	Design Condition	Qty
1	Air Compressor	Centrifugal, multi-stage	60,000 scfm, 190 psia discharge pressure	2
2	Cold Box	Vendor Design	2,150 ton/day O_2	1
3	Oxygen Compressor	Centrifugal, multi-stage	20,000 scfm, 1114 psia discharge pressure	2
4	Nitrogen Compressor	Centrifugal, multi-stage	40,000 scfm, 350 psia discharge pressure	1

ACCOUNT 5 SYNGAS CLEANUP

Equipment No.	Description	Type	Design Condition	Qty
1	COS Hydrolysis Reactor	Packed bed	955 psia, 400°F	1
2	Selexol Absorber	Packed bed	400,000 lbm/h	1
3	Selexol H_2S Concentrator	Packed bed		1
4	Selexol Reabsorber	Packed bed		1
5	Selexol Stripper	Packed bed		1
6	Lean/Rich Exchanger	Shell & tube		1
7	Stripper Reboiler	Shell & tube		1
8	Lean Pump	Horizontal, centrifugal		1
9	Rich Pump	Horizontal, centrifugal		1
10	Sulfur Plant	Claus plant	53 long ton/day	1

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	187 MWe Gas Turbine Generator	Axial flow, single spool based on GE 7FA	900 lbm/sec airflow 2350°F rotor inlet temp.; 15.2:1 pressure ratio	1
2	GE Rotoflow Turboexpander	6,650 kW_e	200 lbm/sec syngas flow, 520 F, 2.2 pressure ratio.	1
3	Enclosure	Sound attenuating	85 dB at 3 ft	1
4	Air Inlet Filter/Silencer	Two-stage	900 lbm/sec airflow 3.0 in. H_2O pressure drop, dirty	1
5	Starting Package	Electric motor, torque converter drive, turning gear	2,000 hp, time from turning gear to full load ~30 minutes	1
6	Air to Air Cooler			1
7	Mechanical Package	CS oil reservoir & pumps dual vertical cartridge filters air compressor		1
8	Oil Cooler	Air-cooled, fin fan		1
9	Electrical Control Package	Distributed control system	1 sec. update time 8 MHz clock speed	1
10	Generator Glycol Cooler	Air-cooled, fin fan		1
11	Compressor Wash Skid			1
12	Fire Protection Package	Halon		1

ACCOUNT 7 WASTE HEAT BOILER, DUCTING, AND STACK (total for plant)

Equipment No.	Description	Type	Design Condition Drums	Qty
1	Heat Recovery Steam Generator	Drum, multi-pressure, with economizer section and integral deaerator	HP-1,800 psig/ 1,000°F 460,000 lbm/h IP-405 psig/ 1,000°F 362,000 lbm/h LP-50 psig/476°F 825,000 lbm/h	1
2	Stack	Carbon steel plate, type 409 stainless steel liner	213 ft high x 28 ft dia.	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	92 MWe Steam Turbine Generator	TC2F40	1,800 psig 1,000°F/1,000°F	1
2	Bearing Lube Oil Coolers	Plate and frame		2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop		1
4	Control System	Digital electro-hydraulic	1,600 psig	1
5	Generator Coolers	Plate and frame		2
6	Hydrogen Seal Oil System	Closed loop		1
7	Surface Condenser	Single pass, divided waterbox	910,000 lbm/h steam @ 2.0 in. Hga with 74°F water, 20°F temp rise	1
8	Condenser Vacuum Pumps	Rotary, water sealed	2,500/25 scfm (hogging/holding)	2

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	Circ. Water Pumps	Vert. wet pit	40,000 gpm @ 60 ft	2
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	810 MMBtu/h 52°F WB/74°F CWT/ 94° HWT	1

ACCOUNT 10 SLAG RECOVERY AND HANDLING

ACCOUNT 10A GASIFIER SLAG DEWATERING & REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Slag Quench Tank	Water bath	12 tph	1
2	Slag Crusher	Roll	12 tph	1
3	Slag Depressurizer	Proprietary	12 tph	1
4	Slag Handling Tank	Horizontal, weir	6 tph	2
5	Slag Conveyor	Drag chain	6 tph	2
6	Slag Separation Screen	Vibrating	8 tph	1
7	Coarse Slag Conveyor	Belt/bucket	8 tph	1
8	Fine Ash Storage Tank	Vertical	10,000 gallons	1
9	Fine Ash Transfer Pumps	Horizontal/centrifugal	100 gpm	2
10	Storage Bin	Vertical	1,400 tons	1
11	Unloading Equipment	Telescoping chute	25 tph	1

ACCOUNT 10B FINE (SCRUBBER) SLAG DEWATERING & REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Clarifier		1 tph	1
2	Sump Pump	Horizontal/centrifugal	50 gpm	2
3	Vacuum Filter	Drum	1 tph	1
4	Slag Transport Conveyor	Belt	1 tph	1

9.1.11. Case-11 Equipment List

ACCOUNT 1 COAL HANDLING

ACCOUNT 1A COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor No. 1	54" belt	900 tph	1
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor No. 2	54" belt	900 tph	1
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor No. 3	48" belt	600 tph	1
9	Crusher Tower	N/A	600 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	600 ton	1
11	Crusher	Granulator reduction	6"x0 - 3"x0	1
12	Crusher	Impactor reduction	3"x0 - 1¼"x0	1
13	As-Fired Coal Sampling System	Swing hammer		2
14	Conveyor No. 4	48" belt	600 tph	1
15	Transfer Tower	N/A	600 tph	1
16	Tripper	N/A	600 tph	1
17	Coal Silo w/ Vent Filter and Slide Gates	N/A	1,600 ton	3
18	Front–End Loader	Rubber tired	1,600 ton	3

ACCOUNT 2 COAL PREPARATION AND FEED

ACCOUNT 2A FUEL SLURRY PREPARATION AND FUEL INJECTION

Equipment No.	Description	Type	Design Condition	Qty
1	Vibratory Feeder		140 tph	3
2	Conveyor No. 1	Belt	280 tph	1
3	Conveyor No. 2	Belt	280 tph	1
4	Rod Mill Feed Hopper	Vertical, double hopper	300 tons	1
5	Vibratory Feeder		100 tph	2
6	Weight Feeder	Belt	100 tph	2
7	Rod Mill	Rotary	100 tph	2
8	Slurry Water Storage Tank with Agitator	Field erected	100,000 gal	1
9	Slurry Water Pumps	Horizontal, centrifugal	1,200 gpm	2
10	Rod Mill Product Tank with Agitator	Field erected	200,000 gal	1
11	Rod Mill Product Pumps	Horizontal, centrifugal	2,000 gpm	2
12	Slurry Storage Tank with Agitator	Field erected	350,000 gal	1
13	Centrifugal Slurry Pumps	Horizontal, centrifugal	3,000 gpm	2
14	PD Slurry Pumps	Progressing cavity	500 gpm	4
15	Slurry Blending Tank with Agitator	Field erected	100,000 gal	1
16	Slurry Blending Tank Pumps	Horizontal, centrifugal	450 gpm	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A CONDENSATE AND FEEDWATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cond. Storage Tank	Vertical, cylindrical, outdoor	50,000 gal	1
2	Condensate Pumps	Vert. canned	900 gpm @ 400 ft	2
3	Deaerator (integral with HRSG)	Horiz. spray type	820,000 lbm/h 210°F to 240°F	2
4	IP Feed Pump	Horiz. centrifugal single stage	200 gpm/1,000 ft	2
5	HP Feed Pump	Barrel type, multi-staged, centr.	900 gpm @ 5,100 ft & 50 gpm @ 1,700 ft	2

ACCOUNT 3B MISCELLANEOUS EQUIPMENT

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fab., water tube	400 psig, 650°F	1
2	Service Air Compressors	Recip., single stage, double acting, horiz.	100 psig, 450 cfm	2
3	Inst. Air Dryers	Duplex, regenerative	450 cfm	1
4	Service Water Pumps	Horiz. centrifugal, double suction	200 ft, 700 gpm	2
5	Closed Cycle Cooling Heat Exchangers	Plate and frame	50% cap. each	2
6	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	70 ft, 700 gpm	2
7	Fire Service Booster Pump	Two-stage horiz., centrifugal	250 ft, 700 gpm	1
8	Engine-Driven Fire Pump	Vert. turbine, diesel engine	350 ft, 1,000 gpm	1
9	Raw Water Pumps	SS, single suction	60 ft, 1,100 gpm	2
10	Filtered Water Pumps	SS, single suction	160 ft, 700 gpm	2
11	Filtered Water Tank	Vertical, cylindrical	50,000 gal	1
12	Makeup Demineralizer	Anion, cation, and mixed bed	700 gpm	2
13	Liquid Waste Treatment System		10 years, 25-hour storm	1
14	Front-End Loaders	Rubber tired, bucket		3

ACCOUNT 4 GASIFIER AND ACCESSORIES

ACCOUNT 4A GASIFICATION (total for plant)

Equipment No.	Description	Type	Design Condition	Qty
1	Gasifier and associated equipment	Pressurized entrained bed	2,250 ton/day/ 965 psia	2
2	Syngas Scrubber	Vertical, upflow	500,000 lbm/h	2
3	Low Temperature Gas Cooling	Syngas Coolers	150,000 scfm syngas	2
5	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	150,000 scfm, medium-Btu gas	1

ACCOUNT 4B AIR SEPARATION PLANT (total for plant)

Equipment No.	Description	Type	Design Condition	Qty
1	Air Compressor	Centrifugal, multi-stage	65,000 scfm, 190 psia discharge pressure	2
2	Cold Box	Vendor Design	2,300 ton/day O_2	1
3	Oxygen Compressor	Centrifugal, multi-stage	22,000 scfm, 1114 psia discharge pressure	2
4	Nitrogen Compressor	Centrifugal, multi-stage	45,000 scfm, 350 psia discharge pressure	2

ACCOUNT 5 FUEL GAS SHIFT, CLEANUP, AND CO_2 PROCESSING

ACCOUNT 5A WATER-GAS SHIFT, RAW GAS COOLING, HUMIDIFICATION, ACID GAS REMOVAL AND RECOVERY

Equipment No.	Description	Type	Design Condition	Qty
1	High-Temperature Shift Reactor	Fixed bed	1,000 psia, 750ºF	1
2	Low-Temperature Shift Reactor	Fixed bed	1,000 psia, 350ºF	1
3	HP Steam Generator	Shell and tube	50 x 10^6 Btu/h @ 1,000 psia and 700ºF	1
4	IP Steam Generator	Shell and tube	30 x 10^6 Btu/h @ 1,000 psia and 500ºF	1
5	LP Steam Generator	Shell and tube	15 x 10^6 Btu/h @ 1,000 psia and 500ºF	1
6	Saturation Water Economizers	Shell and tube	50 x 10^6 Btu/h @ 1,000 psia and 500ºF	1
7	Raw Gas Coolers	Shell and tube with condensate drain	150 x 10^6 Btu/h	3
8	Raw Gas Knockout Drum	Vertical with mist eliminator	1,000 psia, 130ºF	1
9	Fuel Gas Saturator	Vertical tray tower	20 stages 750 psia, 450ºF	1
10	Saturator Water Pump	Centrifugal	1,500 gpm @ 120 ft	1
11	Fuel Gas Reheater 1	Shell and tube	41 x 10^6 Btu/h @ 690 psia, 550ºF	1
12	Fuel Gas Expander	Axial	PR=1.8 @ 940 psia	1
13	Fuel Gas Reheater 2	Shell and tube	39 x 10^6 Btu/h @ 690 psia, 550ºF	1
14	Double-Stage Selexol Unit	Vendor design	130,000 scfm @ 1,000 psia	1
15	Claus Unit	Vendor design	65 ltpd sulfur product	1
16	Hydrogenation Reactor	Vertical fixed bed	7,000 scfm @ 22 psia	1
17	Contact Cooler	Spray contact, tray wash tower	7,000 scfm @ 21 psia	1
18	TGTU Amine Unit	Proprietary amine absorber/stripper	5,100 scfm @ 20 psia	1

Equipment No.	Description	Type	Design Condition	Qty
19	Tail Gas Recycle Compressor	Centrifugal, multi-staged	3,610 scfm, PR = 58	1

ACCOUNT 5B CO_2 COMPRESSION, DRYING, PURIFICATION, AND LIQUEFACTION

Tag No.	Service	Sizing Parameters	MOC
DA	Columns and Towers		
DA-102	CO2 Rectifier	4'/ 12' ID x 30'/ 12' S/S, DP 450 psig	LTCS
E	Heat Transfer Equipment		
EA	Shell & Tube Exchangers		
EA-101	Flue Gas Compressor 1 Stage Trim Cooler	0.33 MMBTU/h, DP S/T, 85 psig/ 125 psig	CS/SS
EA-102	Flue Gas Compressor 2 Stage Trim Cooler	3.1 MMBTU/HR, DP S/T, 175 psig/ 125 psig	CS/SS
EA-103	Flue Gas Compressor 3 Stage Trim Cooler	5.1 MMBTU/HR, DP S/T, 425 psig/ 125 psig	CS/SS
EA-104	CO2 Condenser	75.5 MMBTU/HR, DP S/T, 300 psig/ 425 psig	LTCS/ LTCS
EA-201	Refrig condenser	89.0 MMBTU/HR, DP S/T, 300 psig/ 125 psig	CS/CS
EA-202	Refrig Subcooler	18.8 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/ LTCS
EA-107	CO2 Rectifier Condenser	5.7 MMBTU/HR, DP S/T, 450 psig/ 450 psig	SS/SS
EA -108	Rectifier Ovhd Interchanger	1.38 MMBTU/HR, DP S/T, 450 psig/ 450 psig	LTCS/ SS
EB	Plate Exchangers		
EC	Air Coolers		
EC-101	Flue Gas Compressor 1 Stage Aftercooler	2.4 MMBTU/HR, DP 85 psig	SS
EC-102	Flue Gas Compressor 2 Stage Aftercooler	11.9 MMBTU/HR, DP 175 psig	SS
EC-103	Flue Gas Compressor 3 Stage Aftercooler	20.0 MMBTU/HR, DP 450 psig	SS
FH	Heaters		
FH-101	Dryer Regeneration Gas Heater	Gas fired, 5.46 MMBTU/HR fired duty	
FA	Drums and Vessels		
FA-100	Flue Gas Compressor 1st Stage Suction Drum	5' 6" ID x 10' S/S, DP 75 psig	CS w/ SS liner
FA-101	Flue Gas Compressor 2nd Stage Suction Drum	13' -0" ID x 18' S/S, DP 85 psig	CS w/ SS liner
FA-102	Flue Gas Compressor 3rd Stage Suction Drum	11'-6" ID x 16' S/S DP 175 psig	CS w/ SS liner
FA-103	Flue Gas Compressor Third Stage Discharge K/O Drum	8'- 6" ID x 14' S/S, DP 450 psig	CS w/ SS liner
FA-201	Refrig Surge Drum	12' ID x 30' S/S DP 300 psig	CS
FA-202	Refrig Suction Srubber	18' ID x 26' S/S, DP 300 psig	ITCS
FD	Filters and Dryers		
FD-102	Flue Gas Filter	2530 ACFM, DP 425 psig	CS
FF	Dryers (Dessicant Type)		
FF-101	Flue Gas Dryer	Two Vessels 10' - 0 " ID x 20' S/S DP 450 psig DT 500 F	CS
GA	Pumps Centrifugal		
GA-103	CO2 Pipeline pump	880 gpm, DP 1655 psi	ITCS
GB	Compressors & Blowers		
GB-101	Flue Gas Compressor	See Dresser Rand quote	SS
GB-102	Propane Refrig Compressor	See Dresser Rand quote	ITCS

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	187 MWe Gas Turbine Generator	Axial flow, single spool based on GE 7FA	900 lbm/sec airflow 2350°F rotor inlet temp.; 15.2:1 pressure ratio Modified for Hydrogen Combustion	1
2	GE Rotoflow Turboexpander	6,570 kW_e	50 lbm/sec syngas flow, 520 F, 2.2 pressure ratio.	1
3	Enclosure	Sound attenuating	85 dB at 3 ft	1
4	Air Inlet Filter/Silencer	Two-stage	900 lbm/sec airflow 3.0 in. H_2O pressure drop, dirty	1
5	Starting Package	Electric motor, torque converter drive, turning gear	2,000 hp, time from turning gear to full load ~30 minutes	1
6	Air to Air Cooler			1
7	Mechanical Package	CS oil reservoir & pumps dual vertical cartridge filters air compressor		1
8	Oil Cooler	Air-cooled, fin fan		1
9	Electrical Control Package	Distributed control system	1 sec. update time 8 MHz clock speed	1
10	Generator Glycol Cooler	Air-cooled, fin fan		1
11	Compressor Wash Skid			1
12	Fire Protection Package	Halon		1

ACCOUNT 7 WASTE HEAT BOILER, DUCTING, AND STACK (total for plant)

Equipment No.	Description	Type	Design Condition Drums	Qty
1	Heat Recovery Steam Generator	Drum, multi-pressure, with economizer section and integral deaerator	HP-1,800 psig/ 1,000°F IP-405 psig/ 1,000°F LP-50 psig/476°F	1
2	Stack	Carbon steel plate, type 409 stainless steel liner	213 ft high x 28 ft dia.	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	91 MWe Steam Turbine Generator	TC2F40	1,800 psig 1,000°F/1,000°F	1
2	Bearing Lube Oil Coolers	Plate and frame		2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop		1
4	Control System	Digital electro-hydraulic	1,600 psig	1
5	Generator Coolers	Plate and frame		2
6	Hydrogen Seal Oil System	Closed loop		1
7	Surface Condenser	Single pass, divided waterbox	910,000 lbm/h steam @ 2.0 in. Hga with 74°F water, 20°F temp rise	1
8	Condenser Vacuum Pumps	Rotary, water sealed	2500/25 scfm (hogging/holding)	2

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	Circ. Water Pumps	Vert. wet pit	40,000 gpm @ 60 ft	2
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	1,000 MMBtu/h 52°F WB/74°F CWT/ 94° HWT	1

ACCOUNT 10 SLAG RECOVERY AND HANDLING

ACCOUNT 10A GASIFIER SLAG DEWATERING & REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Slag Quench Tank	Water bath	8 tph	2
2	Slag Crusher	Roll	8 tph	2
3	Slag Depressurizer	Proprietary	8 tph	2
4	Slag Handling Tank	Horizontal, weir	6 tph	2
5	Slag Conveyor	Drag chain	6 tph	2
6	Slag Separation Screen	Vibrating	8 tph	1
7	Coarse Slag Conveyor	Belt/bucket	8 tph	1
8	Fine Ash Storage Tank	Vertical	10,000 gallons	1
9	Fine Ash Transfer Pumps	Horizontal/centrifugal	100 gpm	2
10	Storage Bin	Vertical	1,400 tons	1
11	Unloading Equipment	Telescoping chute	25 tph	1

ACCOUNT 10B FINE (SCRUBBER) SLAG DEWATERING & REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Clarifier		1 tph	1
2	Sump Pump	Horizontal/centrifugal	50 gpm	2
3	Vacuum Filter	Drum	1 tph	1
4	Slag Transport Conveyor	Belt	1 tph	1

9.1.12. Case-12 Equipment List

9.1.12.1. Case-12 Gasifier Island Equipment

Fuel Feeding System:
- Day Silos (3)
- Fuel Silo Isolation Valves
- Fuel Feeders
- Feeder Isolation Valves
- Conveyance to Furnace
Limestone Feeding System:
- Day Silo (1)
- Limestone Silo Isolation Valves
- Rotary Feeder
- Blower
- Conveyance from Blower to Fuel Feeders
Furnace Loop Equipment:
-Combustor (Oxidizer)
-Reductor
-Ring Cone Separators and Sealpots for Combustor and Reductor
-Ash Drain Valve(s)
-Start-up Burner System (Including Burners, Piping, Ducts, and Local Control Equipment) for Reductor
- Metal/Fabric Expansion Joints
MBHE #1 & #2 Equipment:
- Connecting Tubes/Pipes
- Drum Including Internals, Nozzles, and Hanger Rods
–Downcomer System
MBHE #1 & #2 Heat Absorbing Surface:
Horizontal Superheaters
Horizontal Reheater
Horizontal Evaporator
Horizontal Economizer
- Superheater/Reheater Desuperheaters
- Desuperheater Block Valves
- Desuperheater Piping
- Economizer Piping to Drum
- Superheater Interconnecting Piping
- Feedwater Stop, Feedwater Check valves
- Safety Valves/Discharge Piping/Silencers
- Circulation Pumps (3)
- Trim Valves and Piping
- Drum Level Gauge and Indicators
Solids Ducts:
- Transport Duct from Combustor Sealpot to Reductor
- Transport Duct form Reducer to Ring Cone Separator to MBHE Sealpot
- Return Duct form Combustor Sealpot to Reducer
- Bypass Duct from Combustor Sealpot to Combustor
- ash Transport Duct from MBHE to Combustor

Gas Ducts:
- Flue Gas Duct from Combustor to Air Heater
- Flue Gas Duct from Air Heater to ID Fan
- Flue Gas Duct from ID Fan to Stack (by Others)
- Medium Btu Gas Duct from Reductor to Power Generation System
Air Ducts:
- Reductor Startup Air Duct
- Hot Air from Air Heater to Combustor
- PA Fan w/Drive (by Others)
- I.D. Fan w/Drive (by others)
- Expansion Joints
Ash Handling System:
- Bed Ash Drains and Ash Screw Coolers
Structural:
- Structural Steel Including Platforms, Walkways, Stairways, and Ladders
- Gasifier Island Elevator
Instrumentation and Controls:
- Burner Management (FBSS) Logic
- Field Instruments
- Controller Drives
Refractories:
- Material for All Internal Refractory Linings for Furnished Process and Gasifier Equipment
Insulation and Lagging:
- Material for Insulation and Lagging for Heat Conservation and Personnel Protection for Furnished Equipment, Materials Testing and Installation
Painting:
- Shop Prime Paint Coating for Seller Furnished Equipment
Miscellaneous:
- Operator Training Program
- Maintenance Training Program
- Instruction Manuals
- Spare Parts for Commissioning
- Technical Representation During Start-up and Testing
- Field Erection of Equipment Scope
- Freight to Site

9.1.12.2. Case-12 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-12 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor 1	54" belt	450 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	2
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	250 tph	2
3	Conveyor 4	48" belt	250 tph	1
4	Transfer Tower	N/A	250 tph	1
5	Tripper	N/A	250 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	700 ton	2
7	Feeder	Gravimetric	150 tph	2

ACCOUNT 2B LIMESTONE FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Storage Silo	Cylindrical	1,500 Ton	1
2	Weigh Feeder	Gravimetric	25 tph	1
3	Bin Activator		25 tph	
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Cond. Storage Tank	Vertical, cylindrical, outdoor	50,000 gal	1
2	Condensate Pumps	Vert. canned	900 gpm @ 400 ft	2
3	Deaerator (integral with HRSG)	Horiz. spray type	820,000 lbm/h 210°F to 240°F	2
4	IP Feed Pump	Horiz. centrifugal single stage	200 gpm/1,000 ft	2
5	HP Feed Pump	Barrel type, multi-staged, centr.	900 gpm @ 5,100 ft & 50 gpm @ 1,700 ft	2

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 2,500 gpm	2

14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	100,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 Gasifier and Auxiliaries

Equipment No.	Description	Type	Design Condition	Qty
1	Chemical Looping Gasifier System	Advanced	Alstom Design	1
2	Primary Air Fan	Centrifugal	644,138 pph, 140,000 acfm, 47" wg, 1,600 hp	1
3	Fuel Gas Compression System	Intercooled Recip.	327,422 pph 340,000 acfm 14.7 psia suction 300 psia outlet 40,000 hp	1
4	ID Fan	Centrifugal	531,829 pph, 150,000 acfm, 40" wg 1,000 hp	1

ACCOUNT 5 FLUE GAS CLEANUP
Included in Alstom Gasifier System

ACCOUNT 5A PARTICULATE CONTROL
Included in Alstom Gasifier System

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Equipment No	Description	Type	Design Condition	Qty
1	Gas Turbine	GE 7FA	197,000 kW	1

ACCOUNT 7 DUCTING AND STACK

Equipment No.	Description	Type	Design Condition	Qty
1	Boiler Stack	Concrete with Steel liner	160°F 500,000 acfm	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	105 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	750 MMBtu/h 100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	50,000 gpm @ 60 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

Included in Alstom Gasifier System

ACCOUNT 10B ASH HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Air Blower		1,800 cfm	1
2	Bottom Drain Silo	Reinforced concrete	1,500 tons	1
3	Slide Gate Valves			2
4	Wet Unloader		30 tph	1
5	Telescoping Unloading Chute			1

9.1.13. Case-13 Equipment List

9.1.13.1. Case-13 Gasifier Island Equipment

Fuel Feeding System:
- Day Silos (3)
- Fuel Silo Isolation Valves
- Fuel Feeders
- Feeder Isolation Valves
- Conveyance to Furnace
Limestone Feeding System:
- Day Silo (1)
- Limestone Silo Isolation Valves
- Rotary Feeder
- Blower
- Conveyance from Blower to Fuel Feeders
Furnace Loop Equipment:
-Combustor (Oxidizer)
-Reductor
-Calciner
-Ring Cone Separators and Sealpots for Combustor, Reductor and Calciner
-Ash Drain Valve(s)
-Start-up Burner System (Including Burners, Piping, Ducts, and Local Control Equipment) for Reductor
- Metal/Fabric Expansion Joints
MBHE #1 & #2 Equipment:
- Connecting Tubes/Pipes
- Drum Including Internals, Nozzles, and Hanger Rods
–Downcomer System
MBHE #1 & #2 Heat Absorbing Surface:
Horizontal Superheaters
Horizontal Reheater
Horizontal Evaporator
Horizontal Economizer
- Superheater/Reheater Desuperheaters
- Desuperheater Block Valves
- Desuperheater Piping
- Economizer Piping to Drum
- Superheater Interconnecting Piping
- Feedwater Stop, Feedwater Check valves
- Safety Valves/Discharge Piping/Silencers
- Circulation Pumps (3)
- Trim Valves and Piping
- Drum Level Gauge and Indicators
Solids Ducts:
- Transport Duct from Combustor Sealpot to Reducer
- Bypass Duct from Combustor Sealpot to Combustor
- Transport Duct from Reducer Sealpot to MBHE
- Bypass Duct from Reducer Sealpot to Reducer
- Transport Duct from Reducer Sealpot to Calciner
- Transport Duct from MBHE to Reducer

- Return Duct form Calciner Sealpot to MBHE
- Transport Duct from Combustor to Calciner
- Return Duct from Calciner to Combustor

Gas Ducts:
- Flue Gas Duct from Combustor Separator to Air Heater
- Flue Gas Duct from Air Heater to ID Fan
- Flue Gas Duct from ID Fan to Stack (by Others)
- CO_2 Duct from Calciner Separator to Air Heater
- CO_2 Duct from Air Heater to GPS (by Others)
- Medium Btu Gas Duct from Reductor to Power Generation System
Air Ducts:
- Reductor Startup Air Duct
- Hot Air from Air Heater to Combustor
- Duct from PA Fan to Air Heater
- PA Fan w/Drive (by Others)
- I.D. Fan w/Drive (by others)
- Expansion Joints
Ash Handling System:
- Bed Ash Drains and Ash Screw Coolers
Structural:
- Structural Steel Including Platforms, Walkways, Stairways, and Ladders
- Gasifier Island Elevator
Instrumentation and Controls:
- Burner Management (FBSS) Logic
- Field Instruments
- Controller Drives
Refractories:
- Material for All Internal Refractory Linings for Furnished Process and Gasifier Equipment
Insulation and Lagging:
- Material for Insulation and Lagging for Heat Conservation and Personnel Protection for Furnished Equipment, Materials Testing and Installation
Painting:
- Shop Prime Paint Coating for Seller Furnished Equipment
Miscellaneous:
- Operator Training Program
- Maintenance Training Program
- Instruction Manuals
- Spare Parts for Commissioning
- Technical Representation During Start-up and Testing
- Field Erection of Equipment Scope
- Freight to Site

9.1.13.2. Case-13 Gas Processing System Equipment

Number of Trains	Tag no.	Description	Size Parameters	Material
	DA-101	Direct Contact Flue Gas Cooler	18' 6" ID x 34' S/S, DP 10 psig, 3 psi vacuum	CS w/ SS liner
1	EA-2301	CO2 Compr. 1st Stage Aftercooler	8.5 MMBTU/HR, DP S/T, 75 psig/ 125 psig	SS/CS
1	EA-2302	CO2 Compr. 2nd Stage Aftercooler	5.0 MMBTU/HR, DP S/T, 150 psig/ 125 psig	SS/CS
1	EA-2303	CO2 Compr. 3rd Stage Aftercooler	4.5 MMBTU/HR, DP S/T, 350 psig/ 125 psig	CS/CS
1	EA-2304	CO2 Condenser	80.9 MMBTU/hr DP S/T, 300 psig/ 350 psig	CS/CS
1	EA-2401	Propane Refrig Condenser	95.3 MMBTU/HR, DP S/T, 300 psig/ 125 psig	CS/CS
1	EA-2402	Propane Refrig Subcooler	25.44 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
1	EC-2301	CO2 compressor 1st stage air cooler	25.4 MMBTU/HR, DP 75 psig	SS
1	EC-2302	CO2 compressor 2nd stage air cooler	21.3 MMBTU/HR, DP 150 psig	SS
1	EC-2303	CO2 compressor 3rd stage air cooler	19.4 MMBTU/HR, DP 350 psig	SS
1	EB-101	Water Cooler	Total 121 MMBTU/HR, DP P/U, 65 psig/ 125 psig	CS
1	FA-2300	CO_2 Compressor 1st Stage Suction Drum	12'-0" ID x 20' S/S, DP 75 psig	CS/SS
1	FA-2301	CO_2 Compressor 2st Stage Suction Drum	12'- 0" ID x 18' S/S, DP 75 psig	CS/SS
1	FA-2302	CO_2 Compressor 3rd Stage Suction Drum	11'- 6" ID x 14' S/S, DP 150 psig	CS/SS
1	FA-2303	Liquid CO2 Surge Drum	10'- 0" ID x 25' S/S, DP 350 psig	KCS
1	FA-2304	CO2 Compressor 3rd stage Discharge KO Drum	11' 6" ID x 12' S/S, DP 350 psig	CS/SS
1	FA-2401	Propane Refrig Surge Drum	11' ID x 20' S/S, DP 300 psig	CS
1	FA-2402	Propane Refrig Suction Scubber	20' ID x 20' S/S, DP 300 psig	LTCS
1	FA-2403	Propane Refrigeration Economizer	12' ID x 22' S/S DP 300 psig	CS
1	FD-101	Water Filter	6 units, 1020 gpm each, DP 100 psig	SS
1	GA-2301	CO2 Pipeline Pump	980 gpm, DP 1725 psi, 1300 hp	LTCS/CS
1	GA-101	Water Pump	6100 gpm, DP 40 psi	CI w/ SS impeller
1	GB-2301	CO2 Compressor (Motor driven)	25650 hp	SS wheels
1	GB-2401	Propane Refrig Compressor	15600 hp	LTCS
1	PA-2351	CO2 Dryer Package	4 110" x 15' driers DP 350 psig, 430 hp sundyne compressor, 7.1 MMBTU/h heater, 7.8 MMBTU/h air cooler, 37" x 8' KO drum, Dust filter 3150 ACFM	CS w/ SS liner
1		Crane for Compr. Bldg		
		Flue gas ducting		

9.1.13.3. Case-13 Balance of Plant Equipment

This section contains the balance of plant equipment list corresponding to the Case-13 power plant configuration. This list, along with the material and energy balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	200 ton	2
2	Feeder	Vibratory	450 tph	2
3	Conveyor 1	54" belt	450 tph	2
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	2
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	300 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/ Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6" x 0 - 3" x 0	1

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

ACCOUNT 2A COAL PREPARATION AND FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Crusher	Impactor reduction	3" x 0 – 1/4" x 0	1
2	As-Fired Coal Sampling System	Swing hammer	250 tph	2
3	Conveyor 4	48" belt	250 tph	1
4	Transfer Tower	N/A	250 tph	1
5	Tripper	N/A	250 tph	1
6	Coal Silo w/ Vent Filter and Slide Gates	N/A	700 ton	2
7	Feeder	Gravimetric	150 tph	2

ACCOUNT 2B LIMESTONE FEED SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Storage Silo	Cylindrical	1,500 Ton	1
2	Weigh Feeder	Gravimetric	25 tph	1
3	Bin Activator		25 tph	
4	Blowers	Roots	Site	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

ACCOUNT 3A FEEDWATER SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Cond. Storage Tank	Vertical, cylindrical, outdoor	50,000 gal	1
2	Condensate Pumps	Vert. canned	900 gpm @ 400 ft	2
3	Deaerator (integral with HRSG)	Horiz. spray type	820,000 lbm/h 210°F to 240°F	2
4	IP Feed Pump	Horiz. centrifugal single stage	200 gpm/1,000 ft	2
5	HP Feed Pump	Barrel type, multi-staged, centr.	900 gpm @ 5,100 ft & 50 gpm @ 1,700 ft	2

ACCOUNT 3B MISCELLANEOUS SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated water tube	400 psig, 650°F	1
2	Fuel Oil Storage Tank	Vertical, cylindrical	200,000 gal	1
3	Fuel Oil Unloading Pump	Gear	150 ft, 800 gpm	1
4	Fuel Oil Supply Pump	Gear	400 ft, 80 gpm	2
5	Service Air Compressors	SS, double acting	100 psig, 800 scfm	3
6	Inst. Air Dryers	Duplex, regenerative	400 scfm	1
7	Service Water Pumps	SS, double suction	100 ft, 6,000 gpm	2
8	Closed Cycle Cooling Heat Exch.	Shell and tube	50% cap. each	2
9	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	185 ft, 600 gpm	2
11	Fire Service Booster Pump	Two-stage centrifugal	250 ft, 700 gpm	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
13	Raw Water Pumps	SS, single suction	100 ft, 2,500 gpm	2

14	Filtered Water Pumps	SS, single suction	200 ft, 200 gpm	2
15	Filtered Water Tank	Vertical, cylindrical	100,000 gal	1
16	Makeup Demineralizer	Anion, cation, and mixed bed	150 gpm	2
17	Liquid Waste Treatment System	-	Site	1

ACCOUNT 4 Gasifier and Auxiliaries

Equipment No.	Description	Type	Design Condition	Qty
1	Hot Solids Gasifier System	Advanced	Alstom Design	1
2	Primary Air Fan	Centrifugal	700,000 pph, 150,000 acfm, 47" wg, 1,700 hp	1
3	Fuel Gas Compression System	Intercooled Recip.	42,000 pph 24,000 acfm 80 psia suction 300 psia outlet 1,700 hp	1
4	ID Fan	Centrifugal	600,000 pph, 165,000 acfm, 40" wg 1,100 hp	1

ACCOUNT 5 FLUE GAS CLEANUP
Included in Alstom Gasifier System

ACCOUNT 5A PARTICULATE CONTROL
Included in Alstom Gasifier System

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Equipment No	Description	Type	Design Condition	Qty
1	Gas Turbine	GE 7FA	197,000 kW	1

ACCOUNT 7 DUCTING AND STACK

Equipment No.	Description	Type	Design Condition	Qty
1	Boiler Stack	Concrete with Steel liner	160°F 500,000 acfm	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Qty
1	102 MW Turbine Generator	TC2F30	1,800 psig/1,000°F/ 1,000°F/	1
2	Bearing Lube Oil Coolers	Plate and frame	-	2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop	-	1
4	Control System	Electro-hydraulic	1,600 psig	1
5	Generator Coolers	Shell and tube	-	2
6	Hydrogen Seal Oil System	Closed loop	-	1
7	Generator Exciter	Solid state brushless	-	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Mechanical draft	750 MMBtu/h 100,000 gpm 95°F to 75°F	1
2	Circ. W. Pumps	Vertical wet pit	50,000 gpm @ 60 ft	2

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A BOTTOM ASH HANDLING

Included in Alstom Gasifier System

ACCOUNT 10B ASH HANDLING

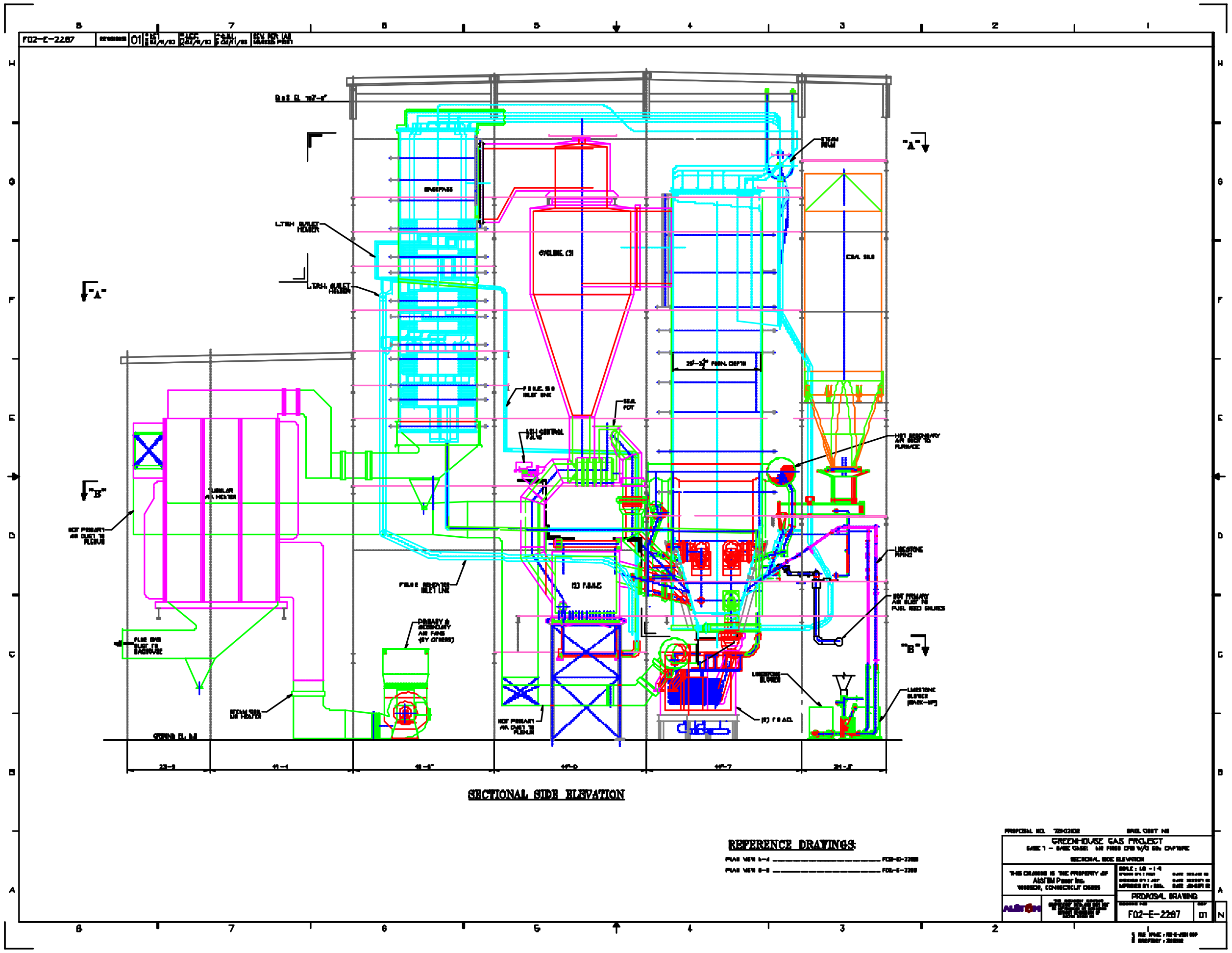
Equipment No.	Description	Type	Design Condition	Qty
1	Air Blower		1,800 cfm	1
2	Bottom Drain Silo	Reinforced concrete	1,500 tons	1
3	Slide Gate Valves			2
4	Wet Unloader		30 tph	1
5	Telescoping Unloading Chute			1

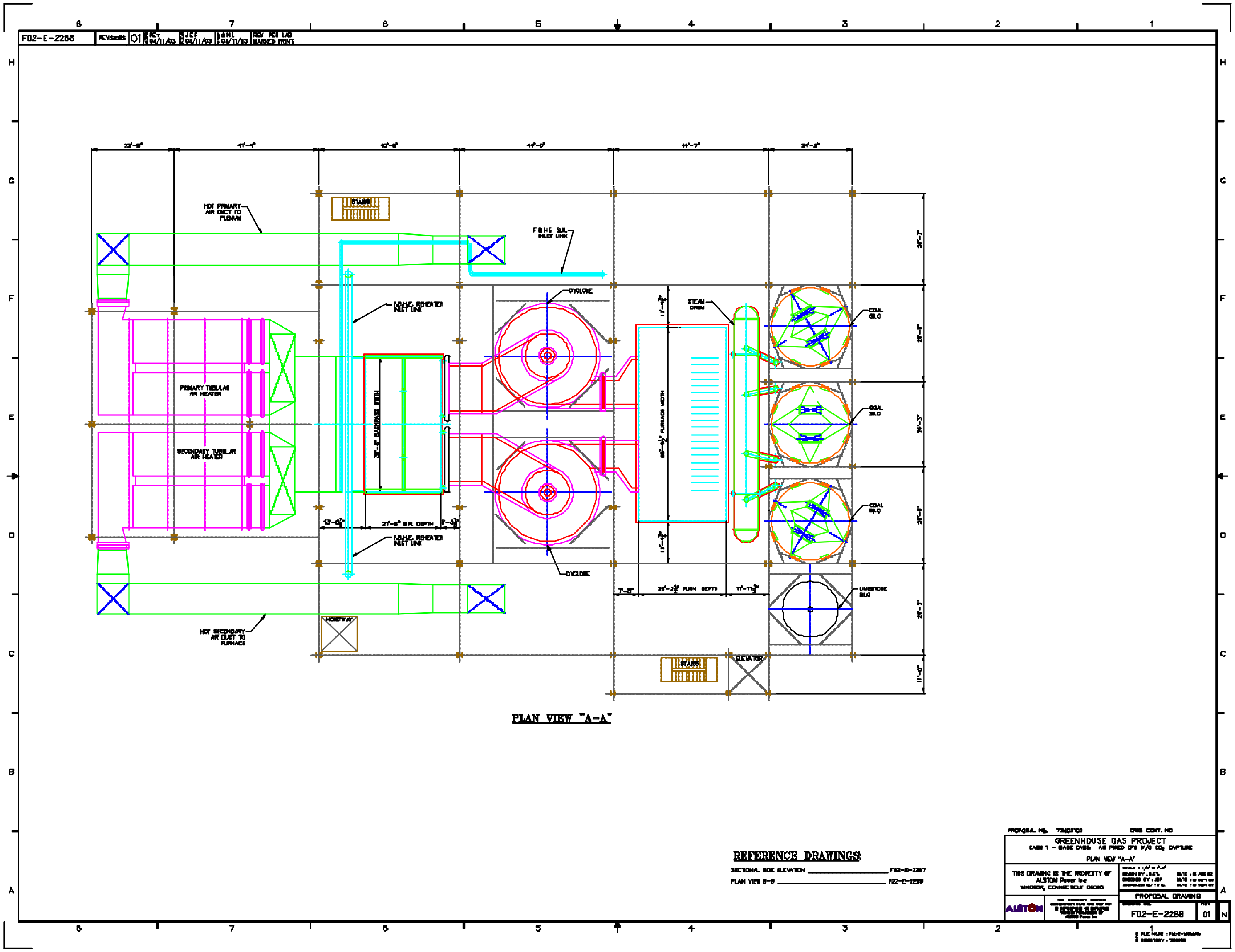
9.2. Appendix II – Drawings

Appendix II provides complete arrangement drawings for each of the thirteen plants studied. The drawings are presented consecutively, starting with Case-1 and ending with Case-13. The drawings are grouped into four separate areas for each case: Boiler Island, Air Separation, Gas Processing, and Balance of Plant. Some of the cases do not have drawings in all four areas.

9.2.1. Case-1 Drawings

9.2.1.1. Case-1 Boiler Island Equipment

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PROPOSAL No. 72402102
ORIG CONT. NO
GREENHOUSE GAS PROJECT
CASE 1 - BASE CASE: AIR FIRED CFB W/O CO2 CAPTURE
PLAN VIEW "A-A"
THIS DRAWING IS THE PROPERTY OF
ALSTOM Power Inc
WINDSOR, CONNECTICUT 06095
PROPOSAL DRAWING
ALSTOM
F02-E-2288
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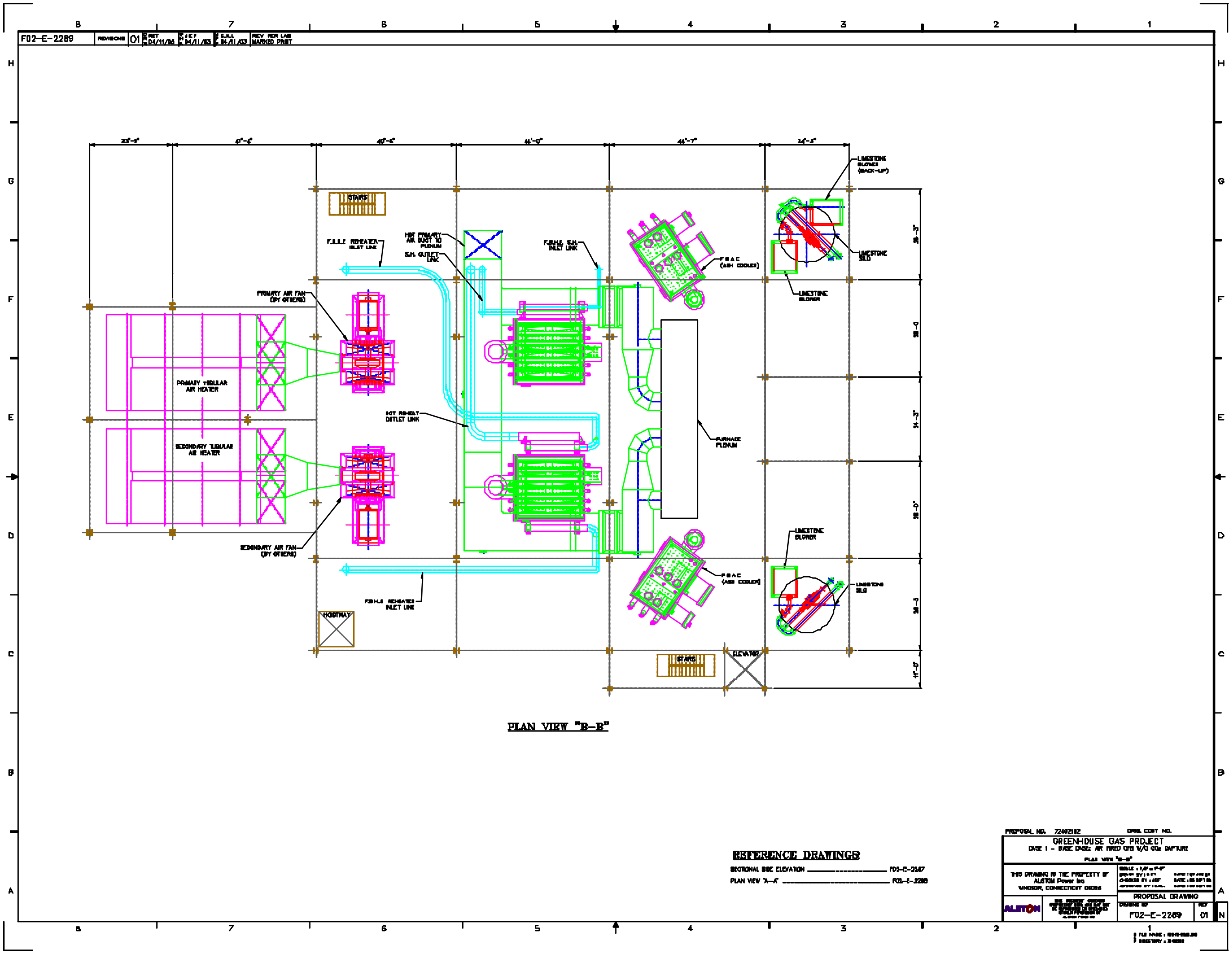
PRIMARY TUBULAR AIR HEATER
SECONDARY TUBULAR AIR HEATER
CYCLONE
COAL SILO
LIMESTONE SILO
STAIRS
ELEVATOR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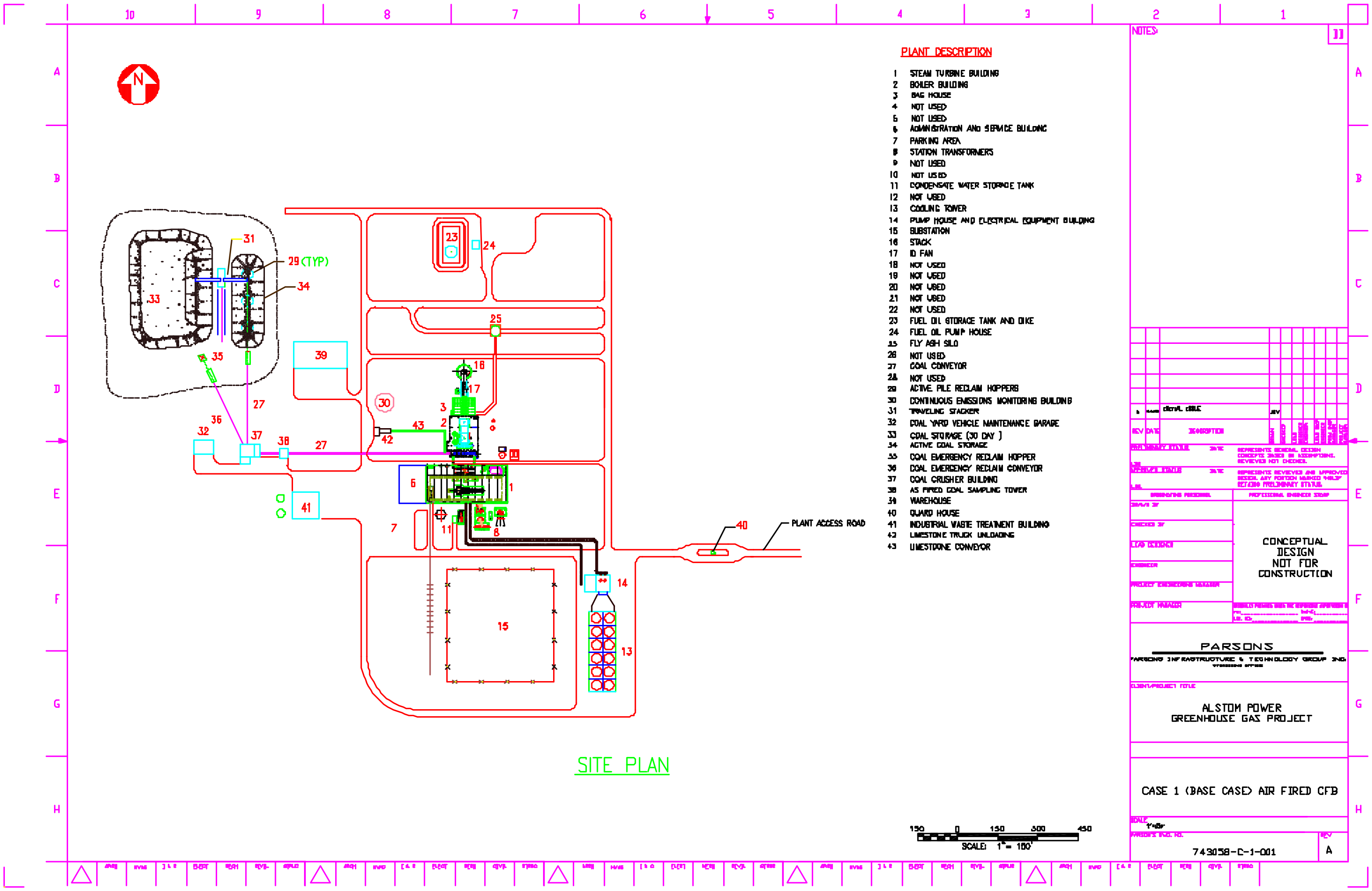
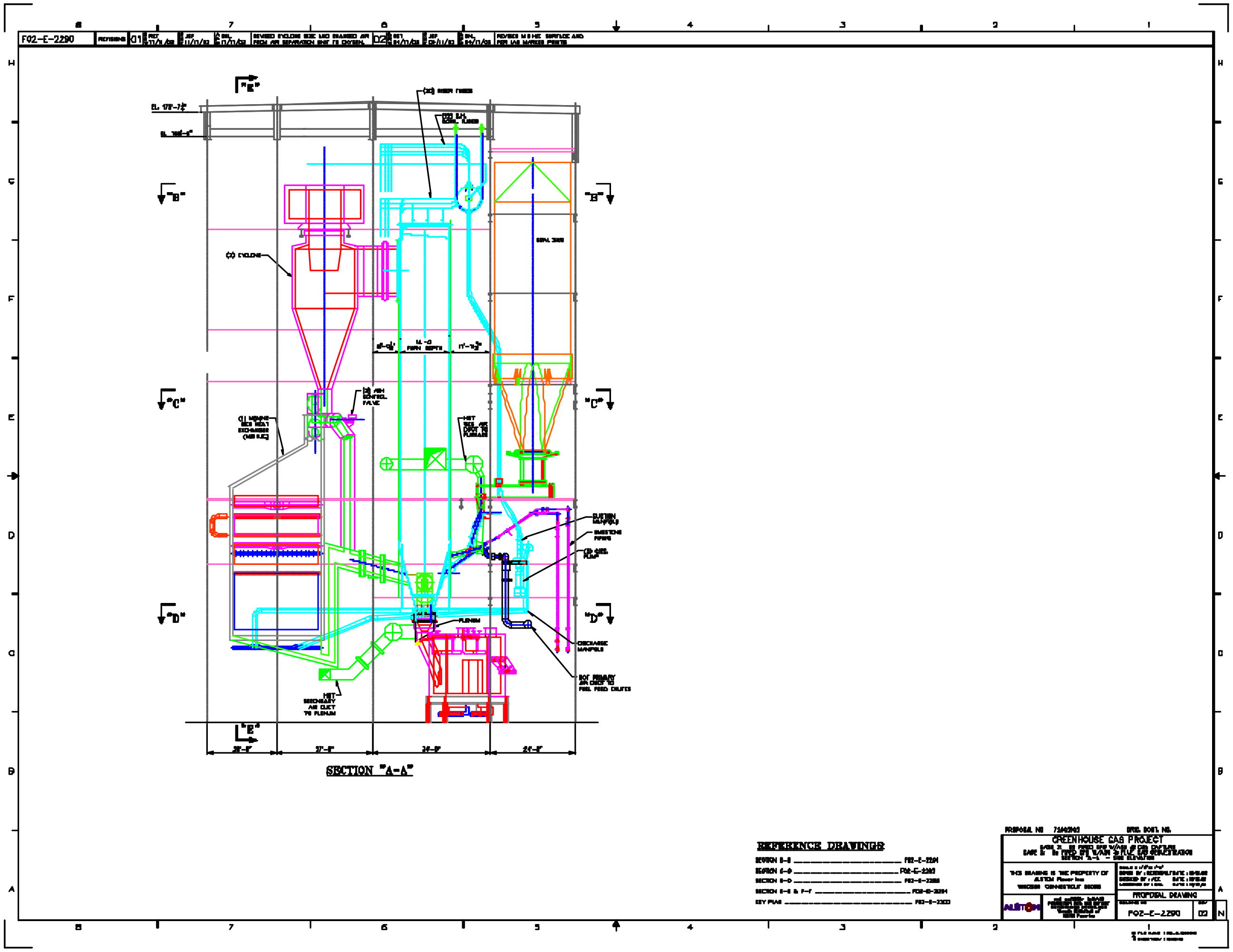
REFERENCE DRAWINGS
GREENHOUSE GAS PROJECT
PROPOSAL DRAWING
ALSTOM

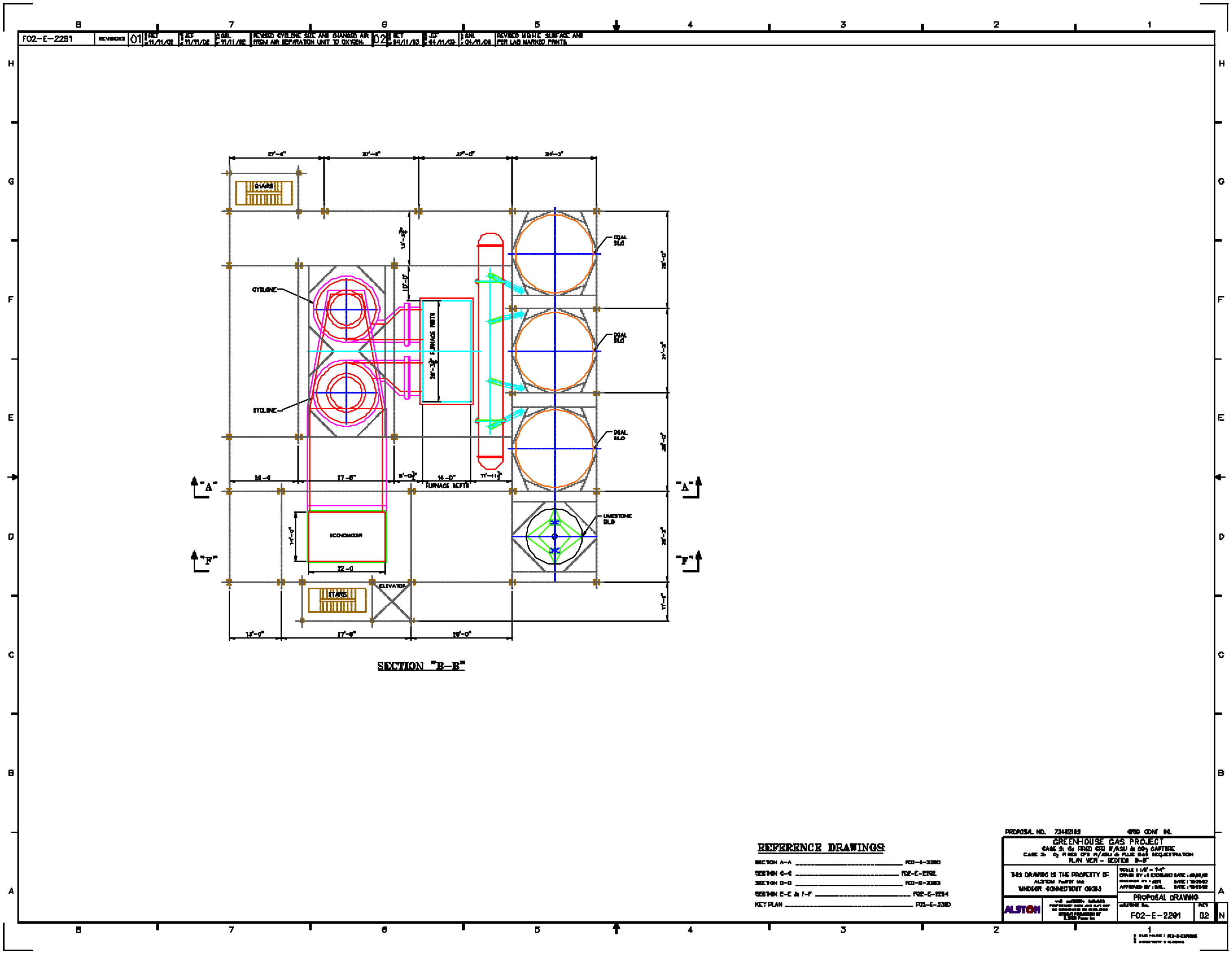
9.2.1.2. Case-1 Site Plan



9.2.2. Case-2 Drawings

9.2.2.1. Case-2 Boiler Island Equipment

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LIMESTONE SILO
ECONOMIZER
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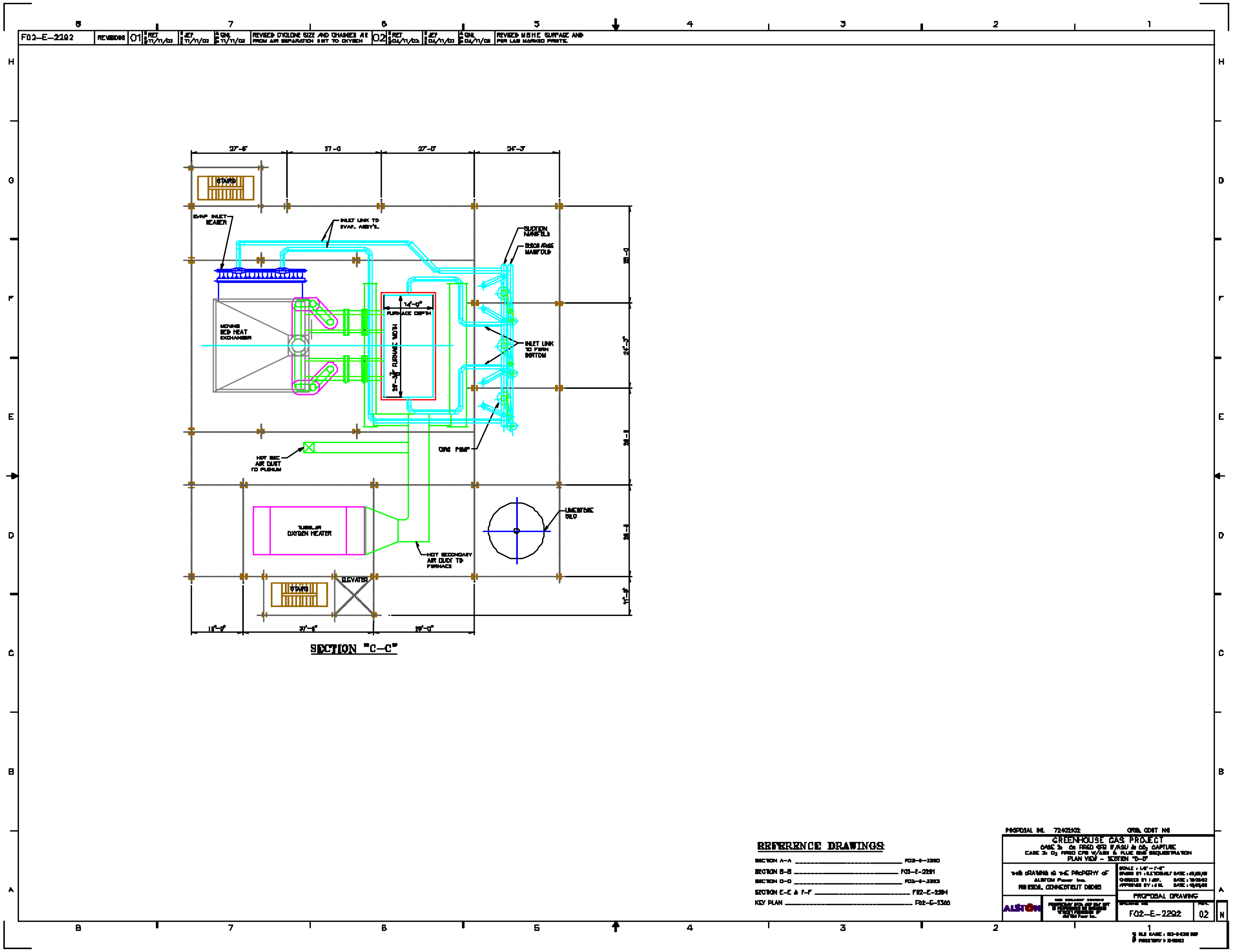
ALSTOM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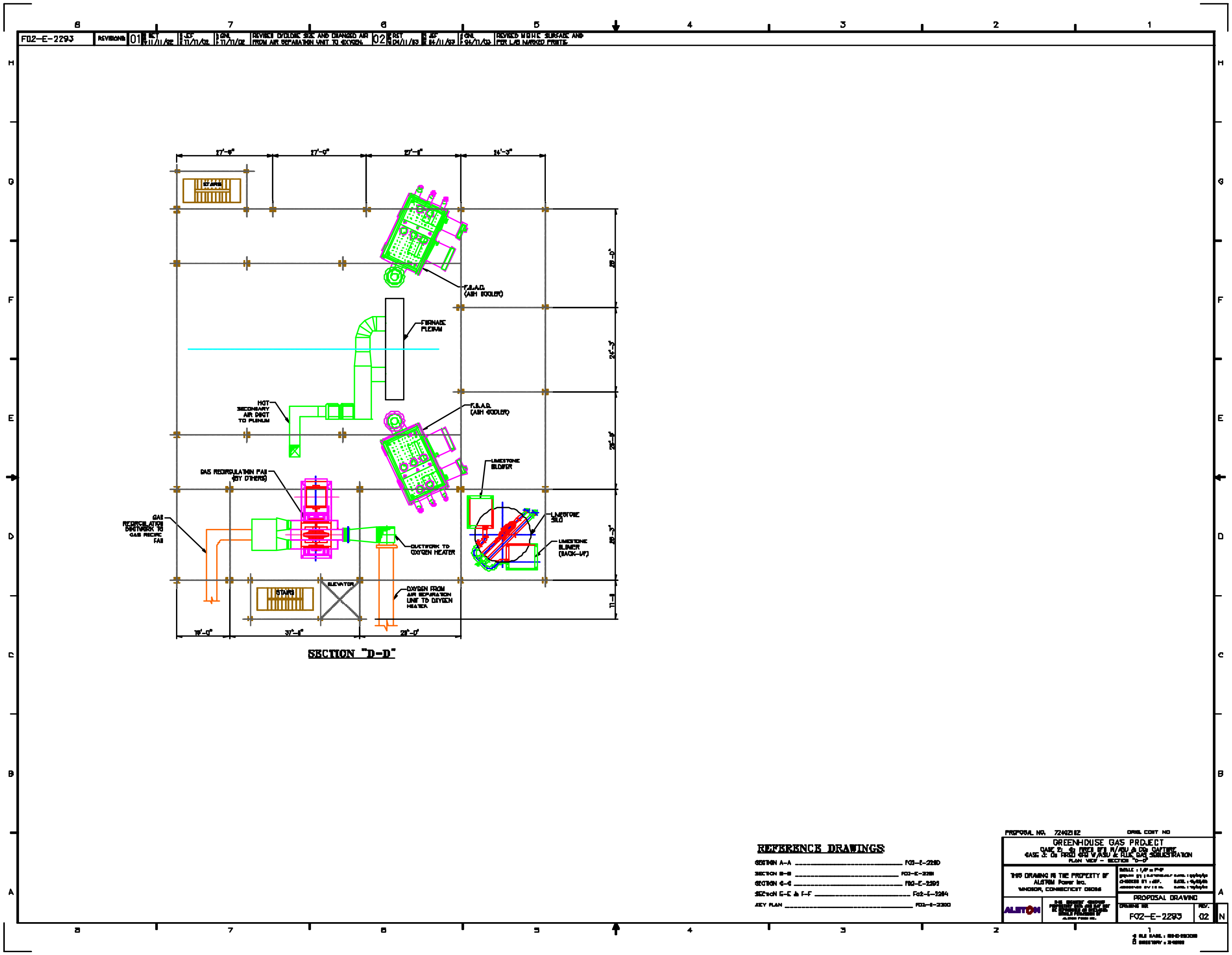
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F.B.A.C.
(ASH COOLER)
FURNACE
PLENUM
HOT
SECONDARY
AIR DUCT
TO PLENUM
F.B.A.C.
(ASH COOLER)
LIMESTONE
BLOWER
GAS RECIRCULATION FAN
(BY OTHERS)
GAS
RECIRCULATION
DUCTWORK TO
GAS RECIRC
FAN
LIMESTONE
SILO
LIMESTONE
BLOWER
(BACK-UP)
DUCTWORK TO
OXYGEN HEATER
ELEVATOR
STAIRS
OXYGEN FROM
AIR SEPARATION
UNIT TO OXYGEN
HEATER
27'-6"
27'-0"
27'-6"
24'-3"
28'-0"
24'-3"
28'-6"
28'-3"
11'-6"
18'-0"
37'-6"
28'-0"
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SECTION B-B F02-E-2281
SECTION C-C F02-E-2292
SECTION E-E & F-F F02-E-2294
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WINDSOR, CONNECTICUT 06095
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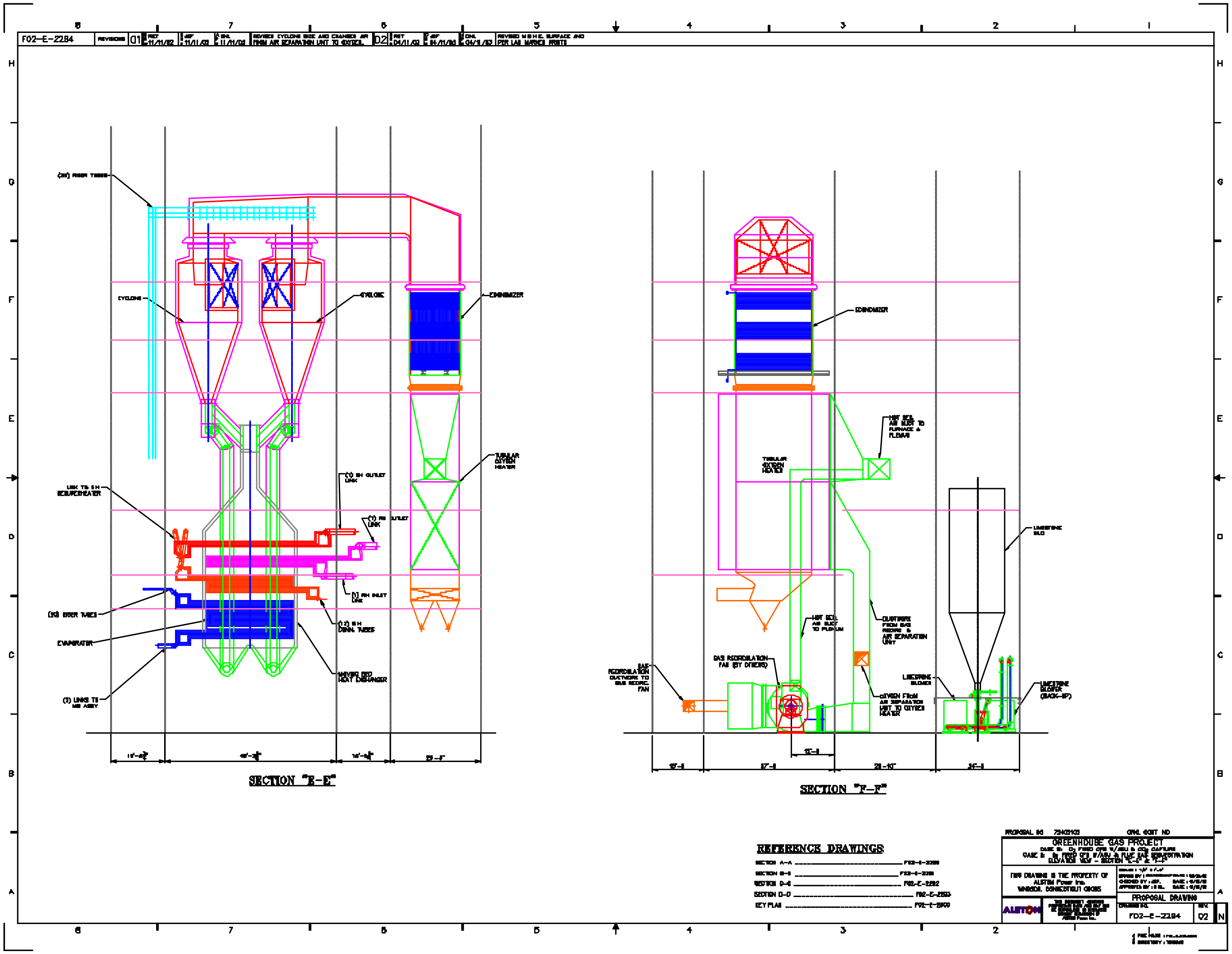
REV.
02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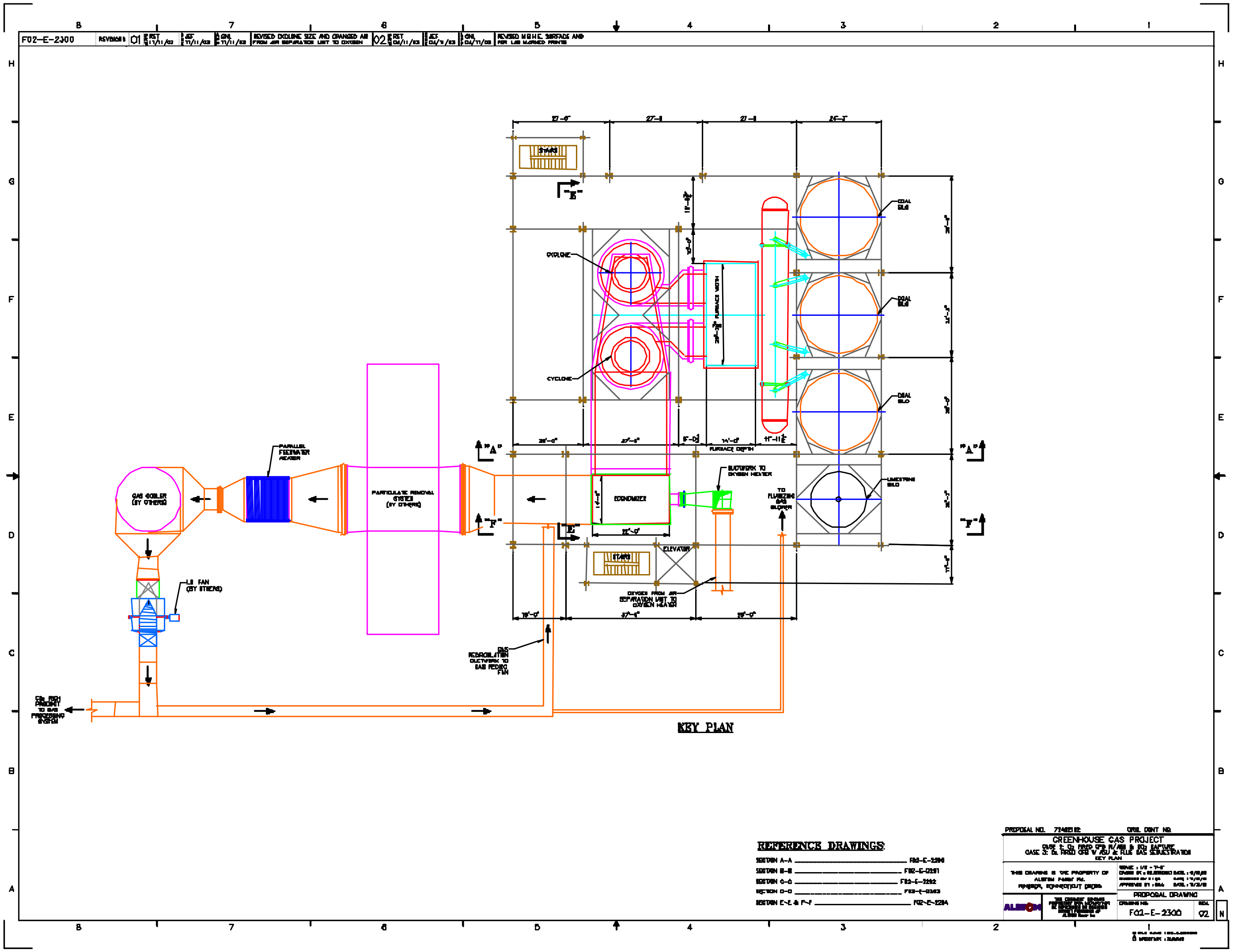
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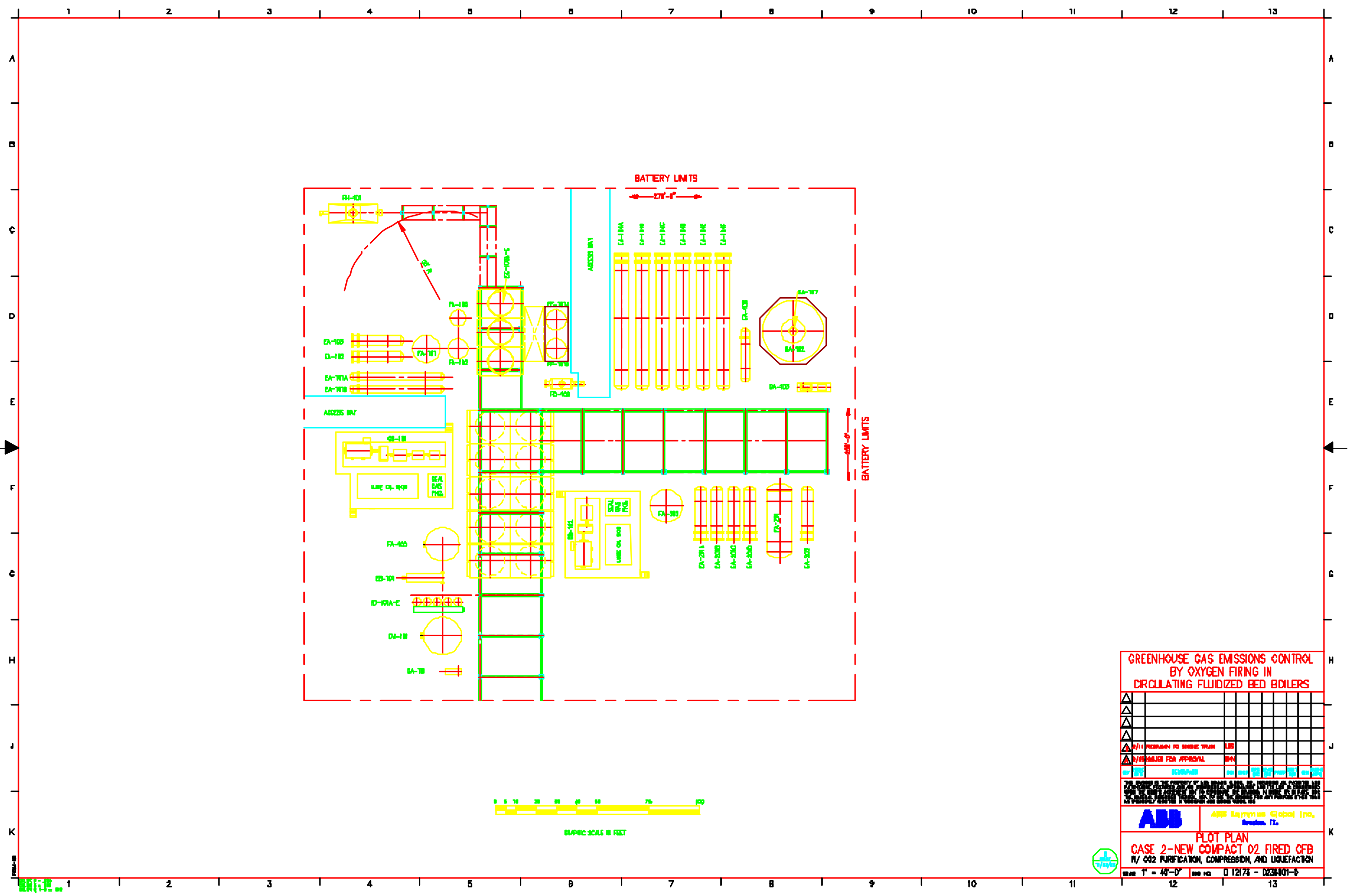
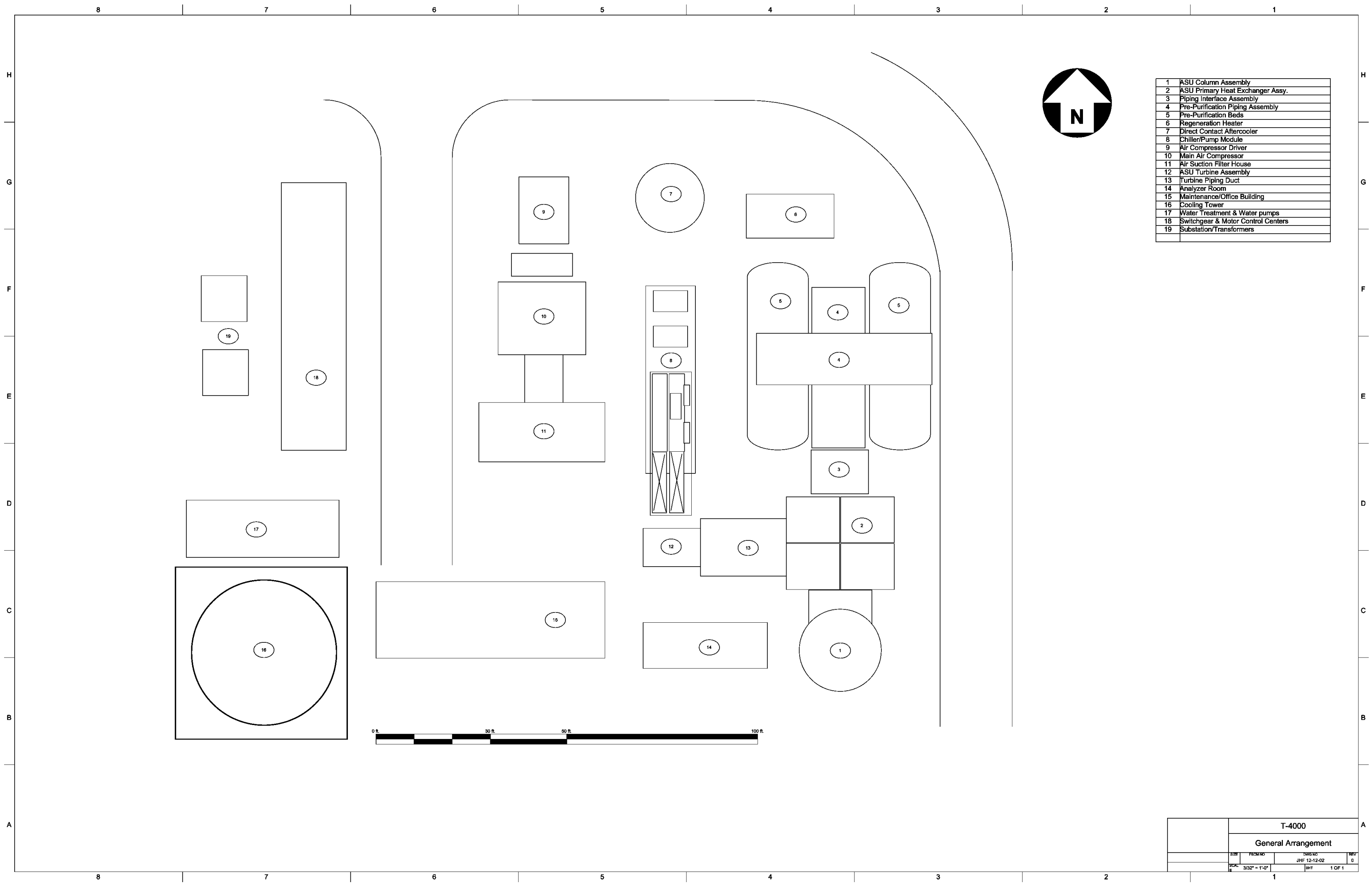
ECONOMIZER
PARTICULATE REMOVAL SYSTEM (BY OTHERS)
GAS COOLER (BY OTHERS)
PARALLEL FEEDWATER HEATER
COAL SILO
LIMESTONE SILO
STAIRS
ELEVATOR
GREENHOUSE GAS PROJECT
PROPOSAL DRAWING
F02-E-2300

9.2.2.2. Case-2 Gas Processing System Equipment

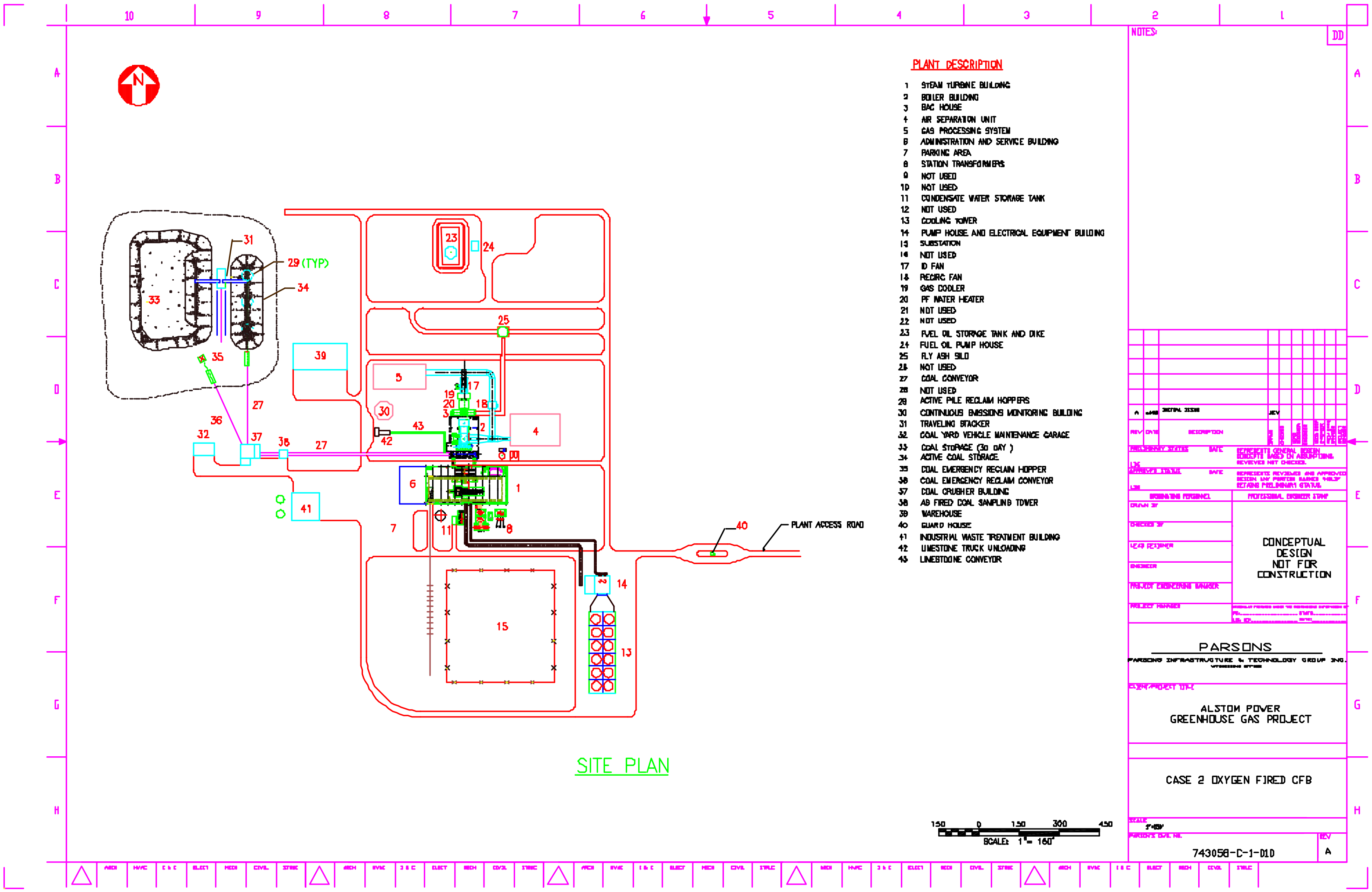


CO2 UNIT CASE 2.dwg LAST REVISED: 02/11/03; 09:08 by LSUTHERL

9.2.2.3. Case-2 Air Separation Unit Equipment



9.2.2.4. Case-2 Site Plan

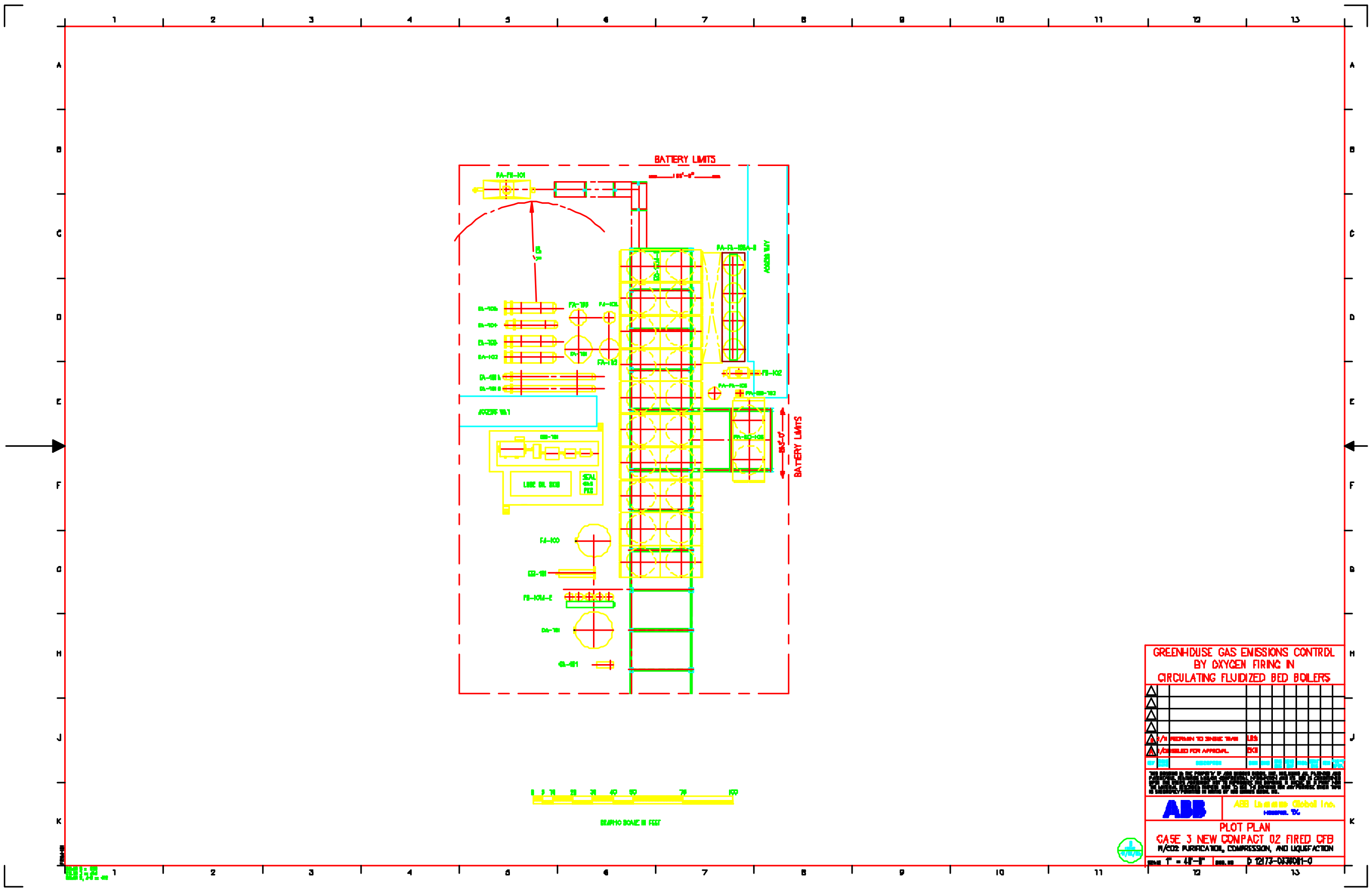


9.2.3. Case-3 Drawings

9.2.3.1. Case-3 Boiler Island Equipment

The Boiler Island for Case 3 is identical to Case 2. Refer to Section 9.2.2.1

9.2.3.2. Case-3 Gas Processing System Equipment

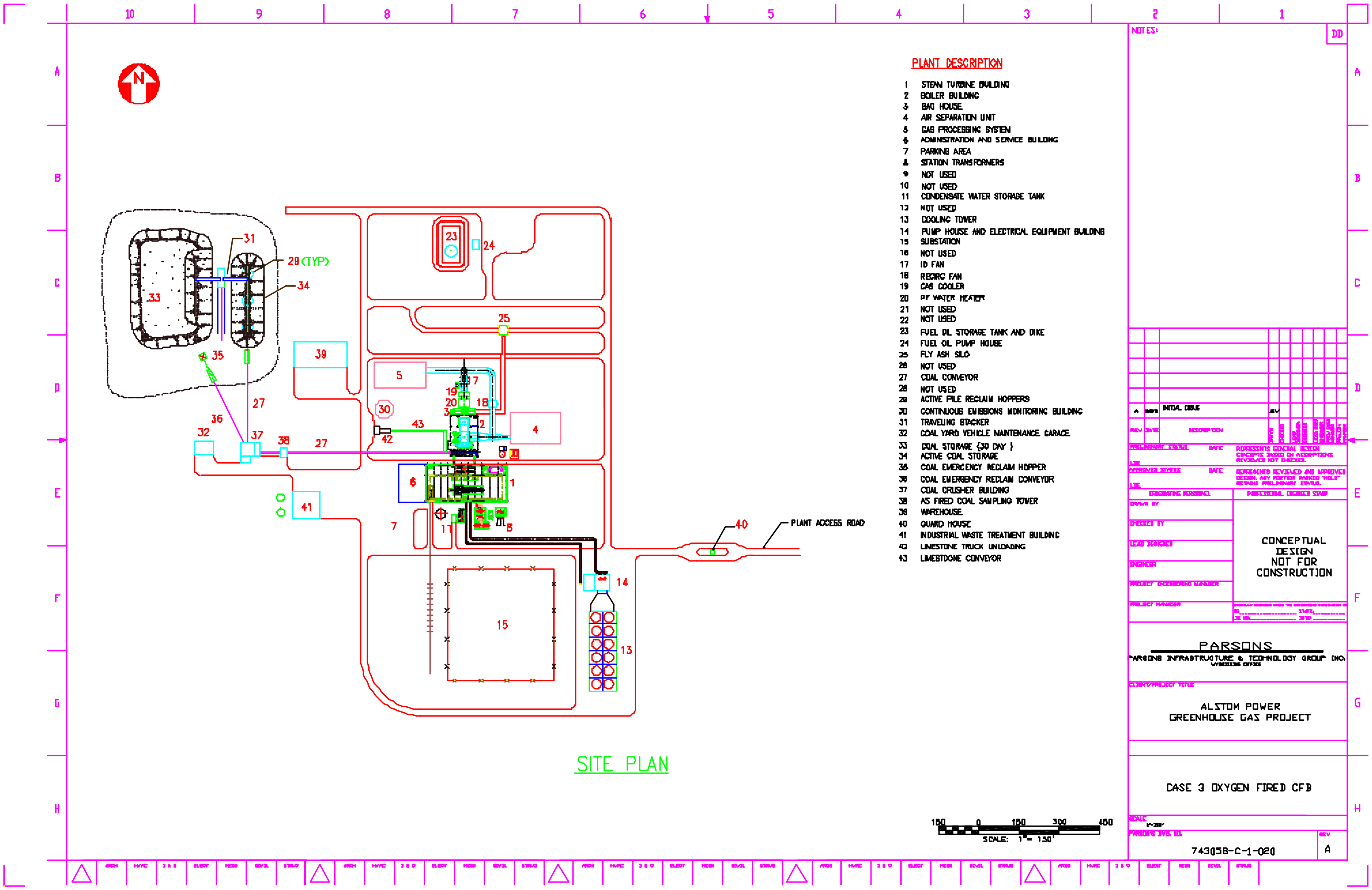


CO2 UNIT CASE 3.dwg LAST REVISED: 02/11/03; 09:05 by LSUTHERL

9.2.3.3. Case-3 Air Separation Unit Equipment

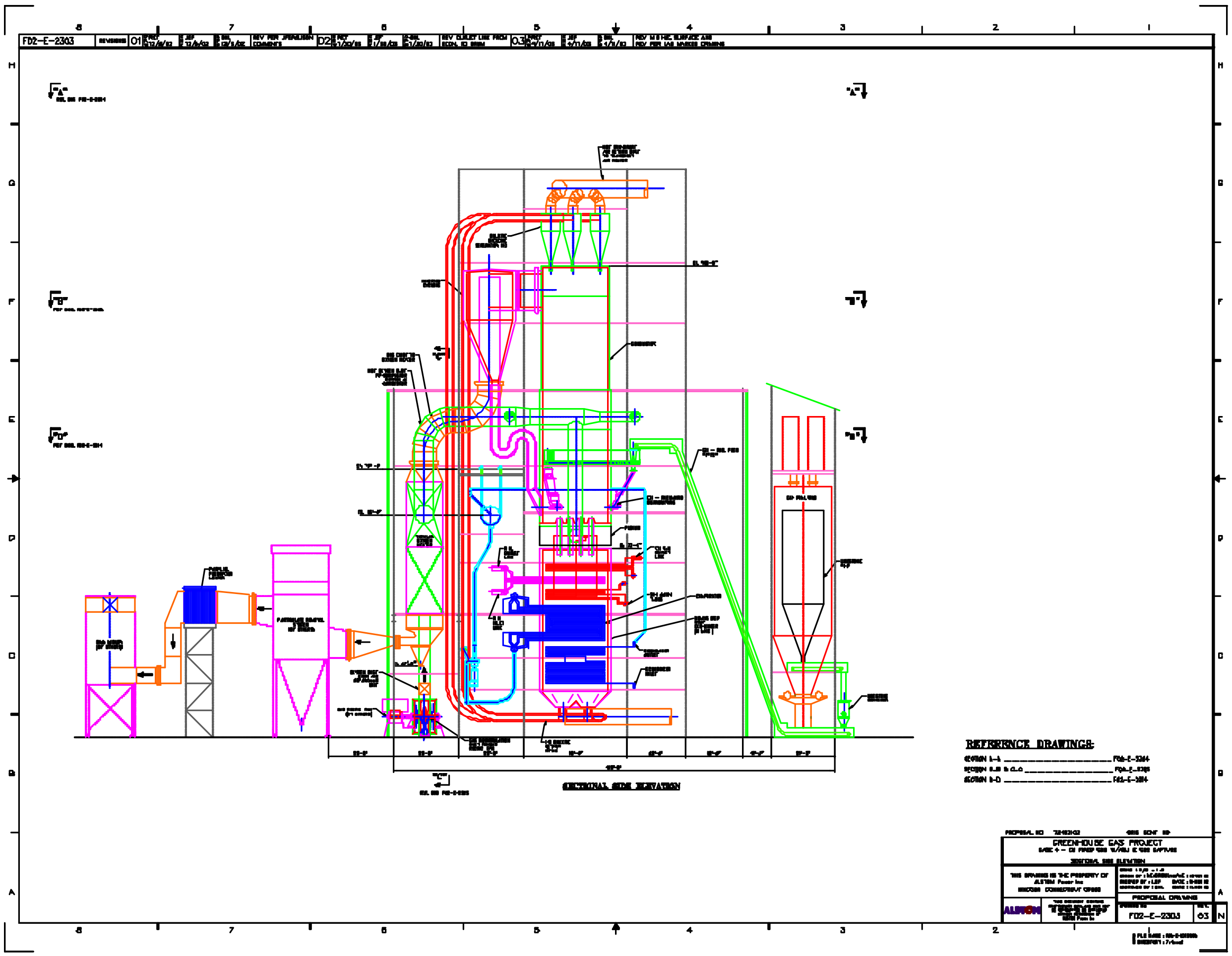
The Air Separation Unit for Case 3 is identical to Case 2. Refer to Section 9.2.3.3.

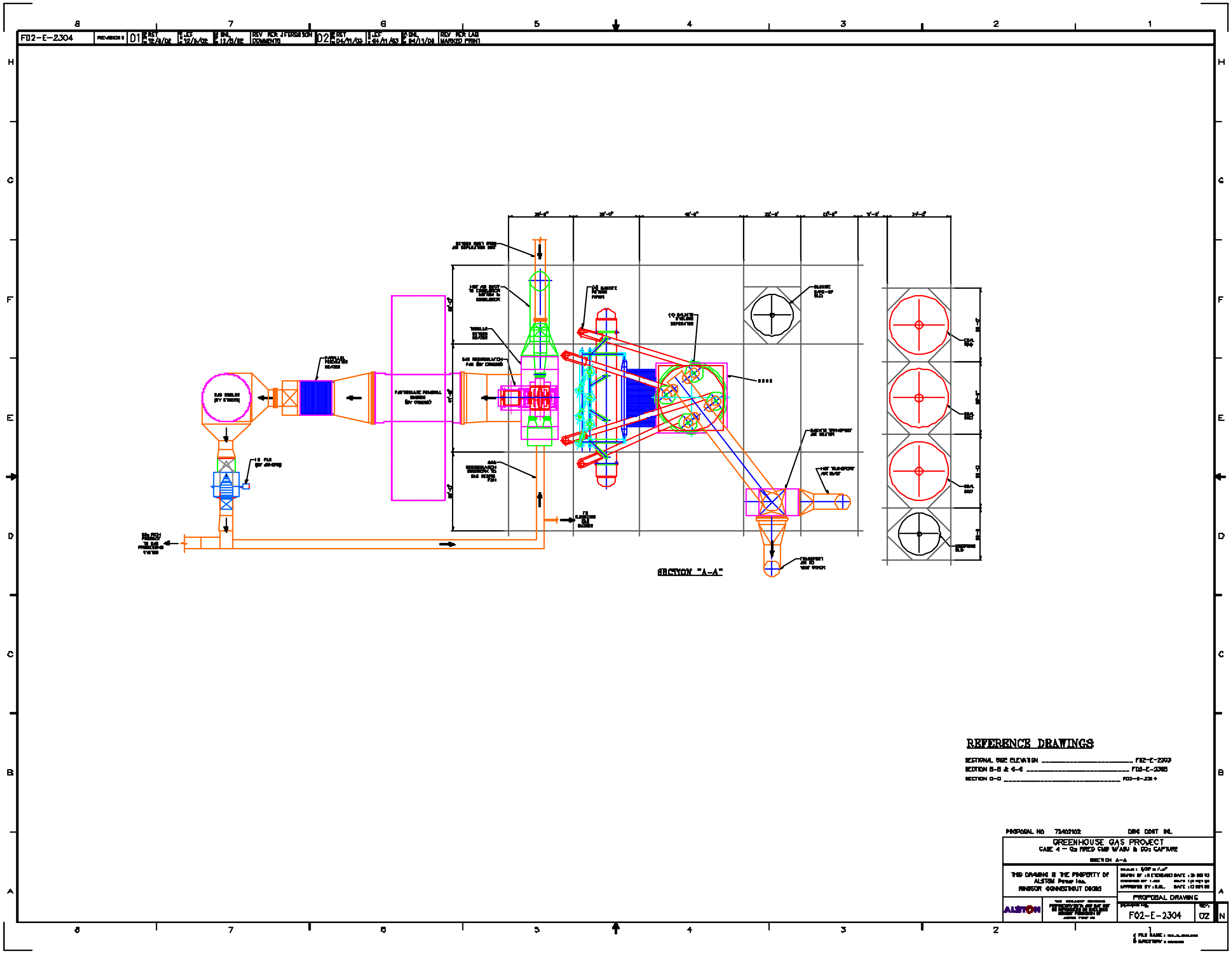
9.2.3.4. Case-3 Site Plan



9.2.4. Case-4 Drawings

9.2.4.1. Case-4 Boiler Island Equipment

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SECTION D-D F02-E-2314
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GREENHOUSE GAS PROJECT
CASE 4 – O2 FIRED CMB W/ASU & CO2 CAPTURE
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WINDSOR CONNECTICUT 06095
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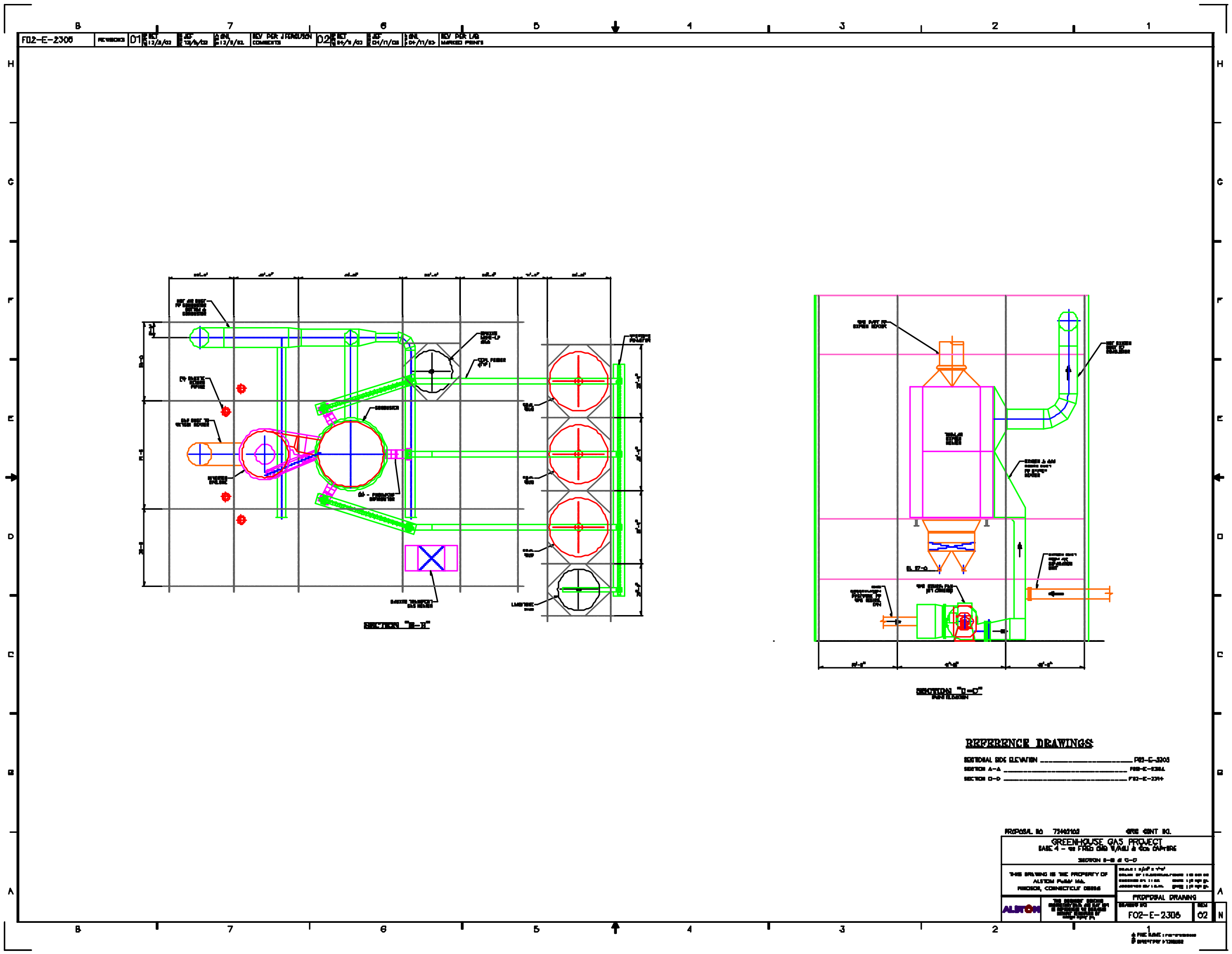
DIRECTORY :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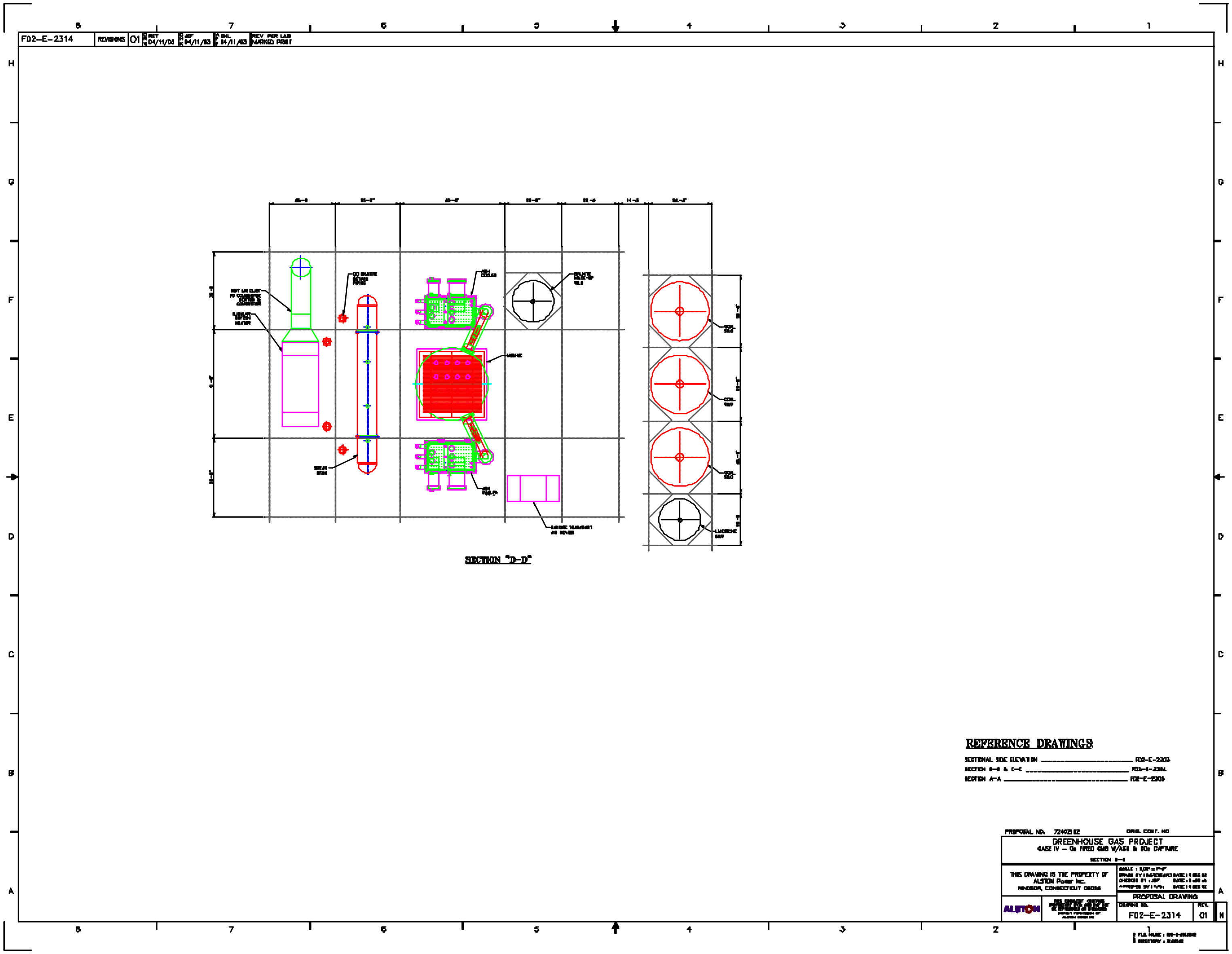
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ALSTOM
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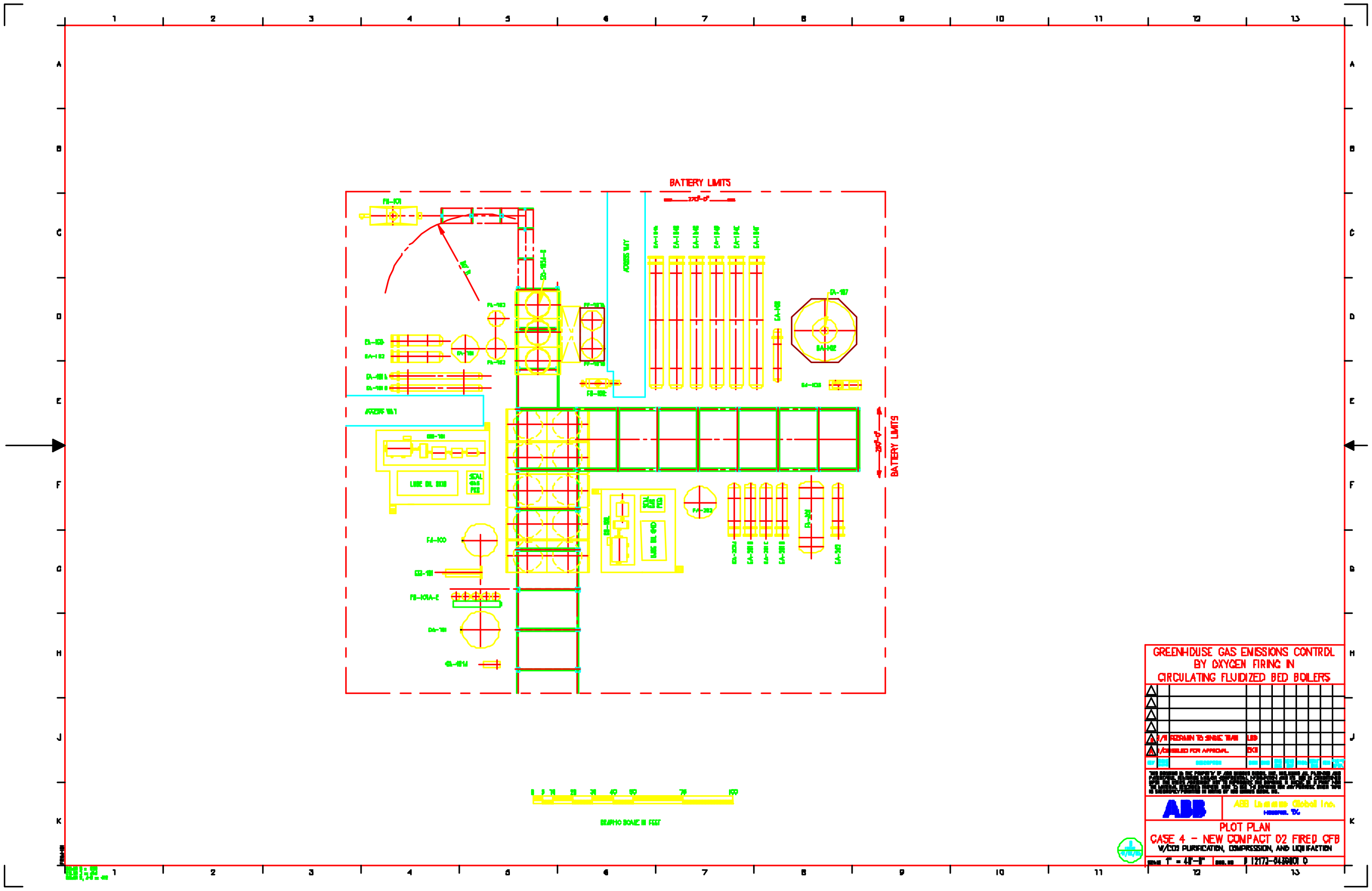
9.2.4.2. Case-4 Gas Processing System Equipment



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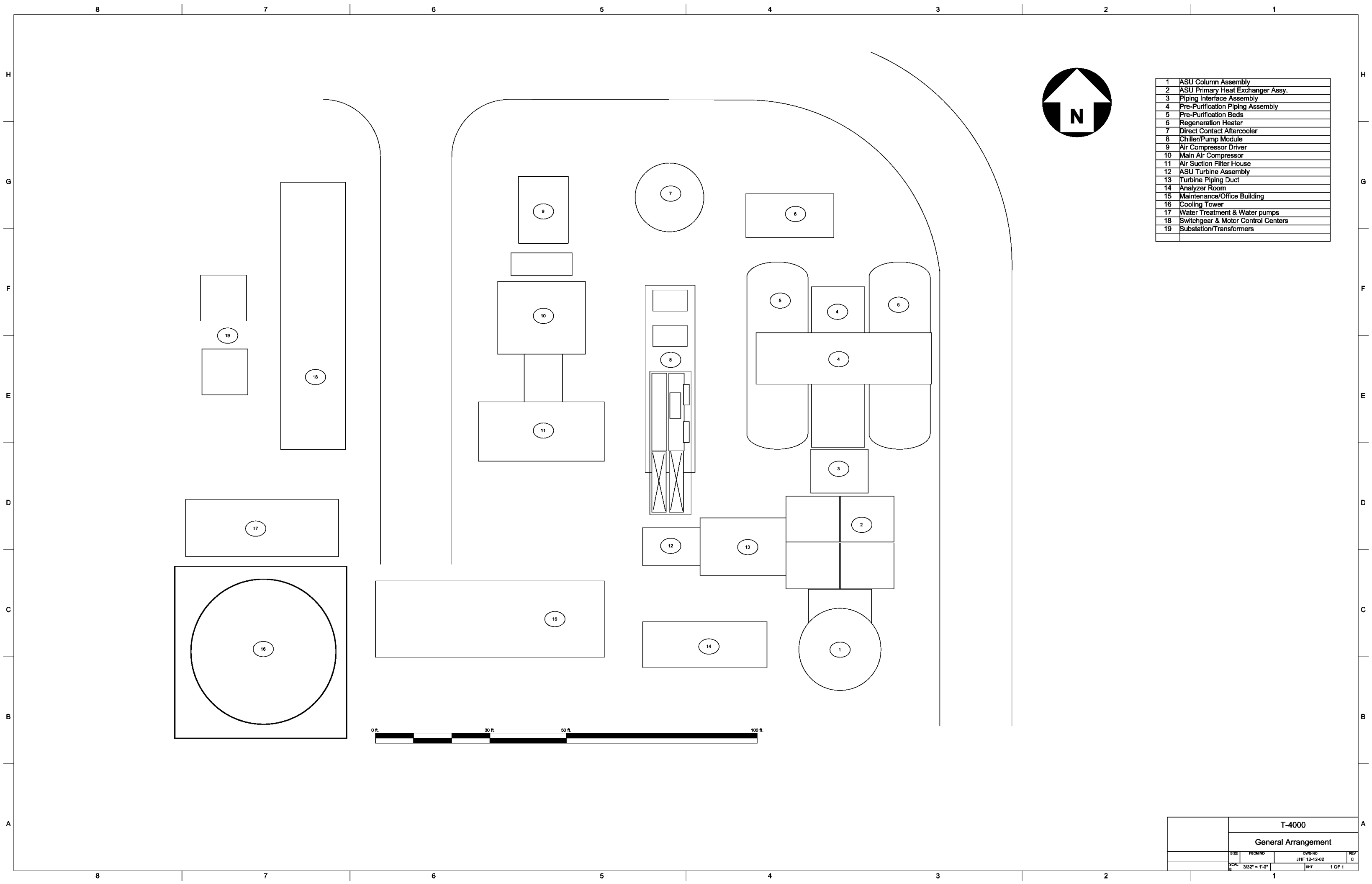
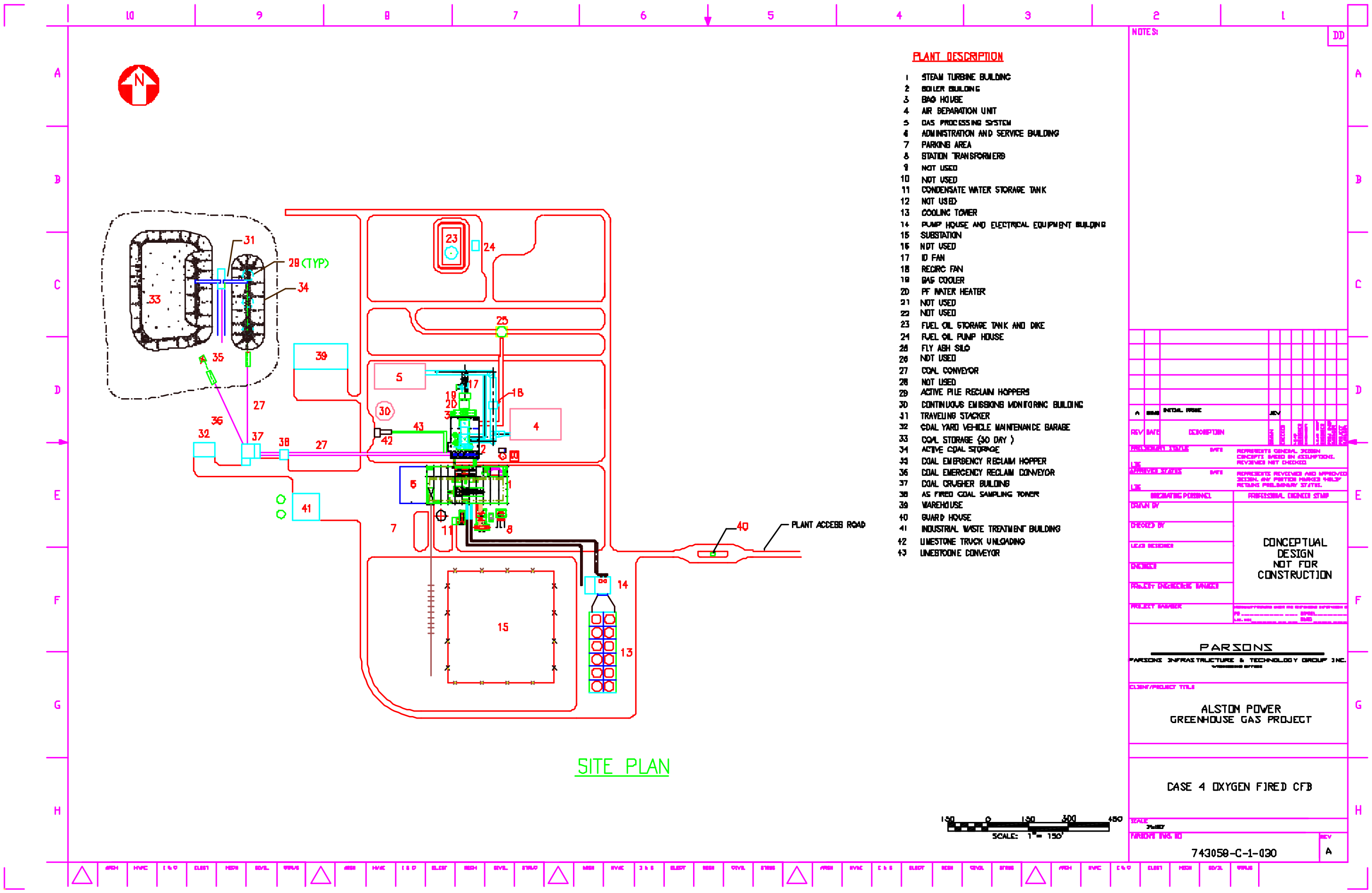
9.2.4.3. Case-4 Air Separation Unit Equipment

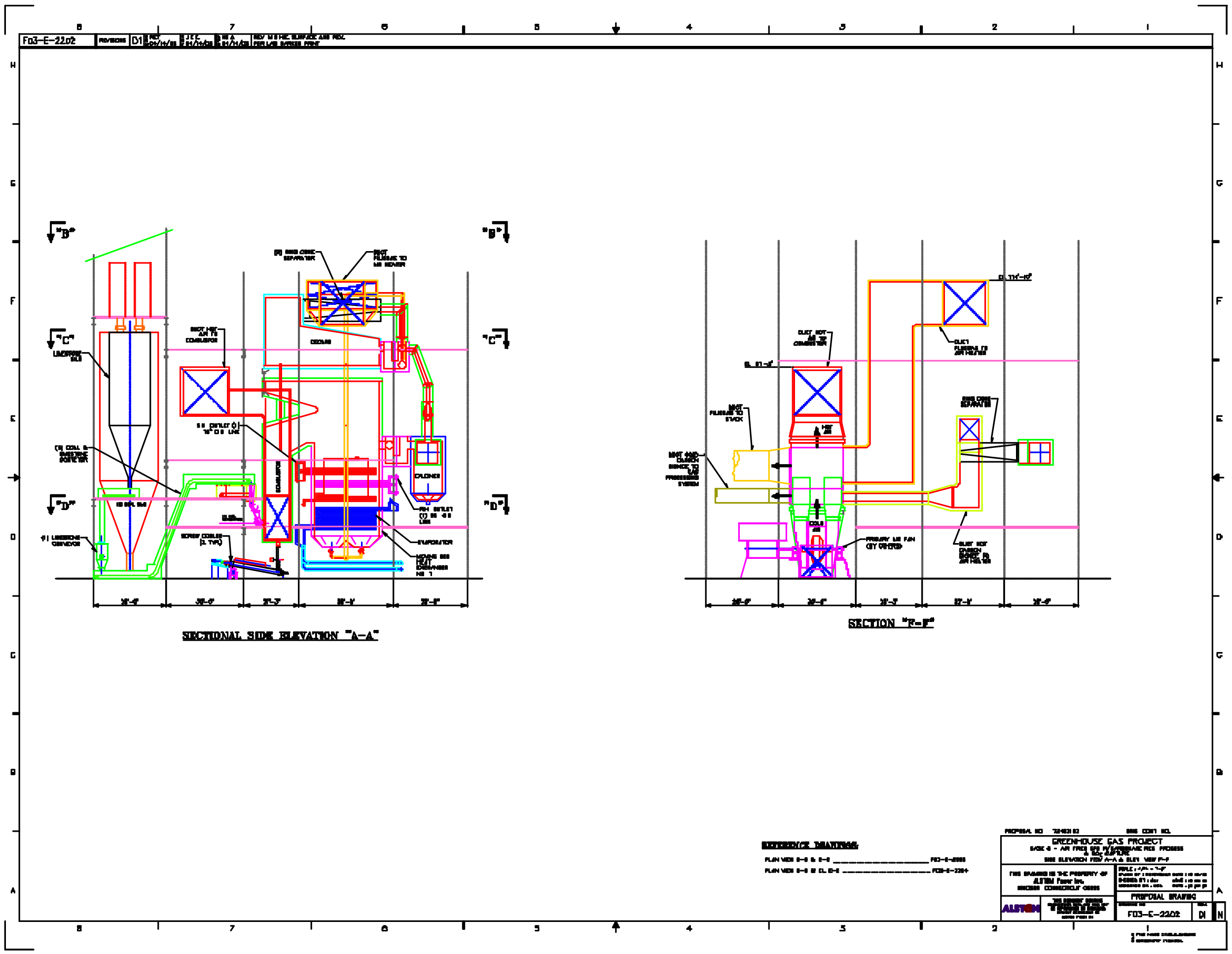


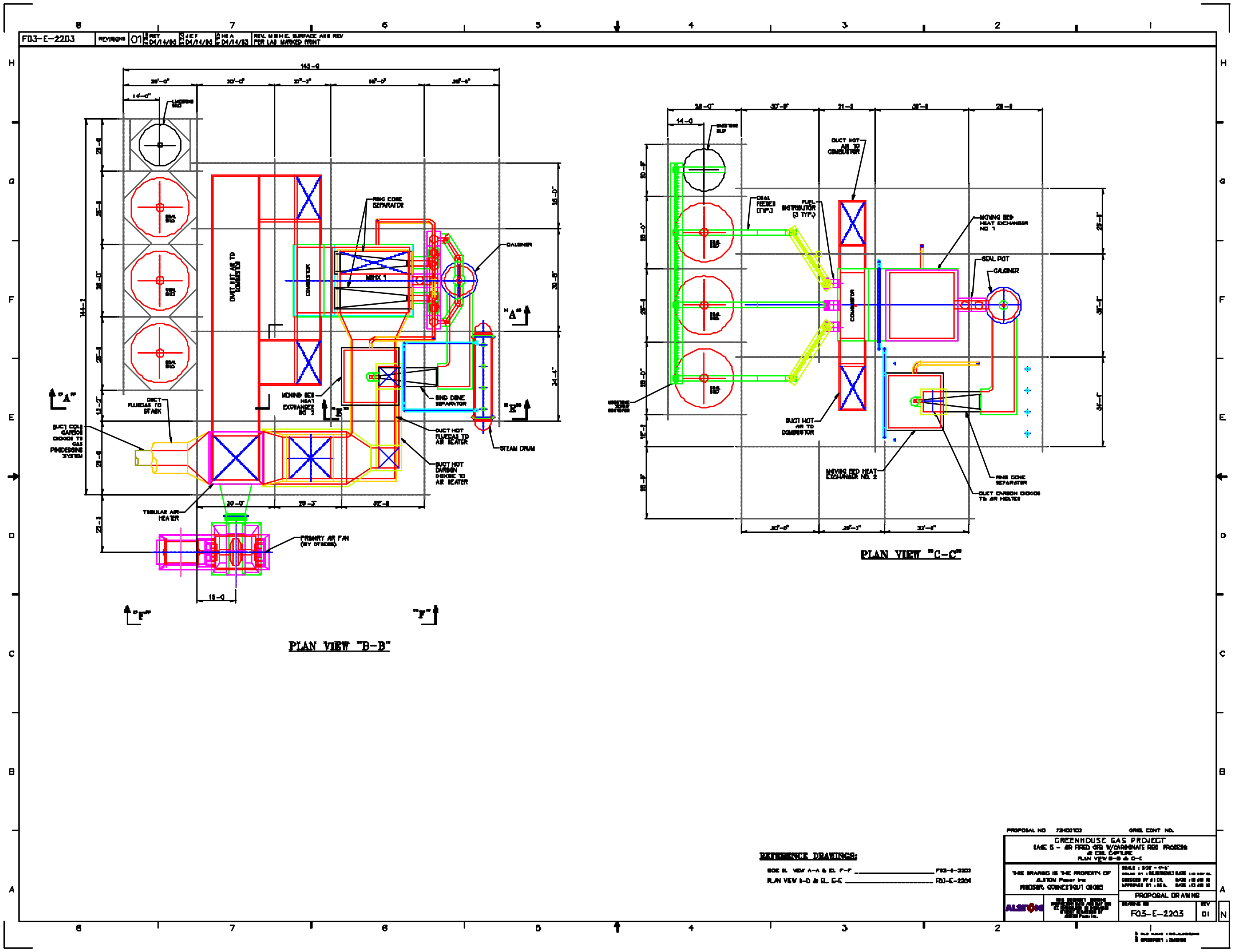
9.2.4.4. Case-4 Site Plan

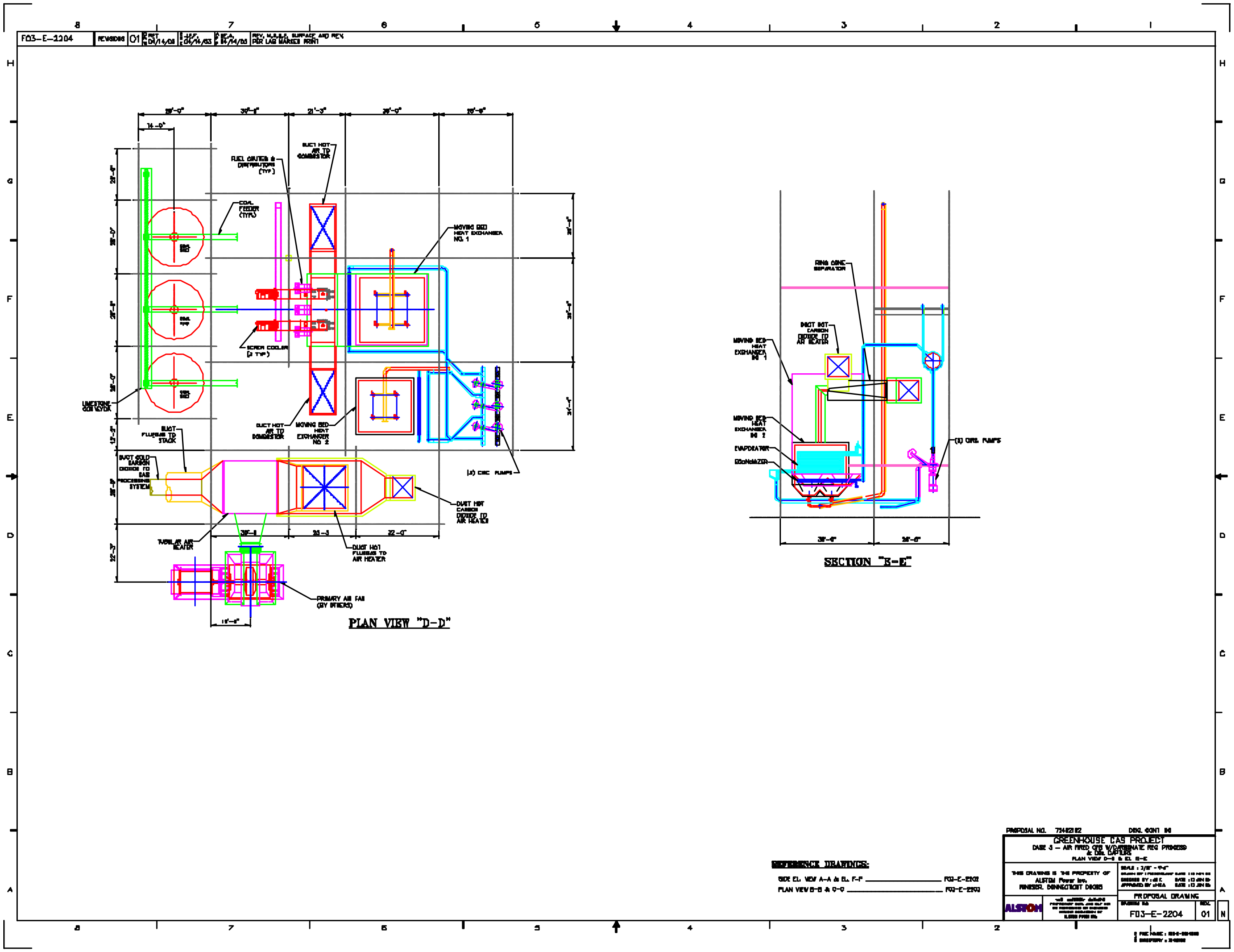


9.2.5. Case-5 Drawings

9.2.5.1. Case-5 Boiler Island Equipment




F03-E-2203
PLAN VIEW "B-B"
PLAN VIEW "C-C"
REFERENCE DRAWINGS:
GREENHOUSE GAS PROJECT
PROPOSAL DRAWING
F03-E-2203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PLAN VIEW "D-D"
SECTION "E-E"
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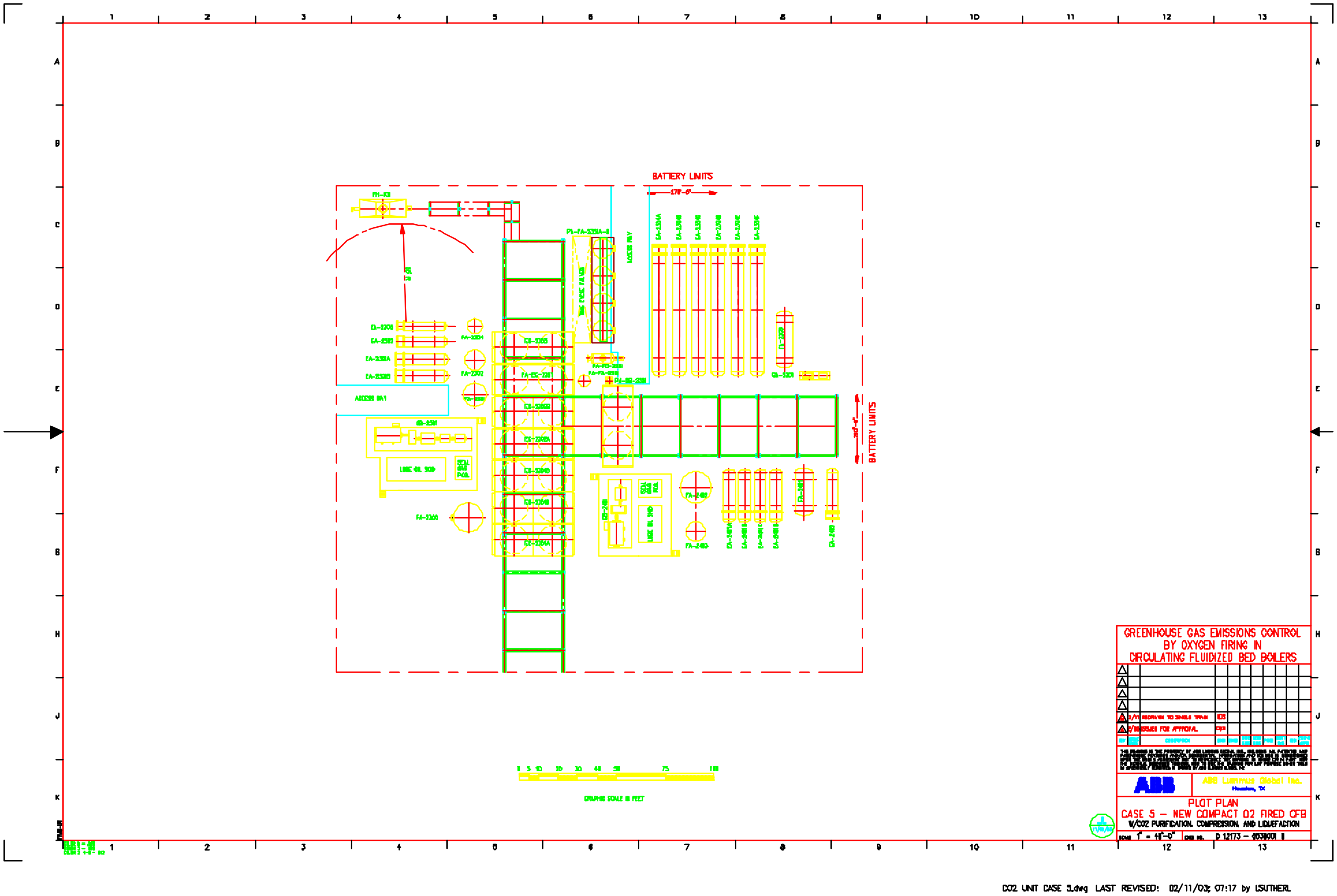
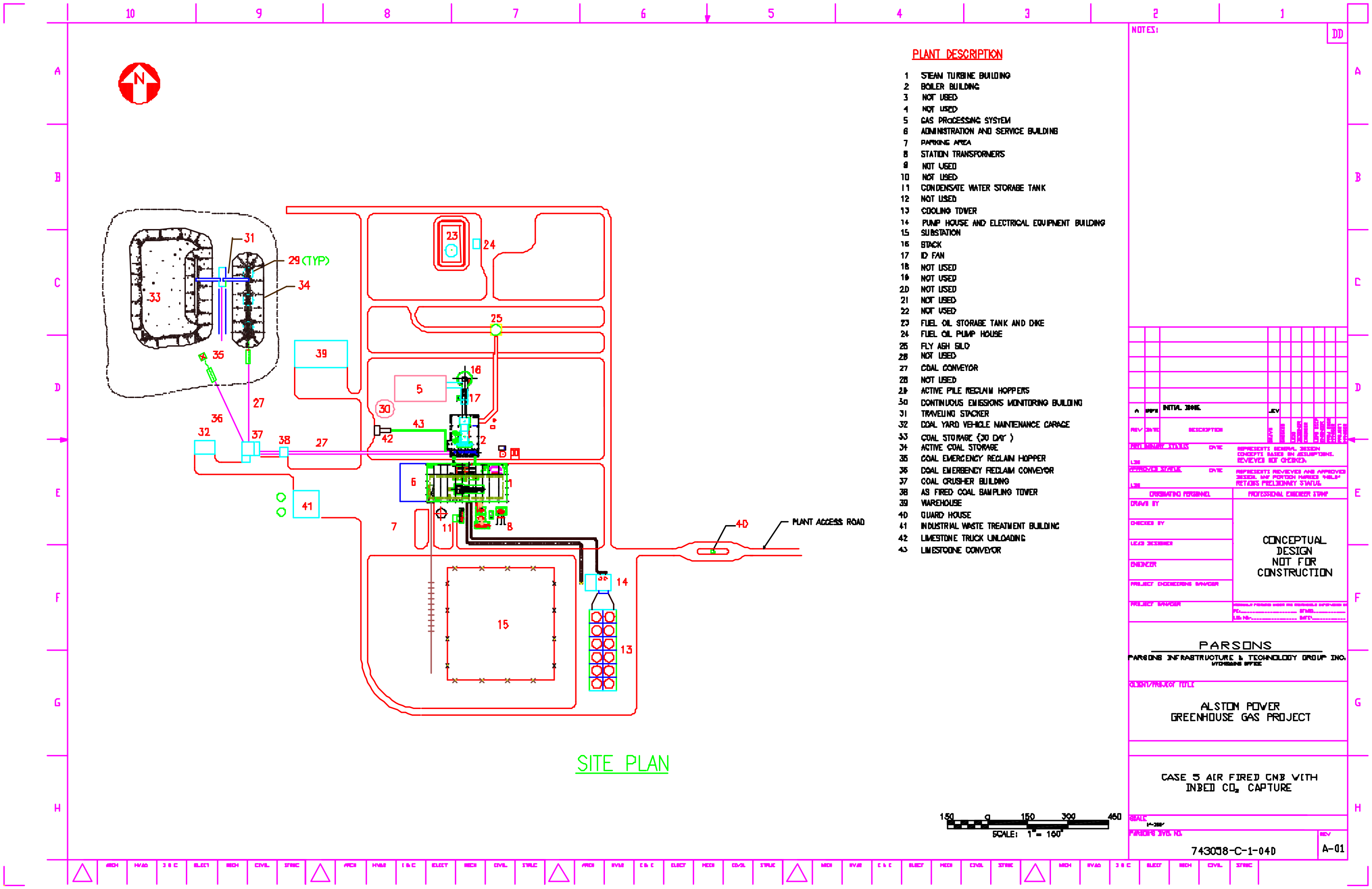
PROPOSAL DRAWING
F03-E-2204
ALSTOM

9.2.5.2. Case-5 Gas Processing System Equipment

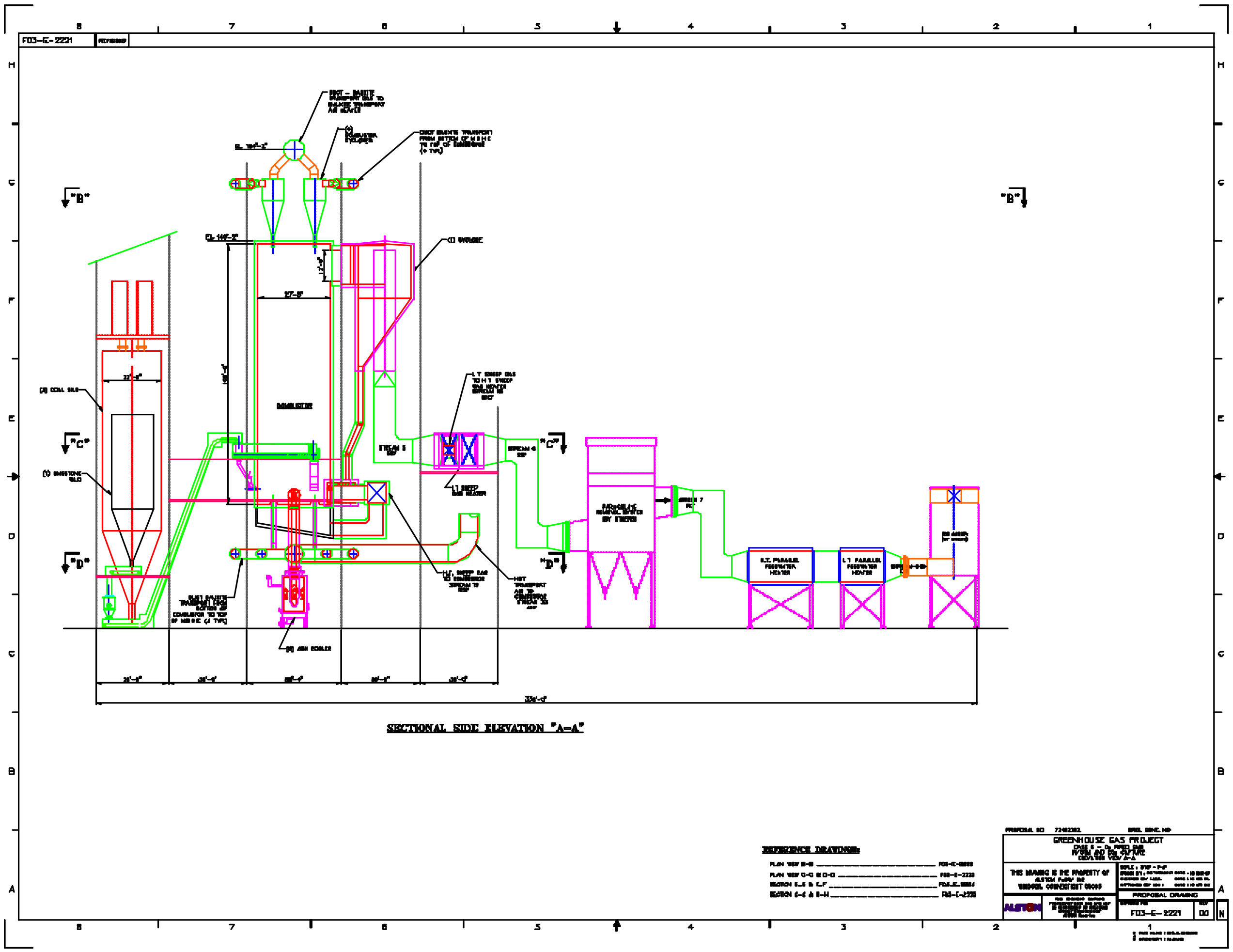


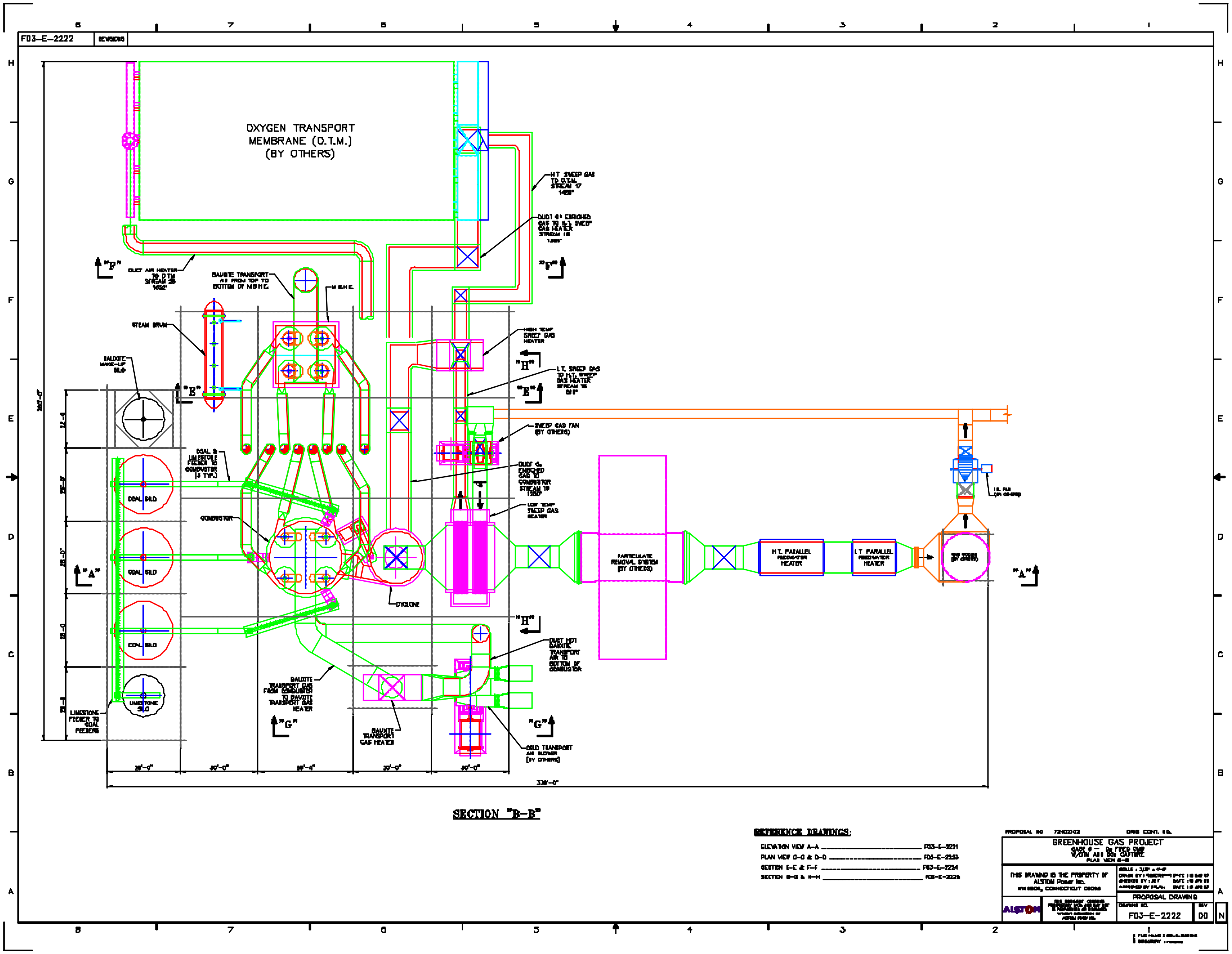
9.2.5.3. Case-5 Site Plan

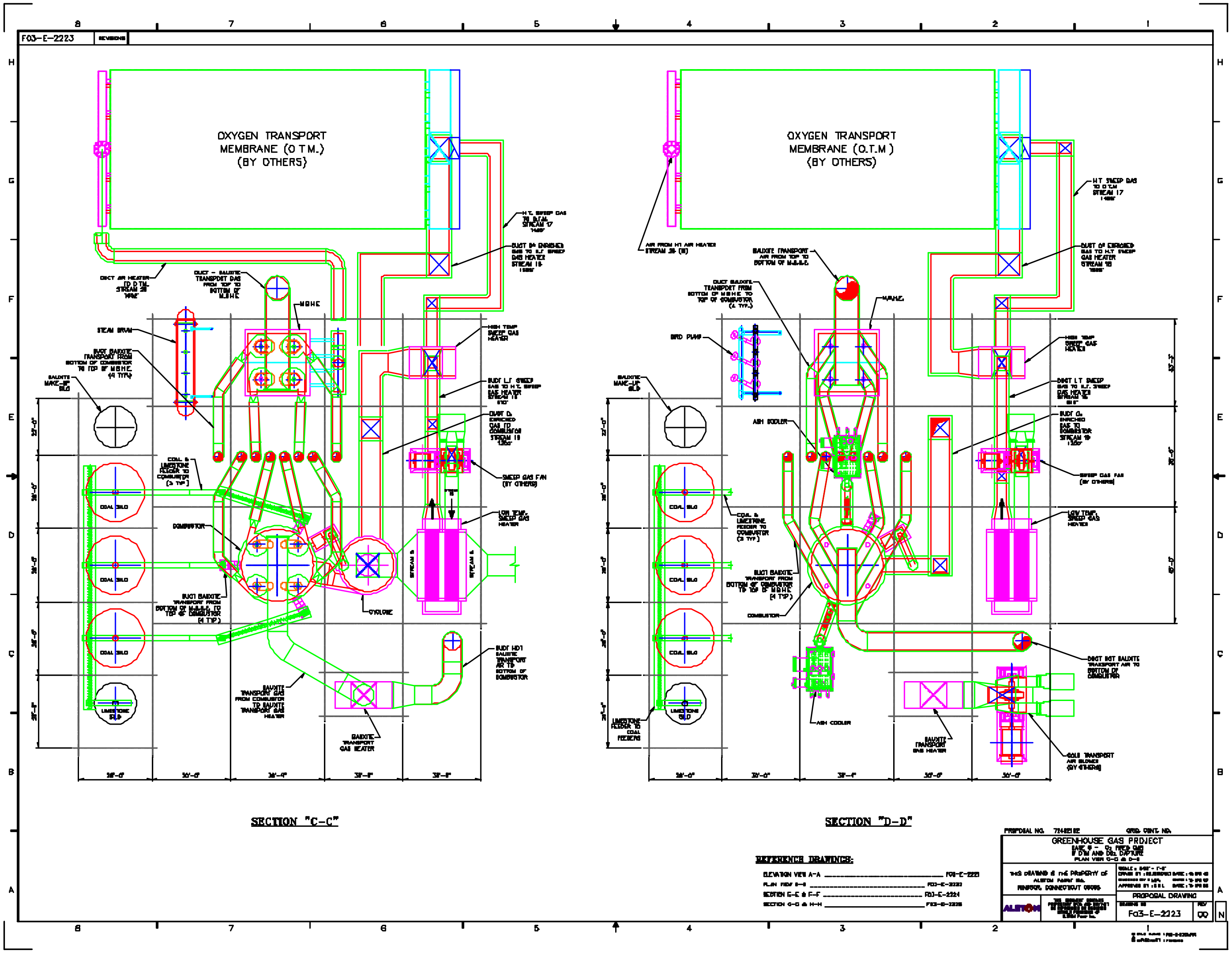


9.2.6. Case-6 Drawings

9.2.6.1. Case-6 Boiler Island Equipment




OXYGEN TRANSPORT
MEMBRANE (O.T.M.)
(BY OTHERS)
SECTION "B-B"
H.T. PARALLEL FEEDWATER HEATER
L.T. PARALLEL FEEDWATER HEATER
PARTICULATE REMOVAL SYSTEM (BY OTHERS)
STEAM DRUM
CYCLONE
COMBUSTOR
COAL SILO
LIMESTONE SILO
SWEEP GAS FAN (BY OTHERS)
REFERENCE DRAWINGS:
BREENHOUSE GAS PROJECT
F03-E-2222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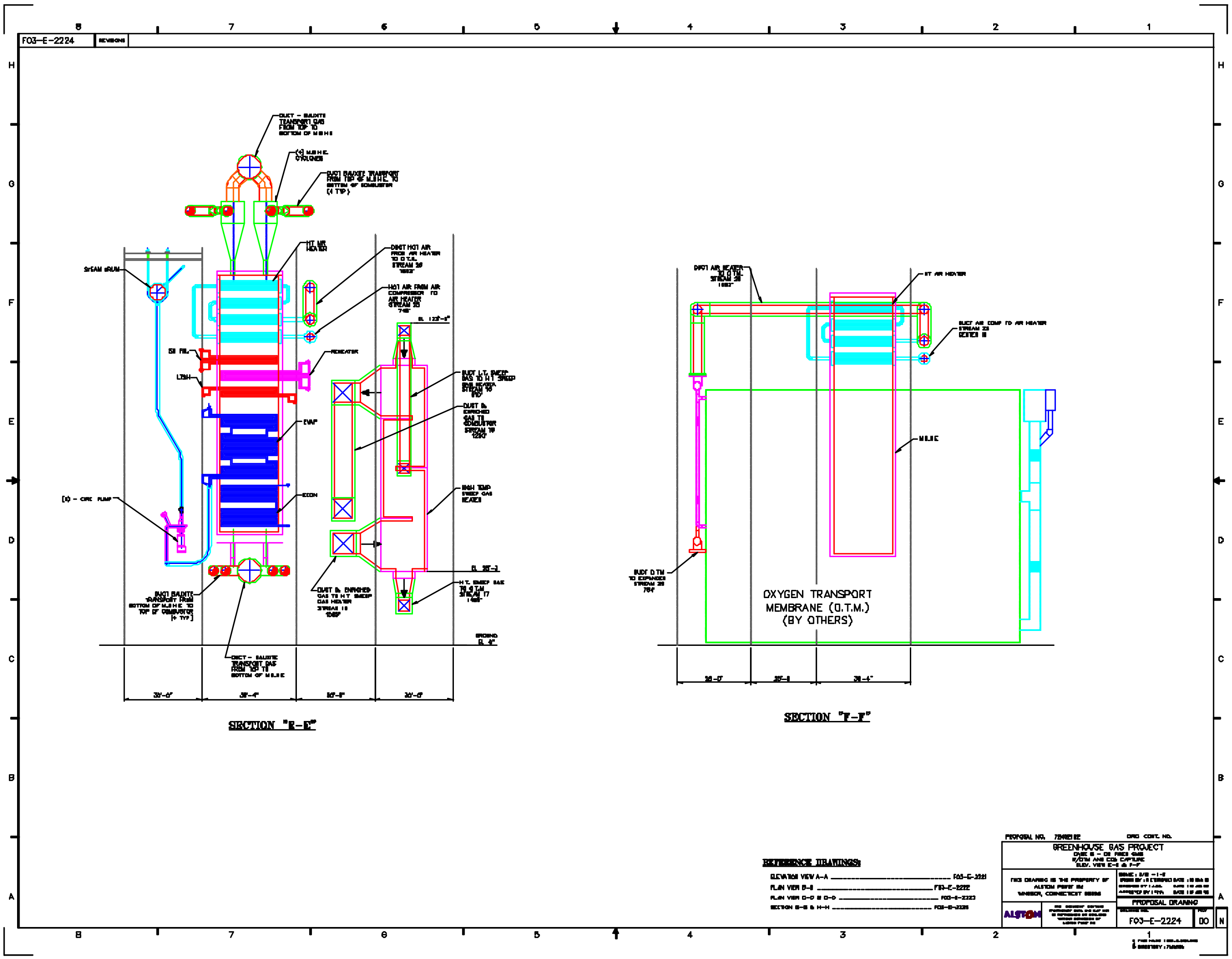
OXYGEN TRANSPORT
MEMBRANE (O.T.M.)
(BY OTHERS)
OXYGEN TRANSPORT
MEMBRANE (O.T.M.)
(BY OTHERS)
COAL SILO
COAL SILO
COAL SILO
LIMESTONE SILO
CYCLONE
SECTION "C-C"
SECTION "D-D"
REFERENCE DRAWINGS:
GREENHOUSE GAS PROJECT
PROPOSAL DRAWING
F03-E-2223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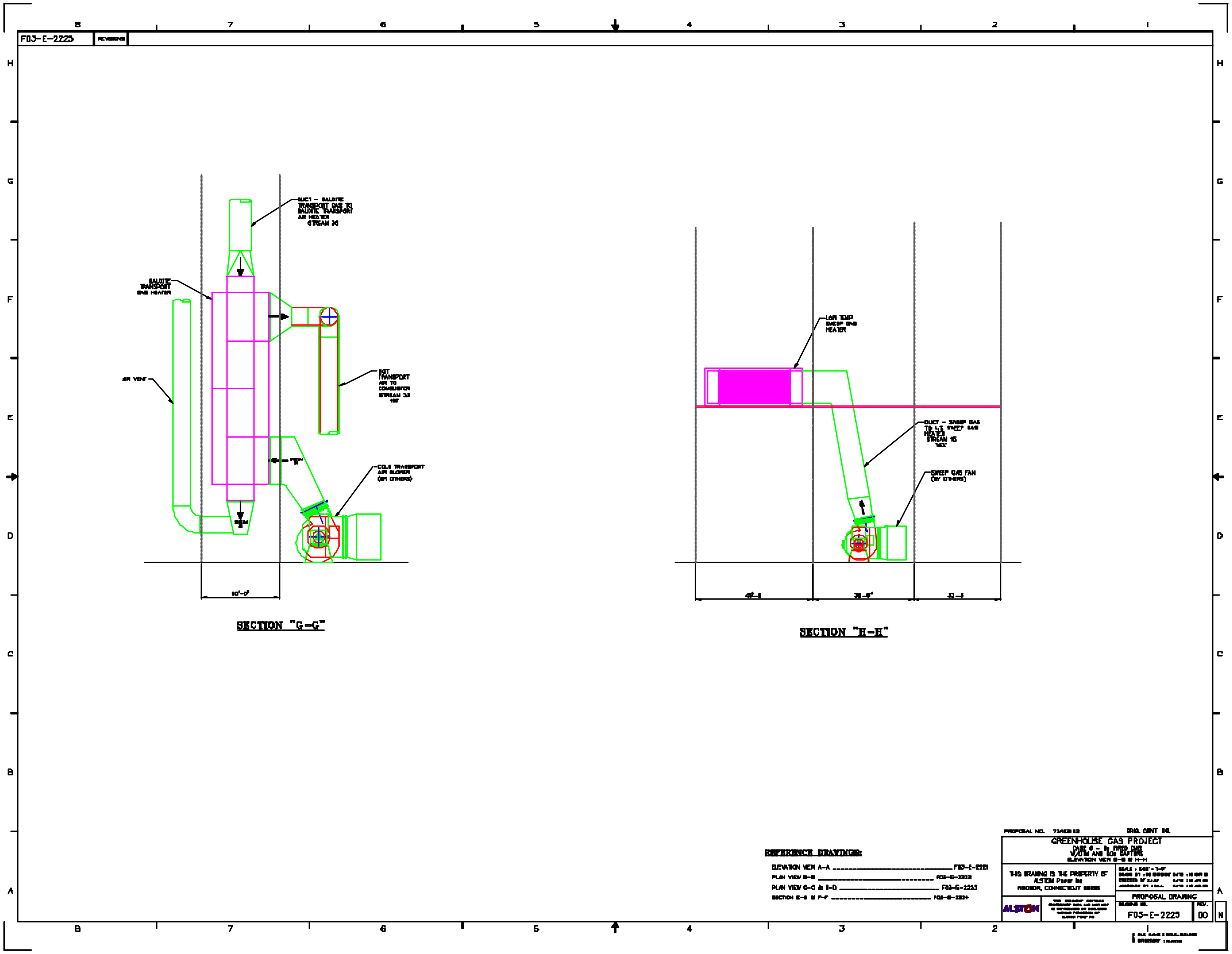
SECTION "E-E"
SECTION "F-F"
OXYGEN TRANSPORT
MEMBRANE (O.T.M.)
(BY OTHERS)
REFERENCE DRAWINGS
GREENHOUSE GAS PROJECT
PROPOSAL DRAWING
F03-E-2224


F03-E-2225
BAUXITE TRANSPORT GAS HEATER
AIR VENT
HOT TRANSPORT AIR TO COMBUSTOR
30'-0"
SECTION "G-G"
SWEEP GAS FAN (BY OTHERS)
SECTION "H-H"
REFERENCE DRAWINGS:
ELEVATION VIEW A-A F03-E-2221
PLAN VIEW B-B F03-E-2222
PLAN VIEW C-C & D-D F03-E-2223
SECTION E-E & F-F F03-E-2224
PROPOSAL NO. 7248202
GREENHOUSE GAS PROJECT
ELEVATION VIEW G-G & H-H
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WINDSOR, CONNECTICUT
ALSTOM
PROPOSAL DRAWING
DRAWING NO.
F03-E-2225
REV.
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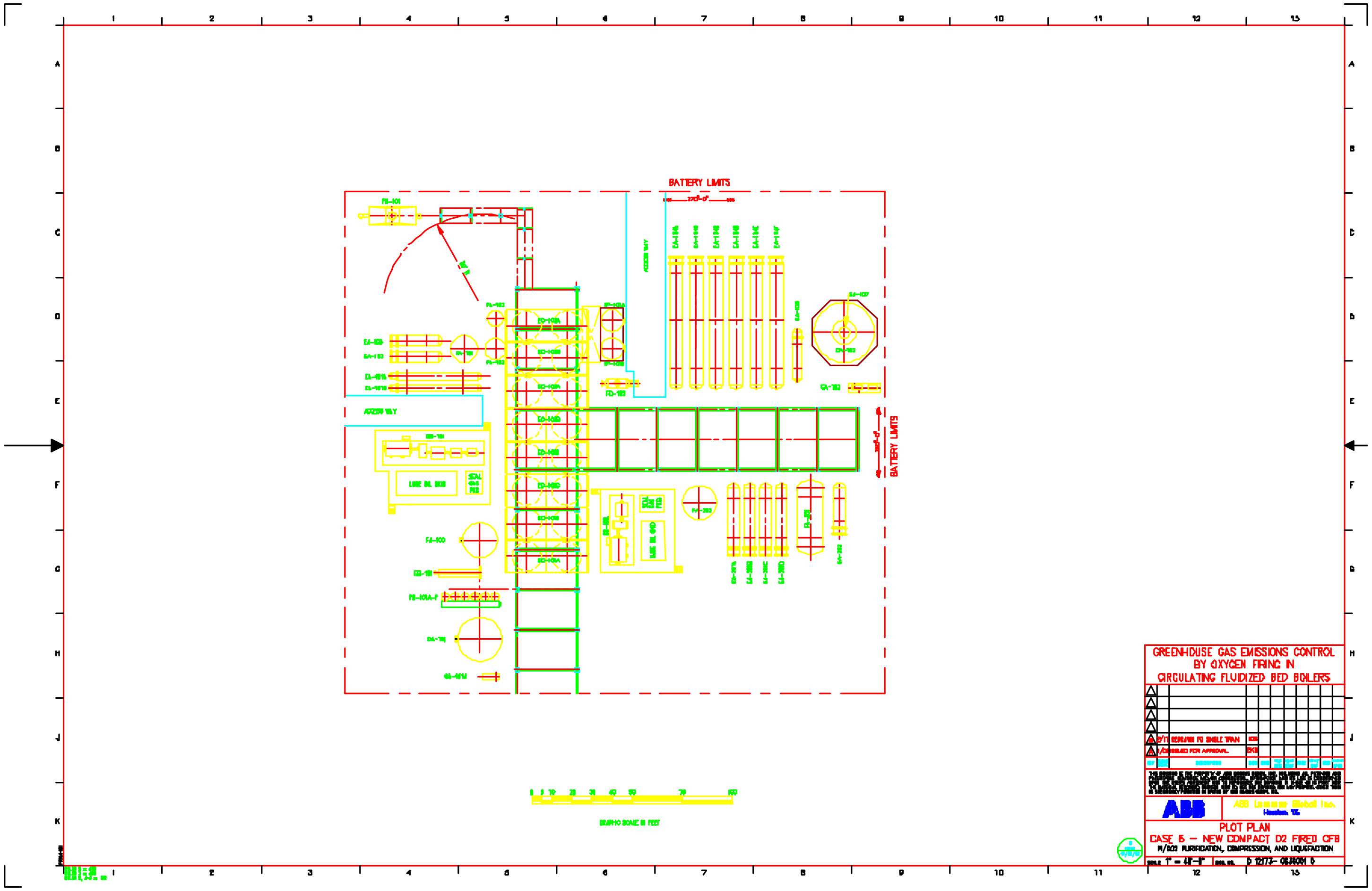
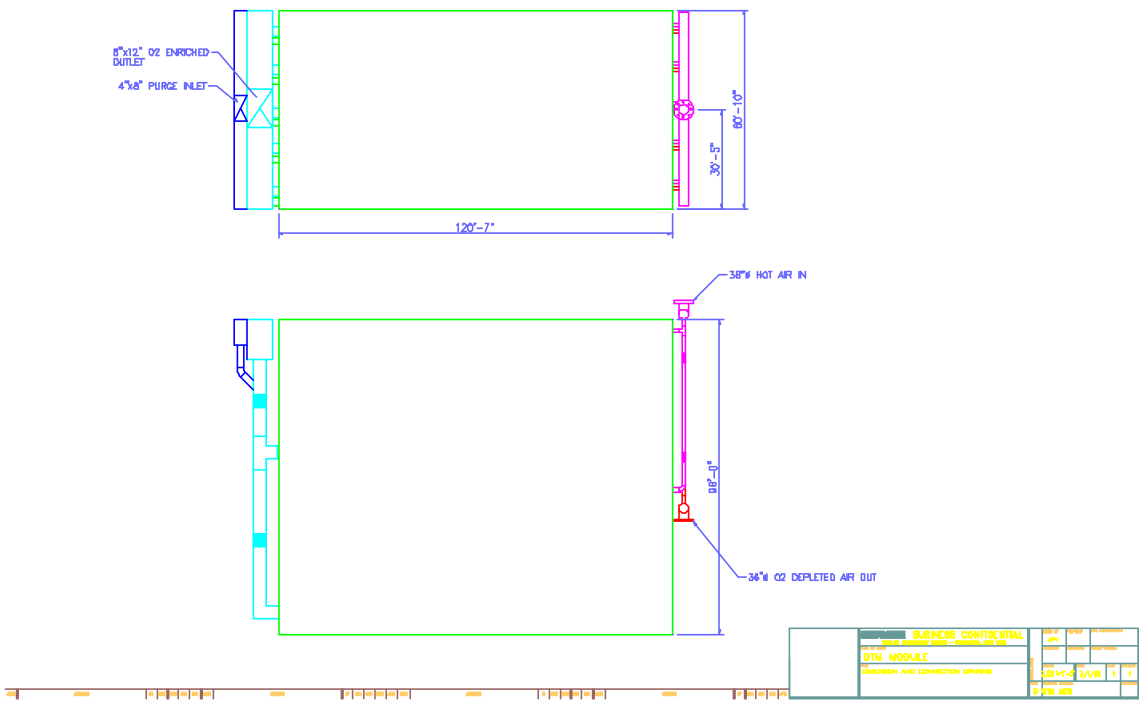
9.2.6.2. Case-6 Gas Processing System Equipment

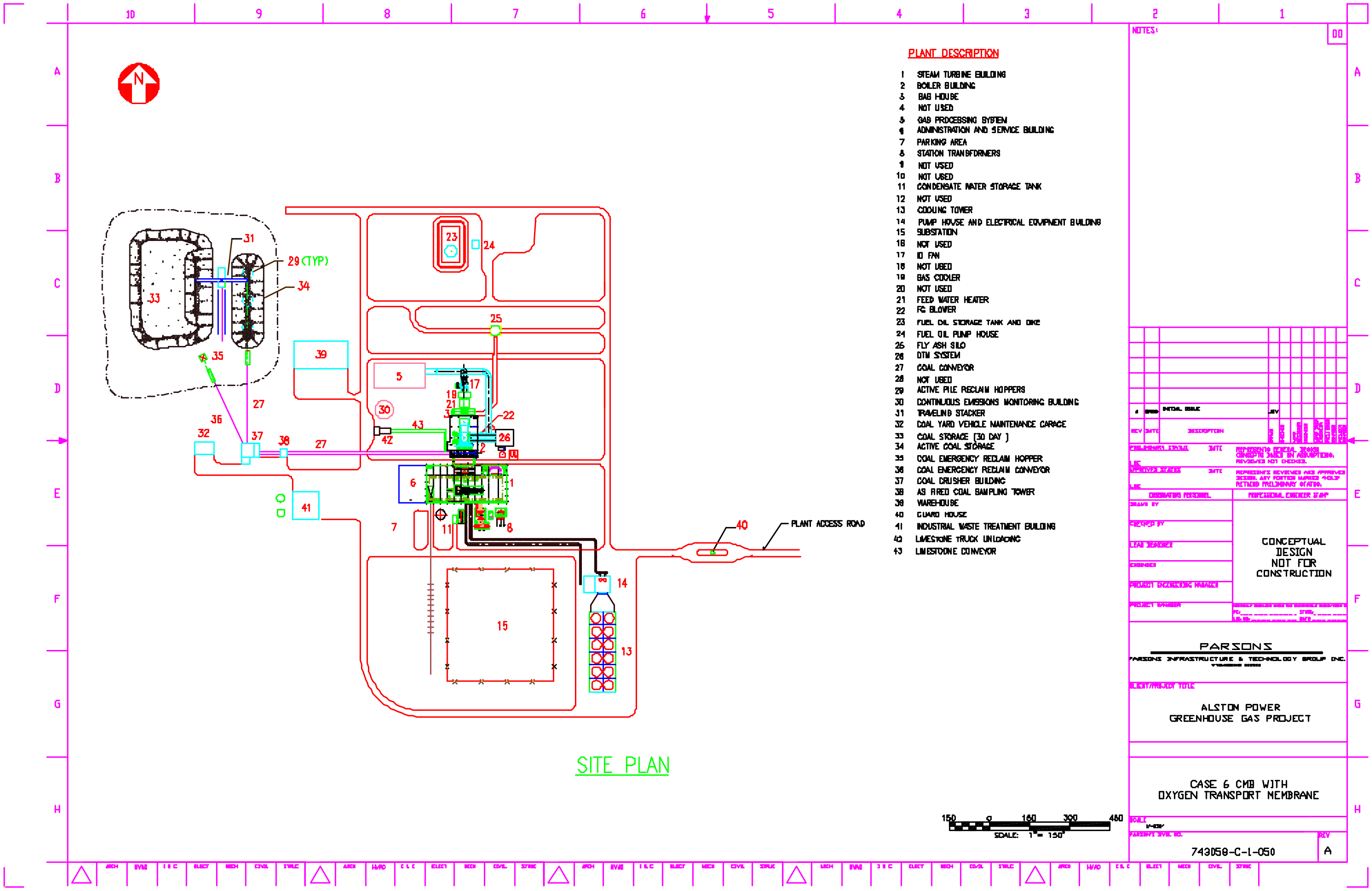


CO2 UNIT CASE 6.dwg LAST REVISED: 02/14/03; 13:03 by LSUTHERL

9.2.6.3. Case-6 Oxygen Transport Membrane System Equipment

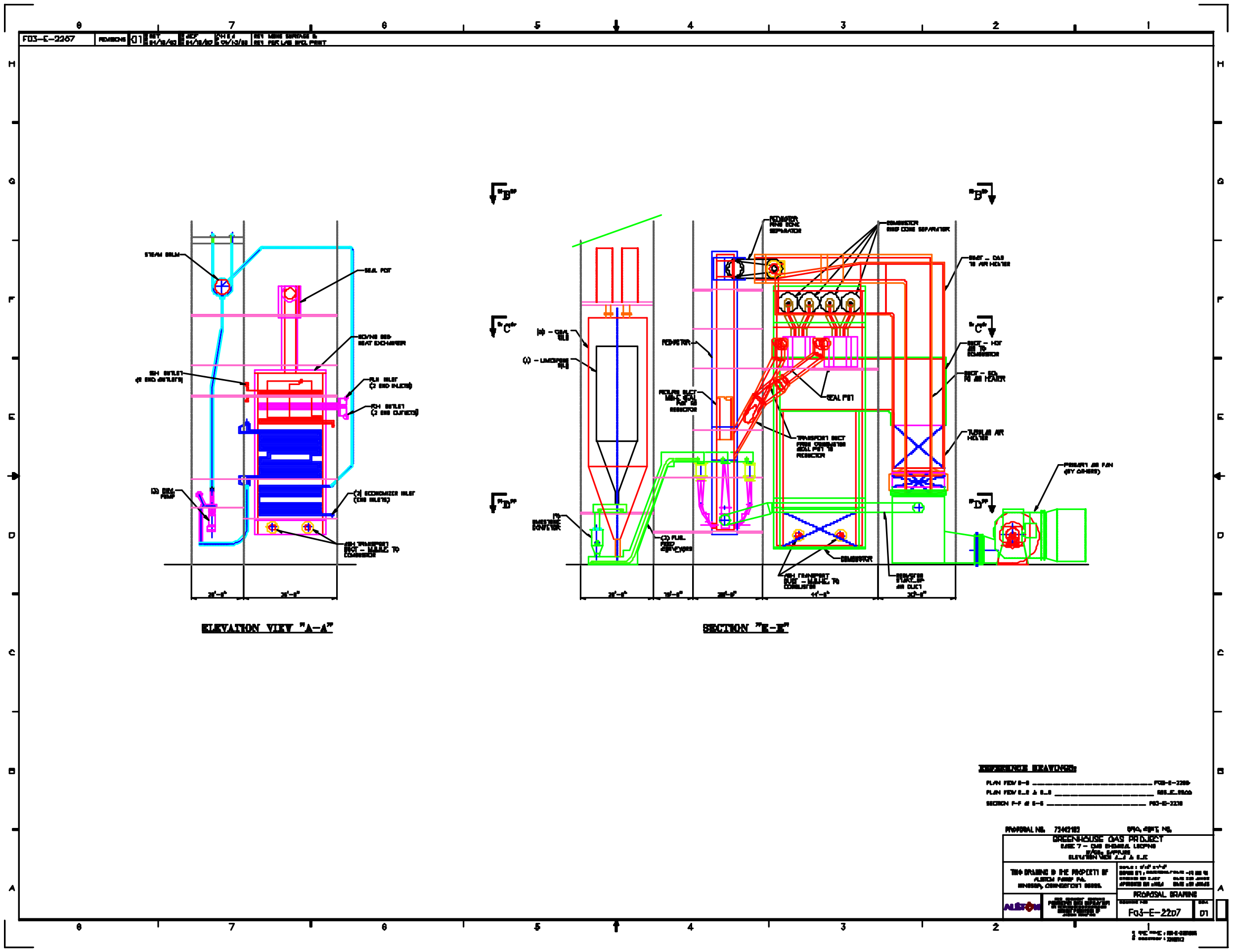


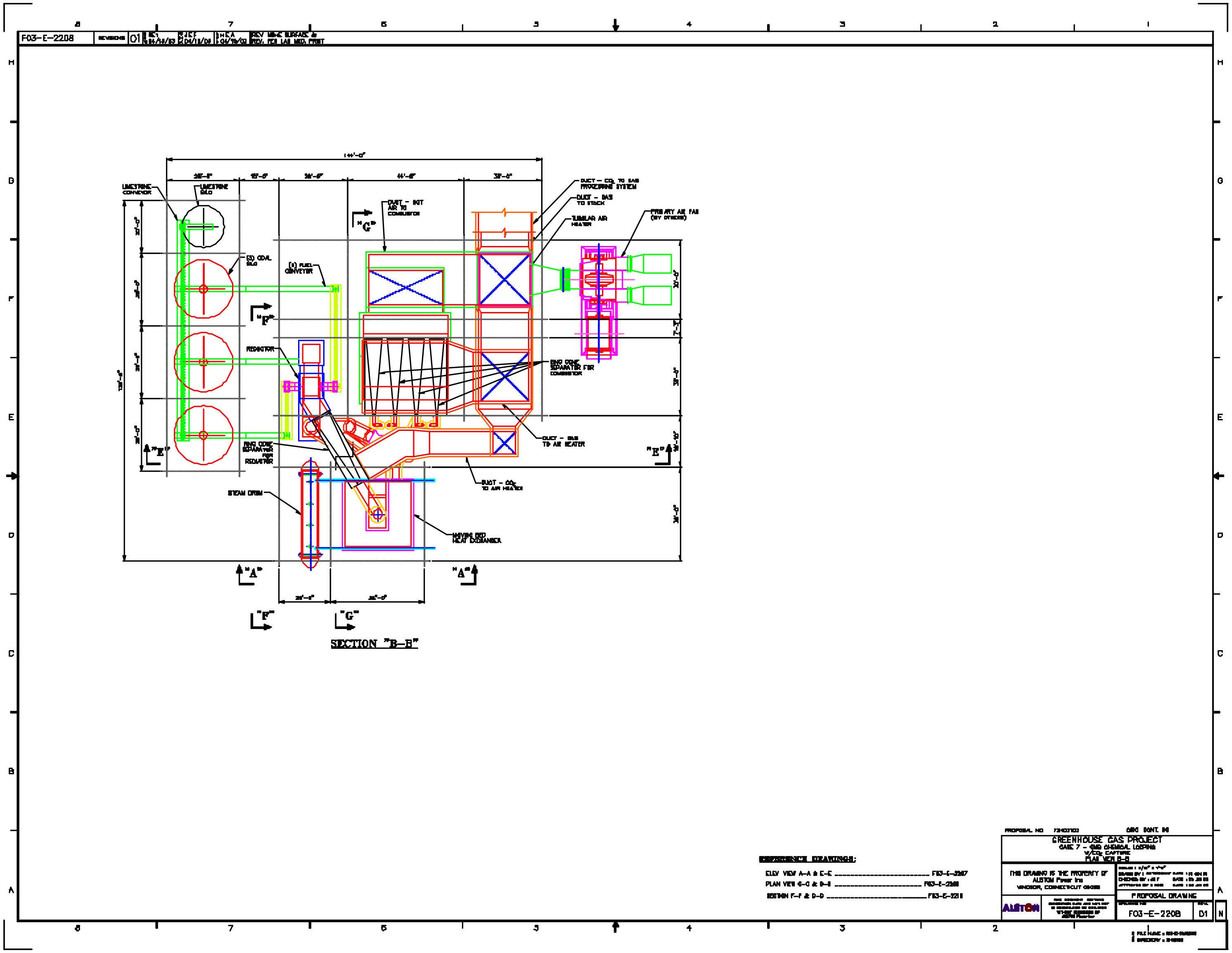
9.2.6.4. Case-6 Site Plan

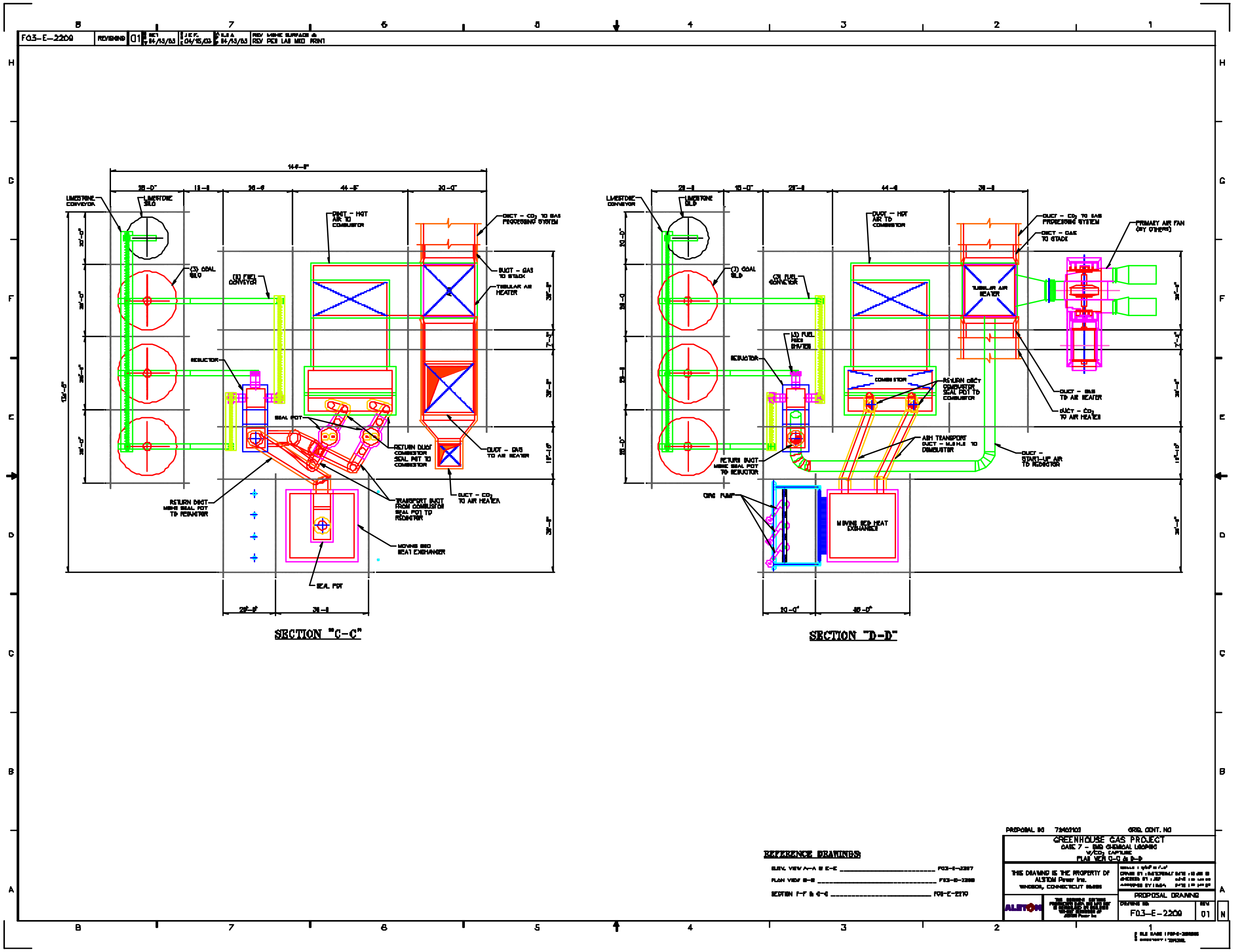


9.2.7. Case-7 Drawings

9.2.7.1. Case-7 Boiler Island Equipment



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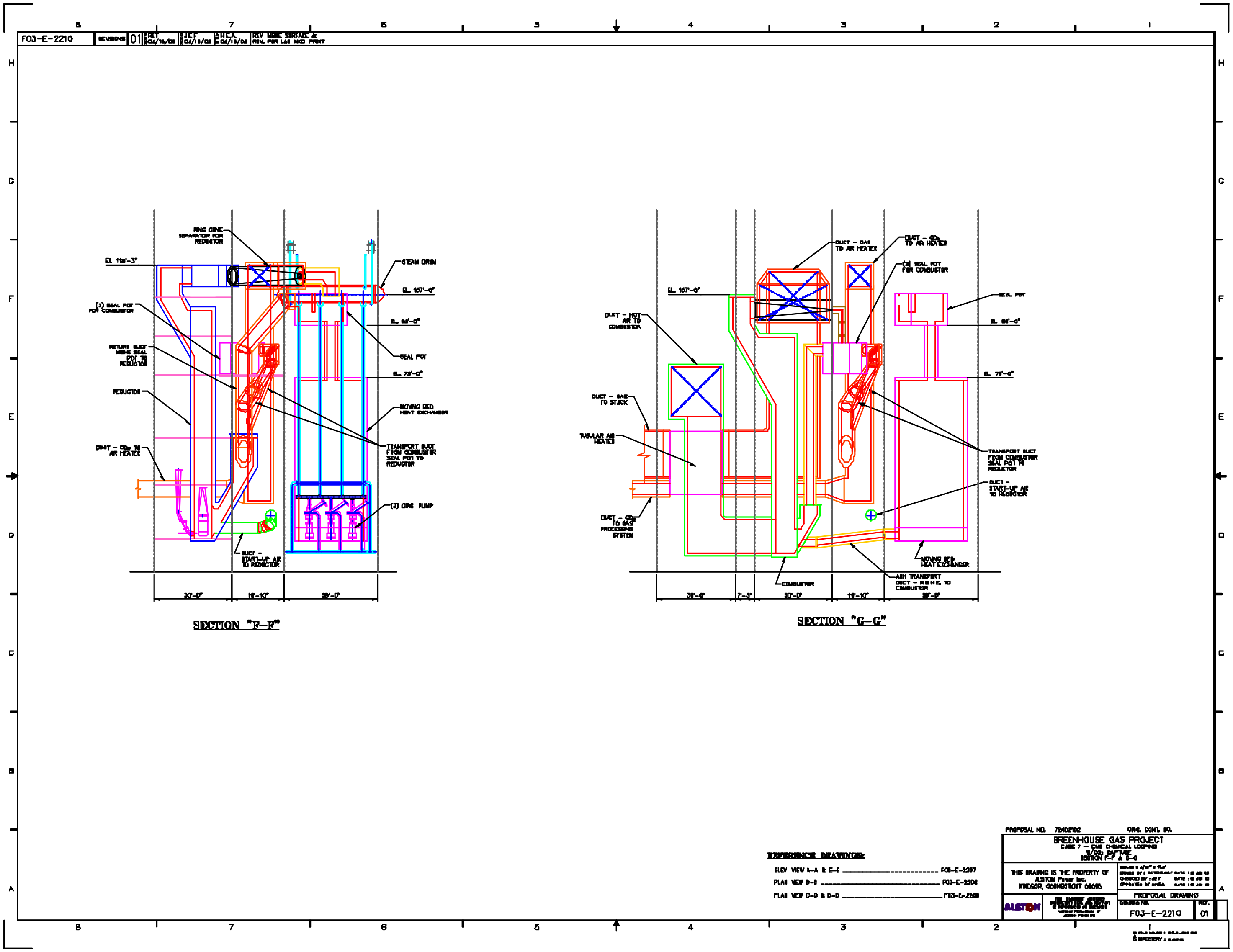
SECTION "D-D"
REFERENCE DRAWINGS
GREENHOUSE GAS PROJECT
PROPOSAL DRAWING
F03-E-2209


F03-E-2210
REVISIONS 01
RING CONE SEPARATOR FOR REDUCTOR
EL 118'-3"
STEAM DRUM
EL. 107'-0"
(2) SEAL POT FOR COMBUSTOR
EL. 95'-0"
SEAL POT
EL. 73'-0"
REDUCTOR
MOVING BED HEAT EXCHANGER
AIR HEATER
TRANSPORT DUCT FROM COMBUSTOR SEAL POT TO REDUCTOR
(3) CIRC. PUMP
DUCT - START-UP AIR TO REDUCTOR
30'-0"
18'-10"
30'-0"
SECTION "F-F"
DUCT - GAS TO AIR HEATER
EL. 107'-0"
DUCT - HOT AIR TO COMBUSTOR
(2) SEAL POT FOR COMBUSTOR
SEAL POT
EL. 95'-0"
EL. 73'-0"
DUCT - GAS TO STACK
TUBULAR AIR HEATER
TRANSPORT DUCT FROM COMBUSTOR SEAL POT TO REDUCTOR
DUCT - START-UP AIR TO REDUCTOR
PROCESSING SYSTEM
MOVING BED HEAT EXCHANGER
COMBUSTOR
ASH TRANSPORT DUCT - M.B.H.E. TO COMBUSTOR
7'-3"
30'-0"
18'-10"
SECTION "G-G"
REFERENCE DRAWINGS:
PLAN VIEW B-B
PLAN VIEW D-D & D-D
PROPOSAL NO. 724D2782
GREENHOUSE GAS PROJECT
CASE 7 — CMB CHEMICAL LOOPING
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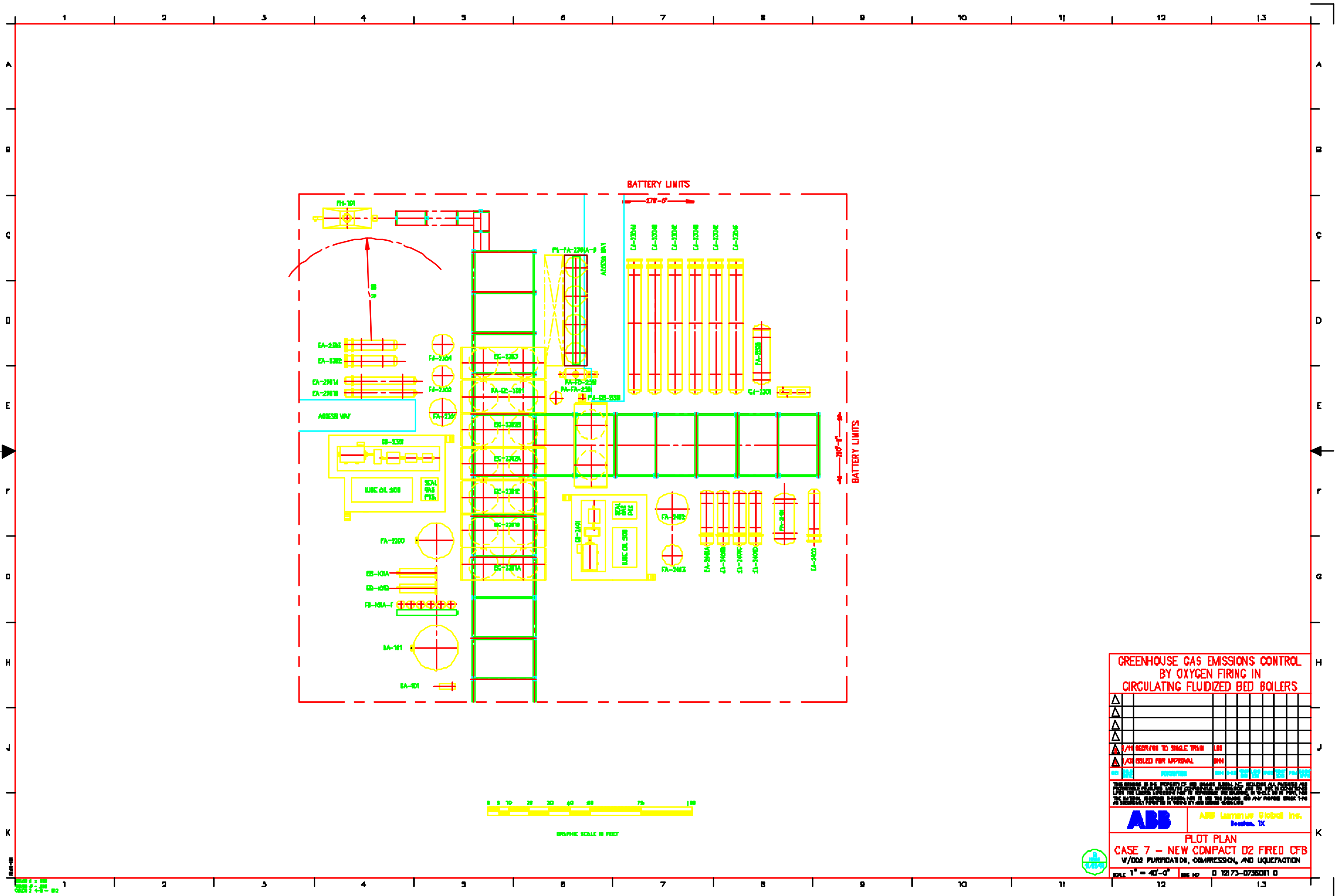
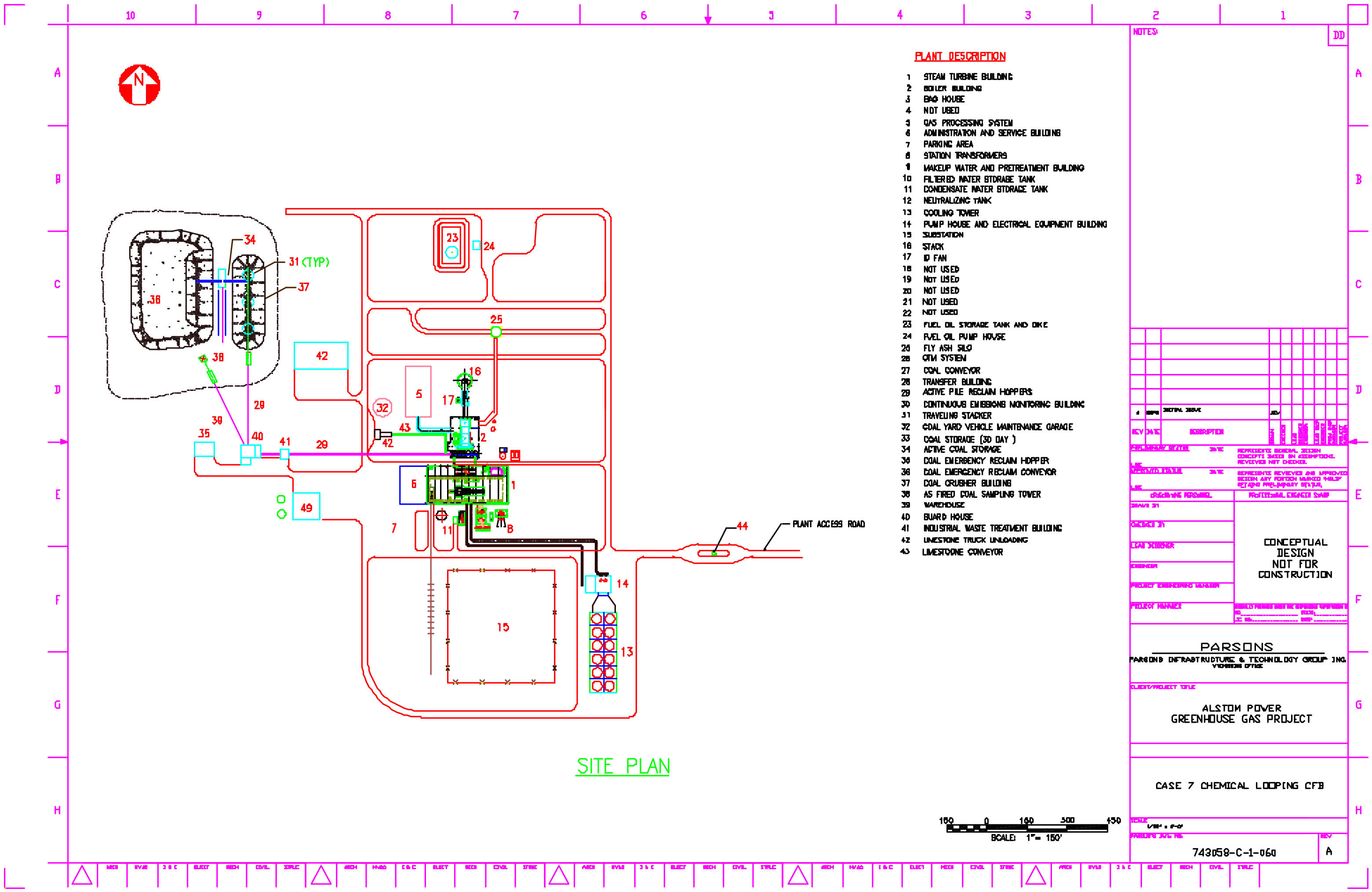
PROPOSAL DRAWING
DRAWING NO.
F03-E-2210
REV.
01

9.2.7.2. Case-7 Gas Processing System Equipment

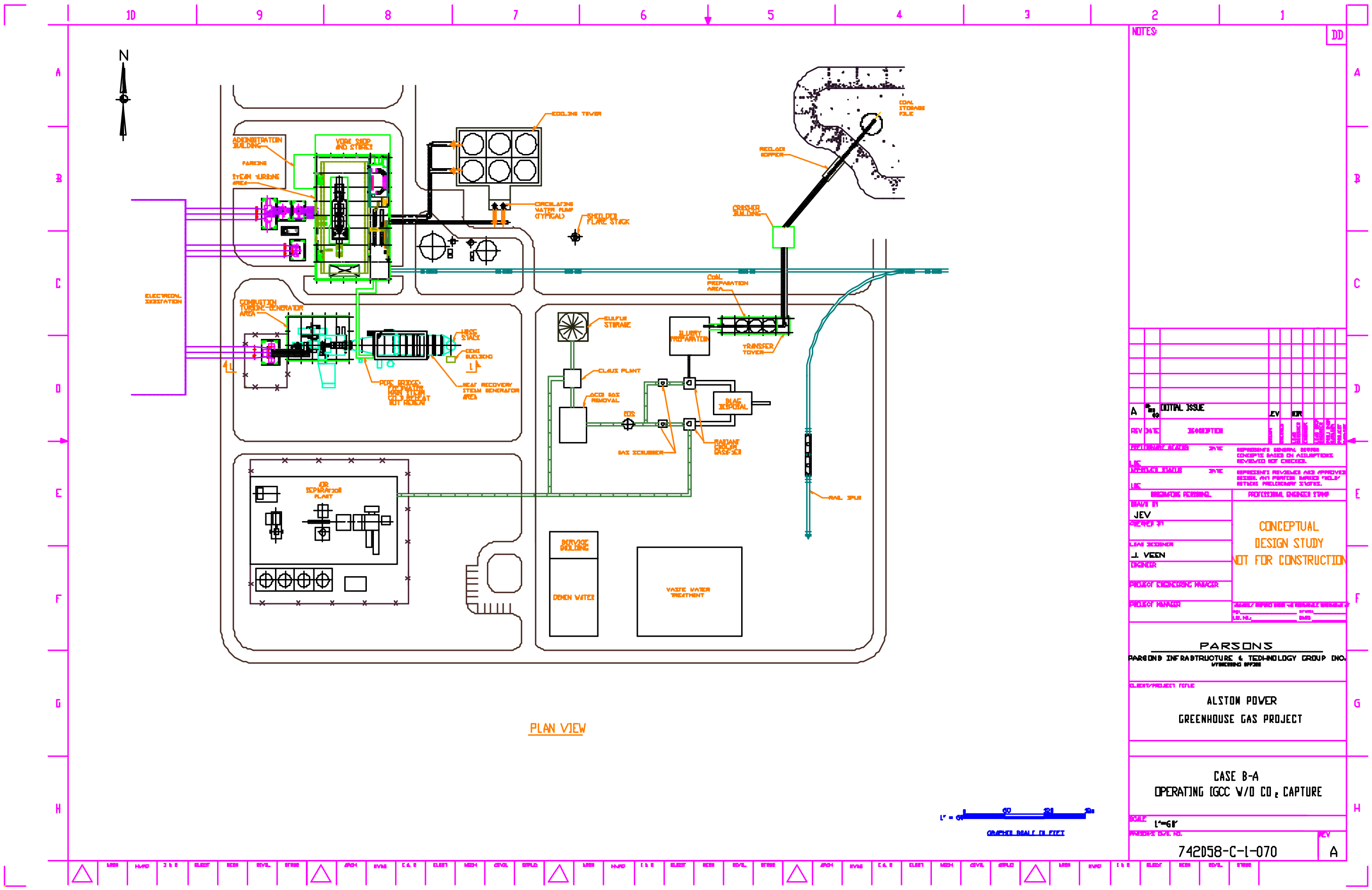


C02 UNIT CASE 7.dwg LAST REVISED: 02/10/03, 15:39 by LSUTHERL

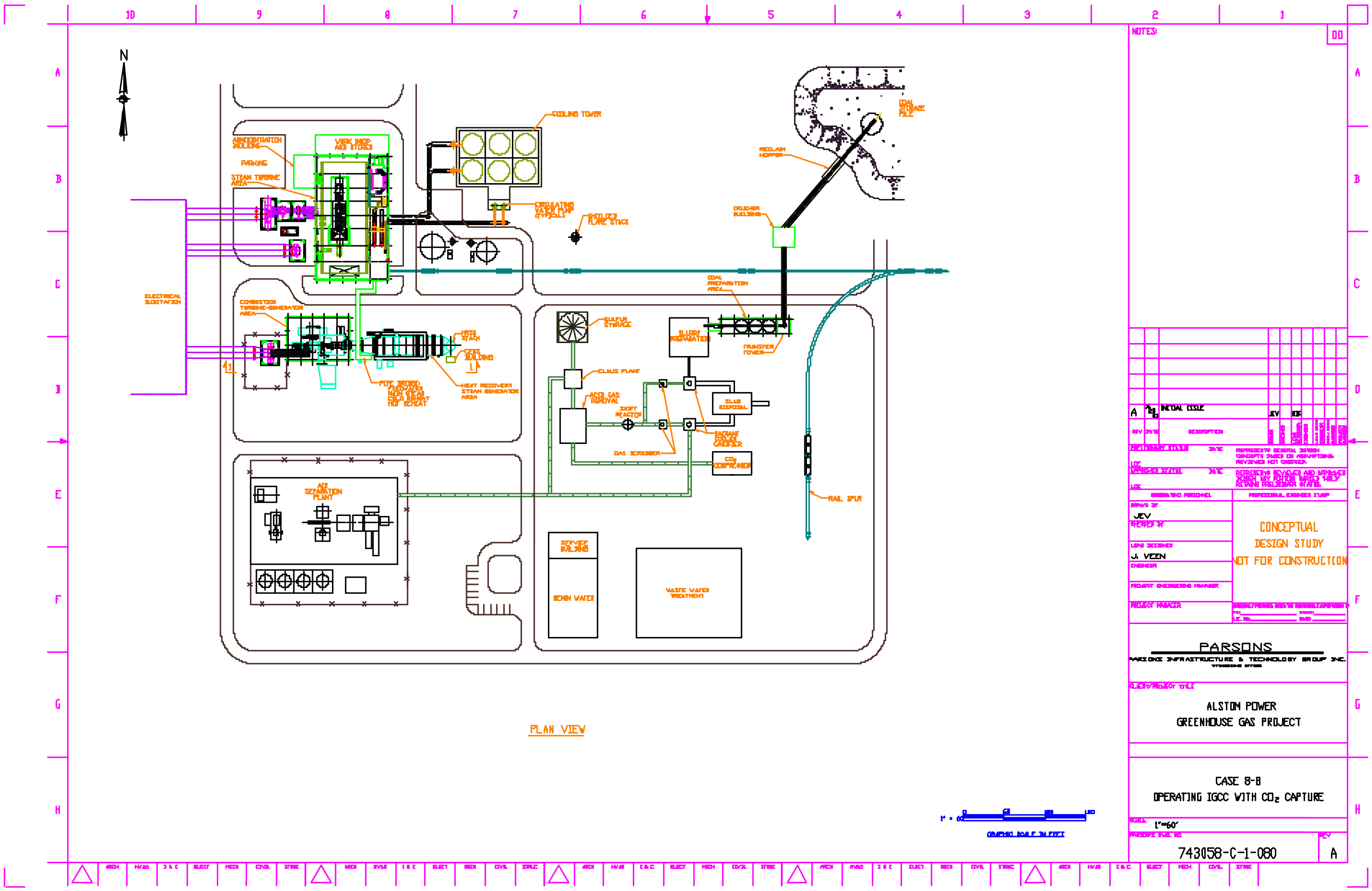
9.2.7.3. Case-7 Site Plan



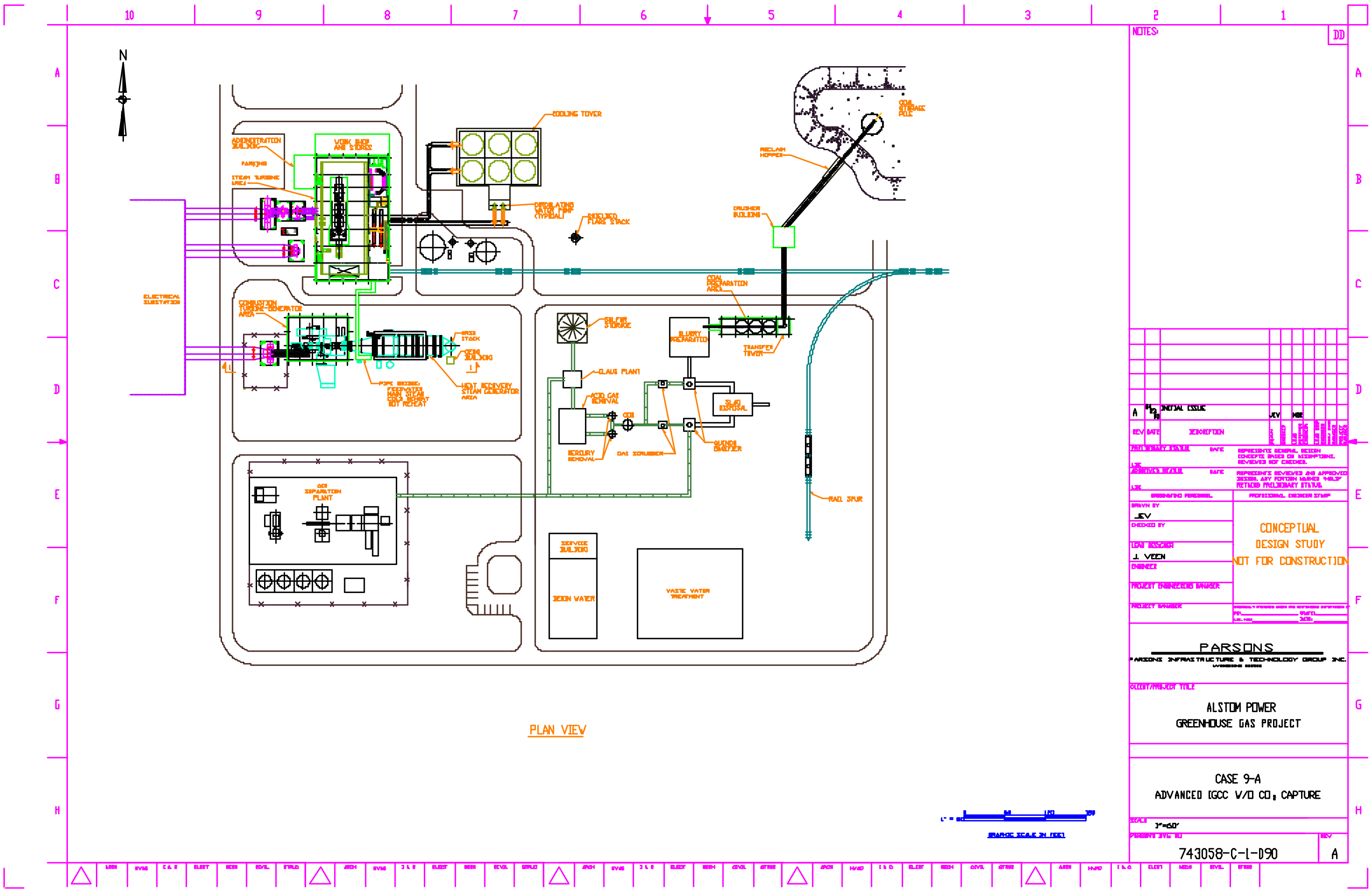
9.2.8. Case-8 Site Plan Drawing



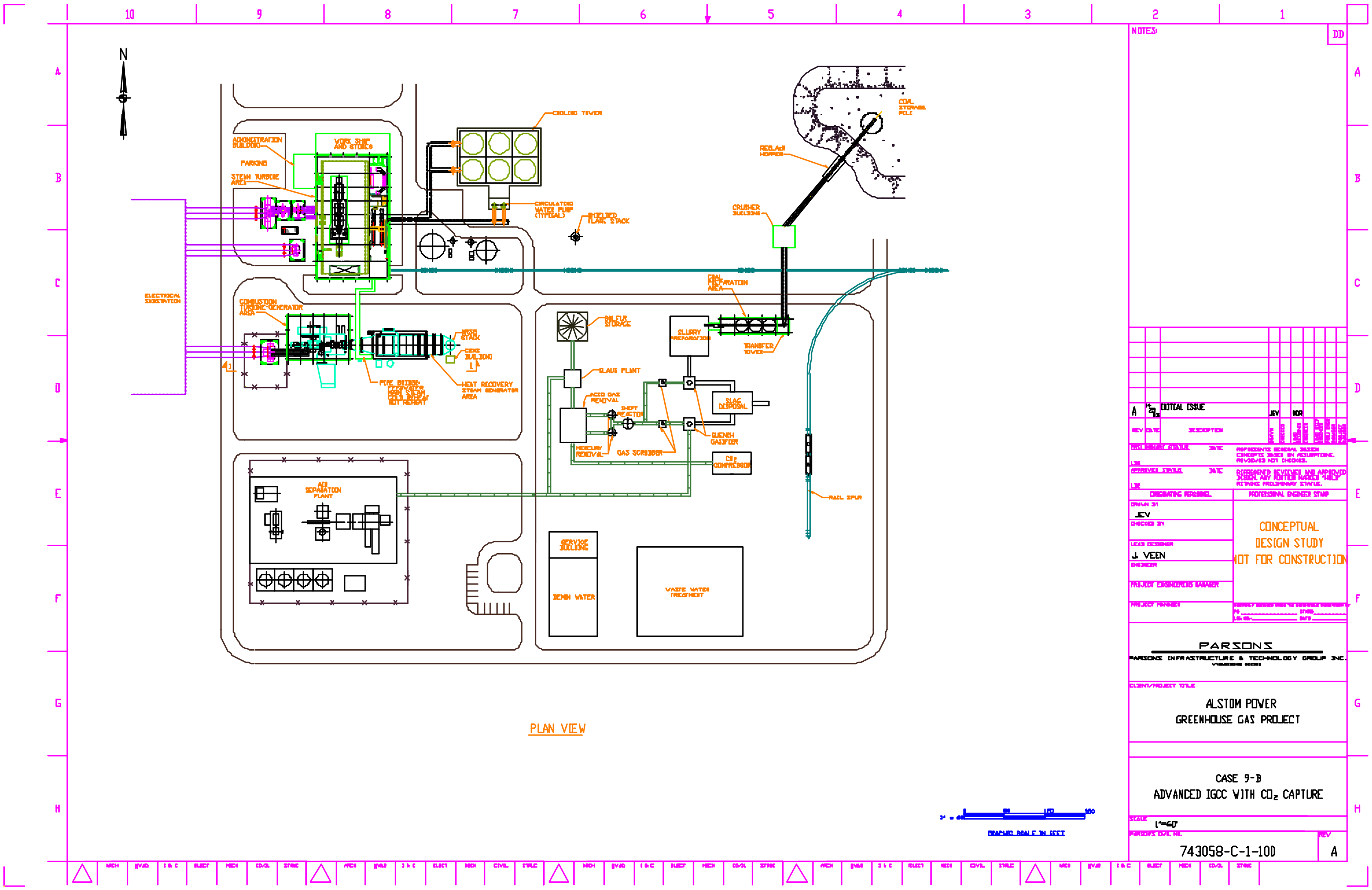
9.2.9. Case-9 Site Plan Drawing



9.2.10. Case-10 Site Plan Drawing



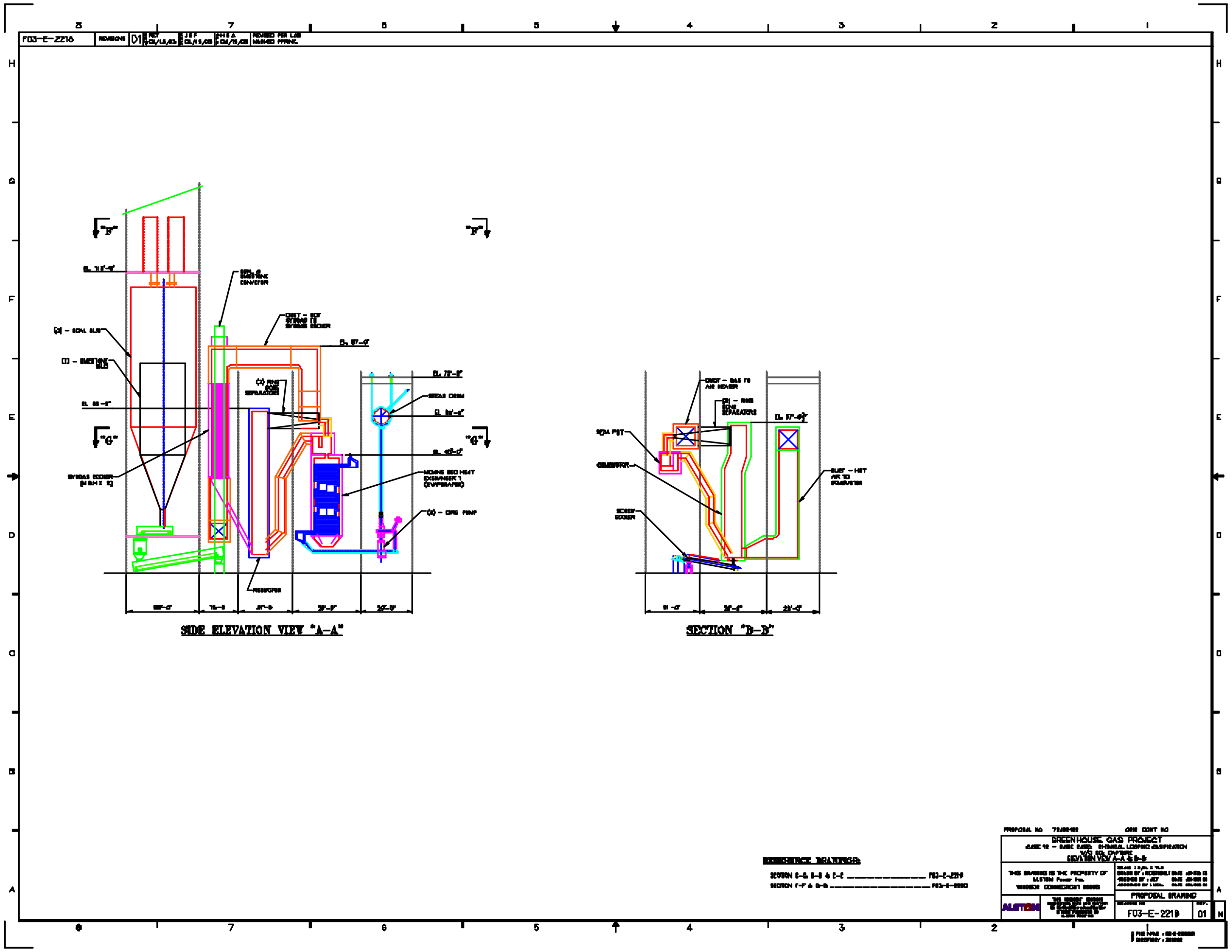
9.2.11. Case-11 Drawings

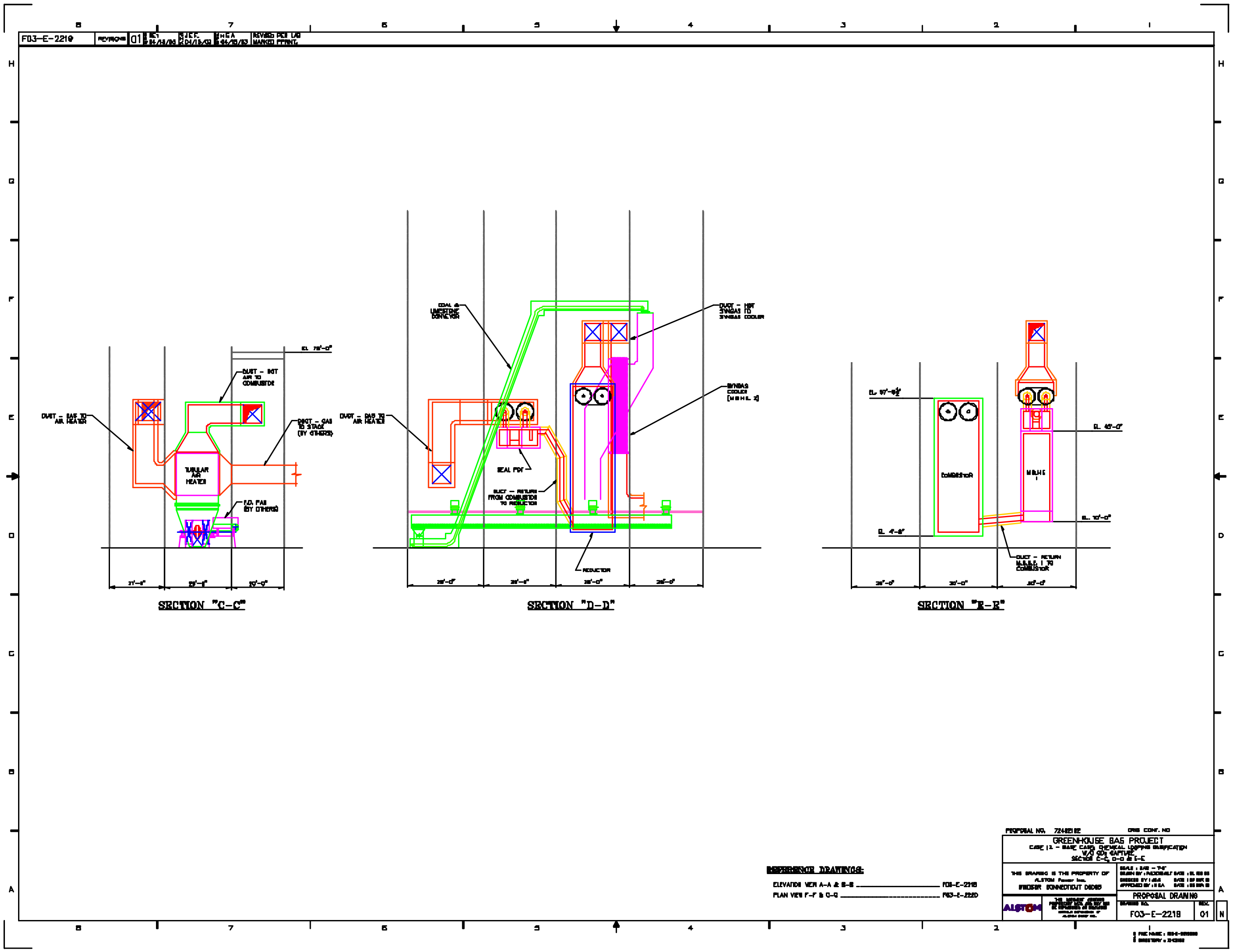


9.2.12.

Case-12 Drawings

9.2.12.1. Case-12 Gasifer Island Equipment

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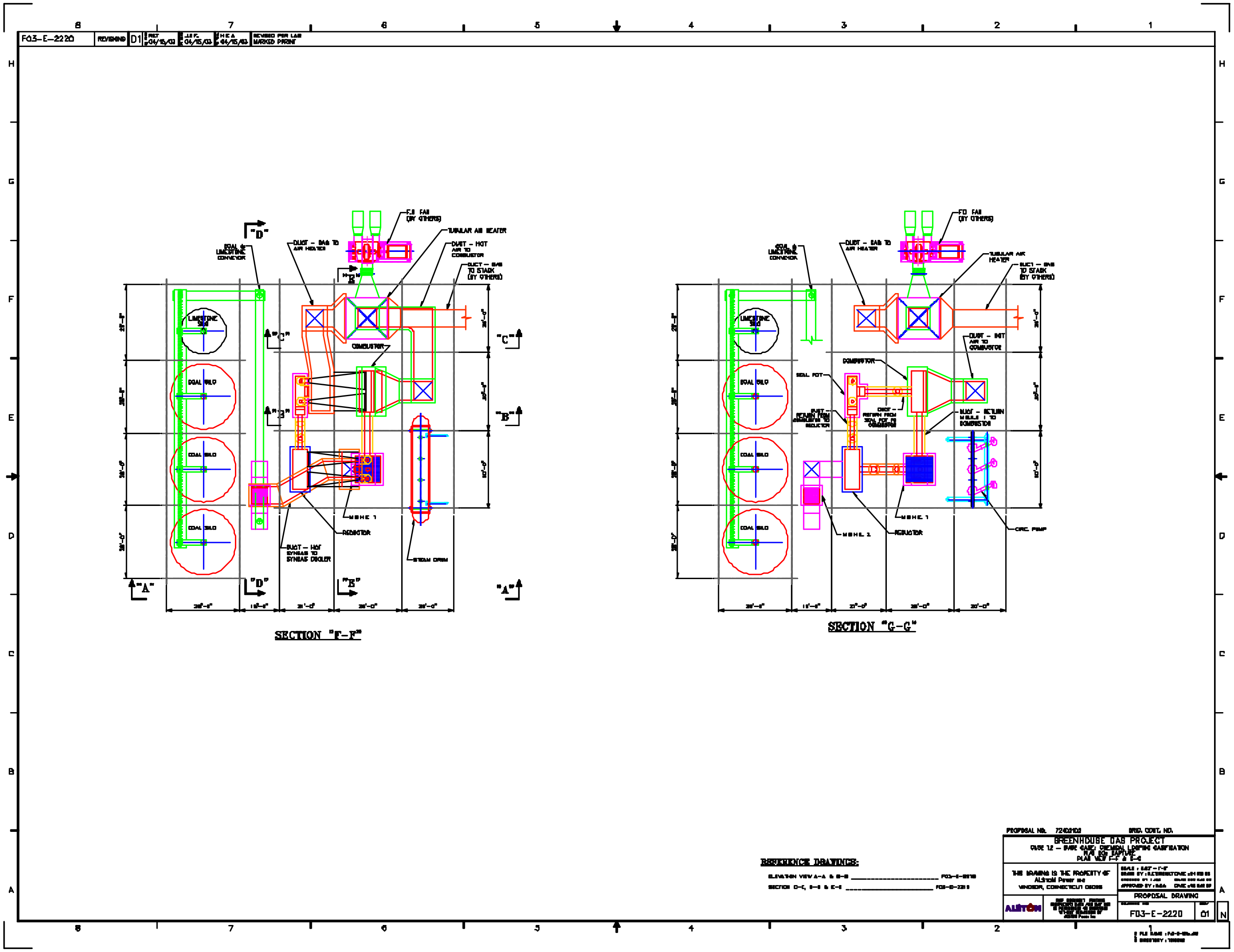
SECTION "D-D"
SECTION "E-E"
F03-E-2218
GREENHOUSE GAS PROJECT
PROPOSAL DRAWING
REFERENCE DRAWINGS: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SECTION "G-G"
REFERENCE DRAWINGS:
PROPOSAL NO. 724D21D2
ORIG. CONT. NO.
GREENHOUSE GAS PROJECT
PLAN VIEW F-F & G-G
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WINDSOR, CONNECTICUT
PROPOSAL DRAWING
F03-E-2220
01
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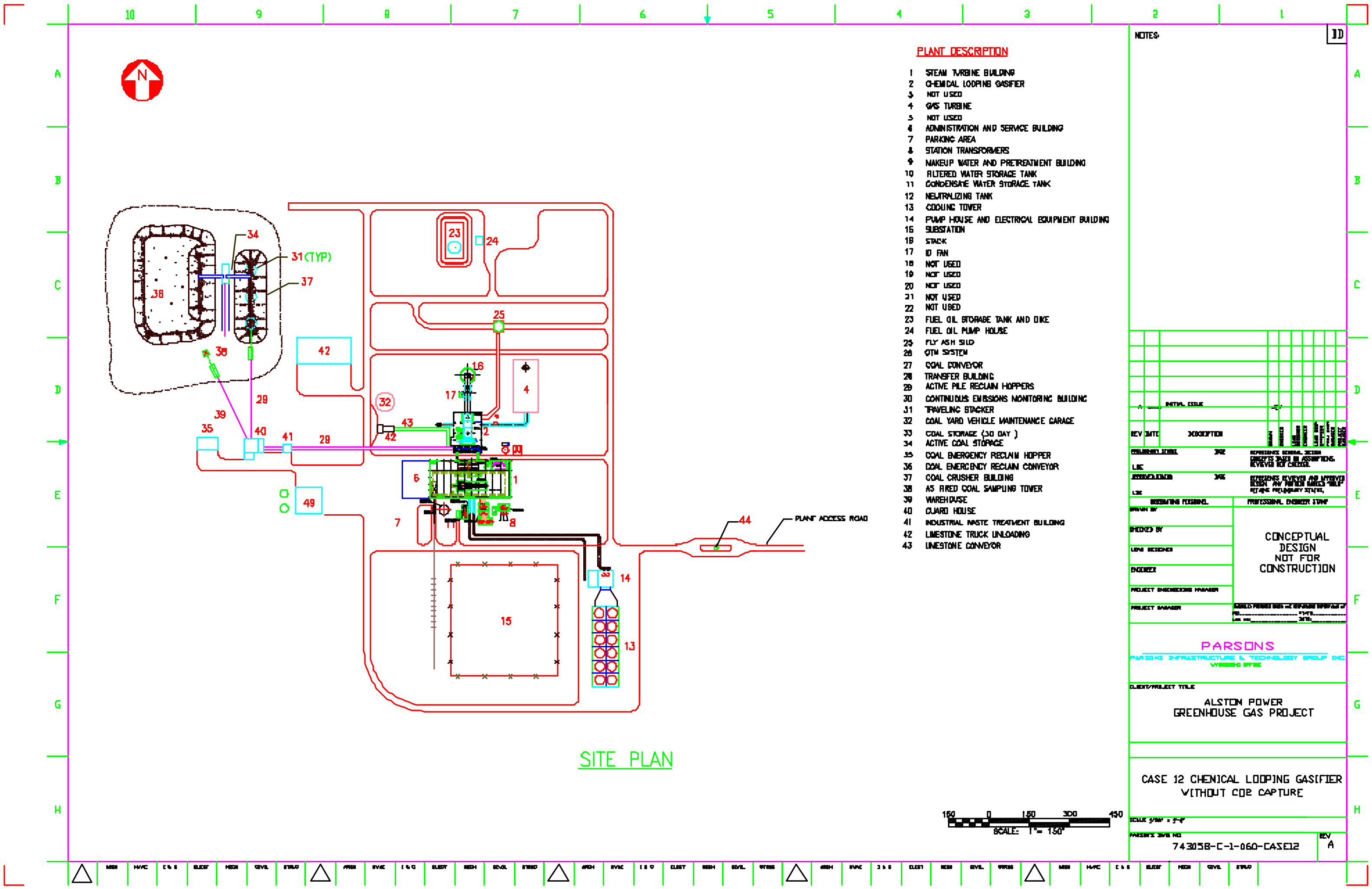
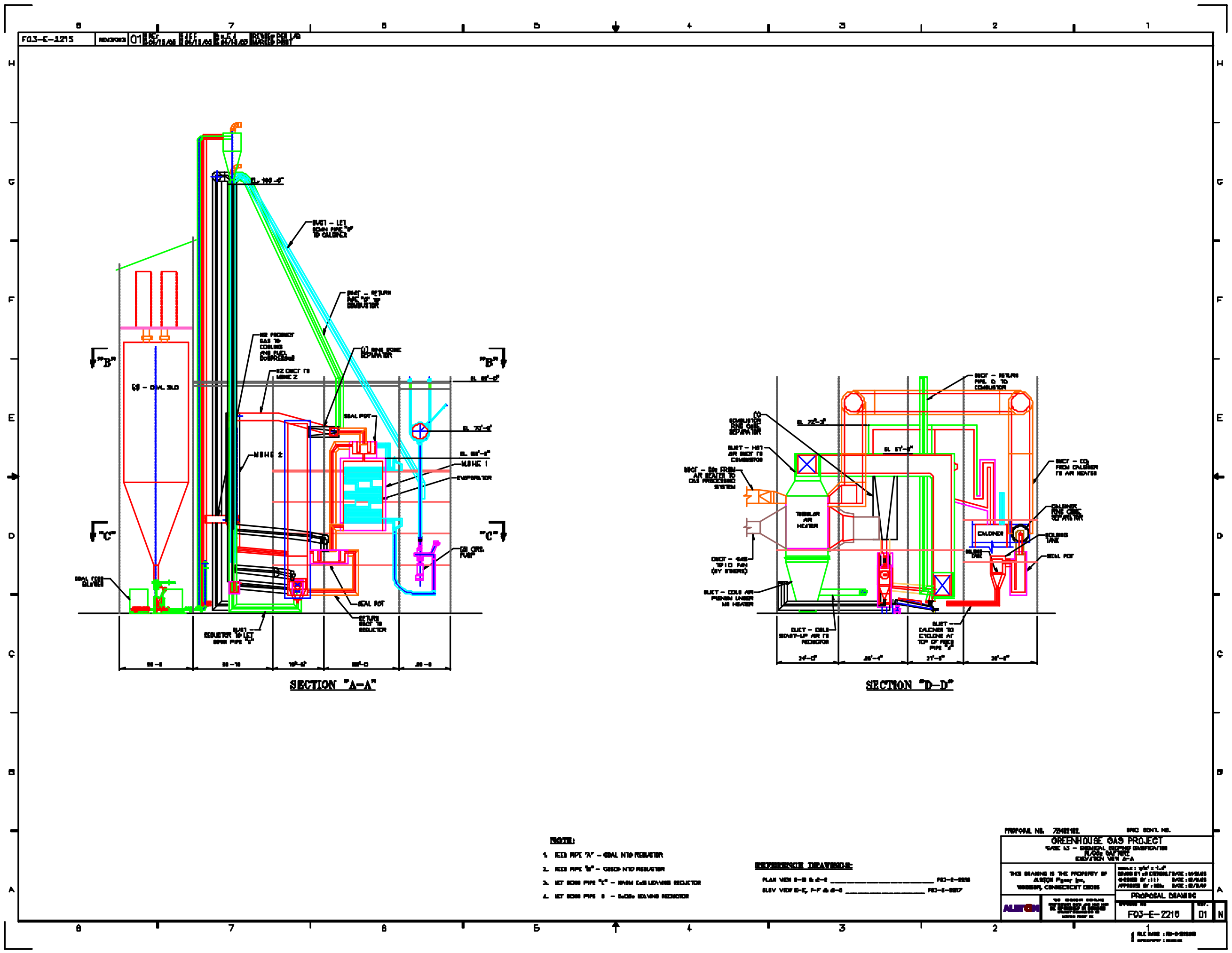
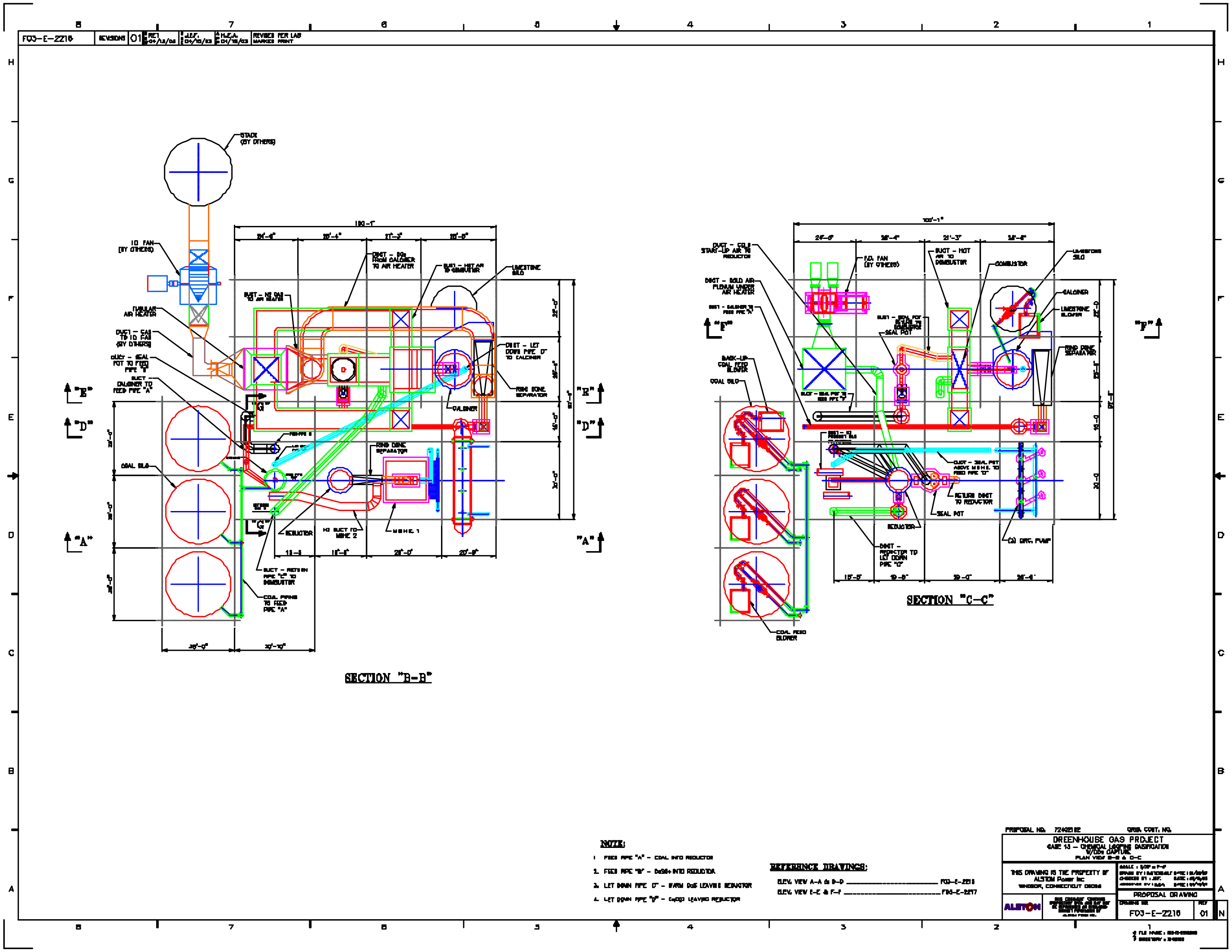
9.2.12.2. Case-12 Site Plan

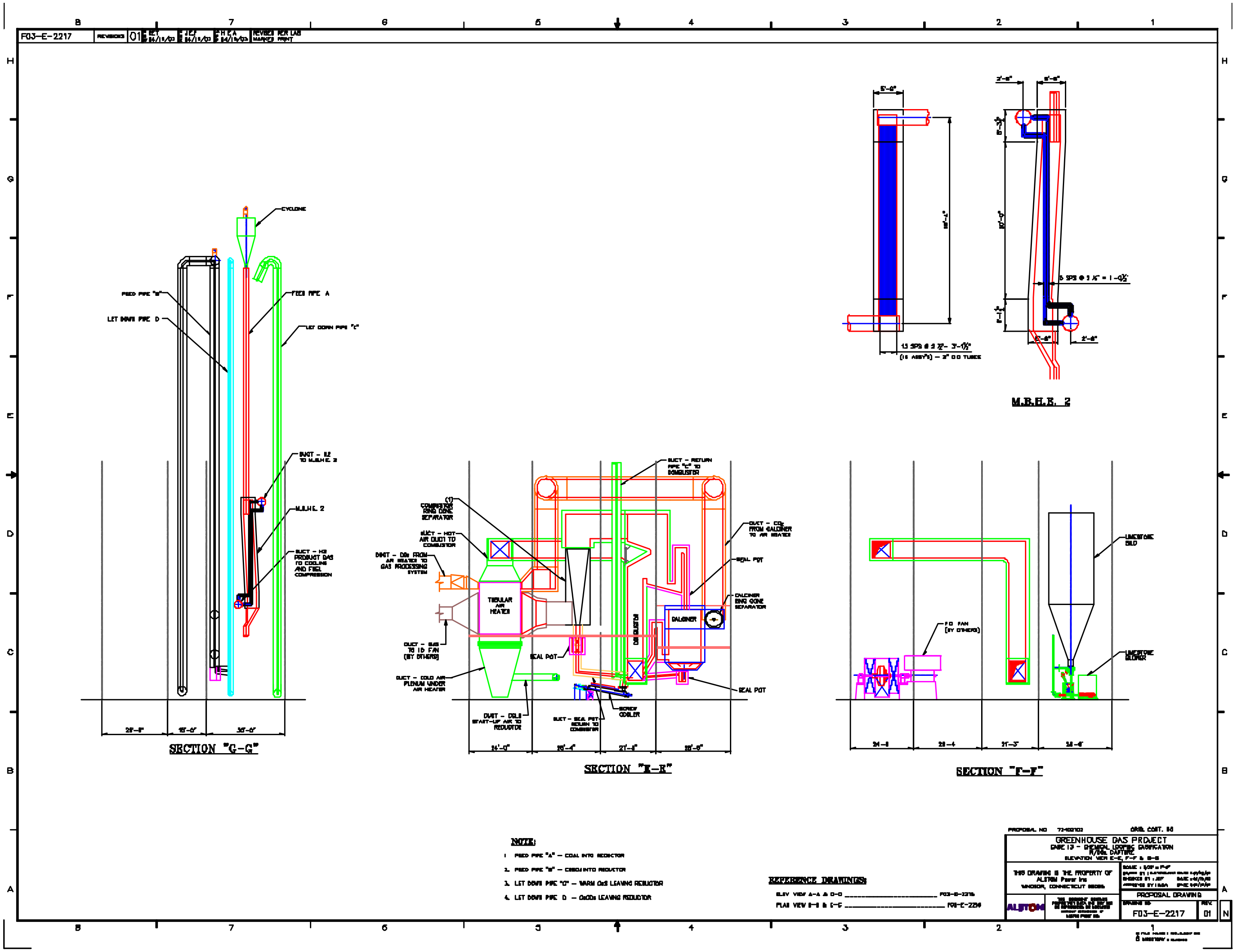


9.2.13. Case-13 Drawings

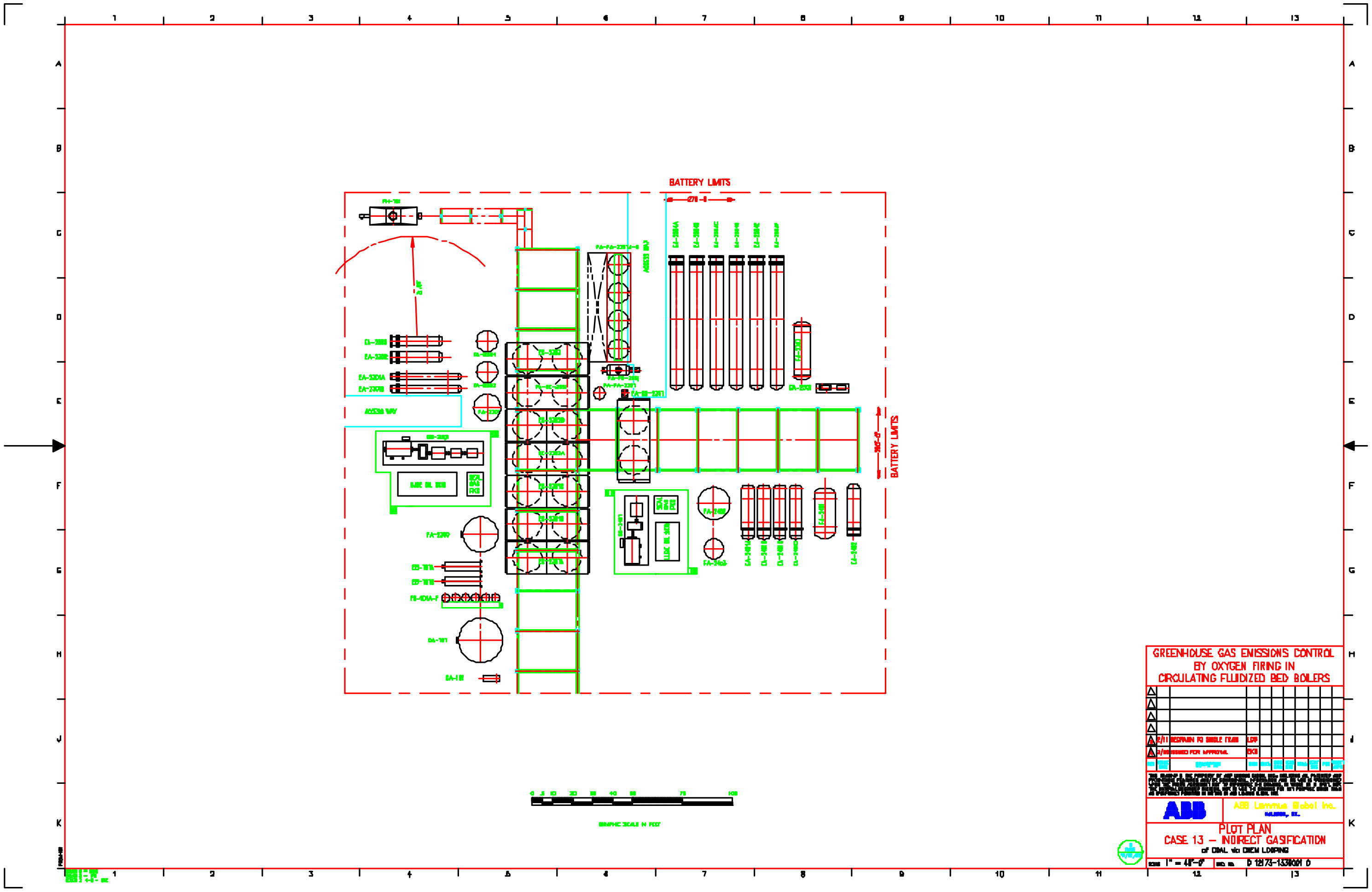
9.2.13.1. Case-13 Gasifer Island Equipment



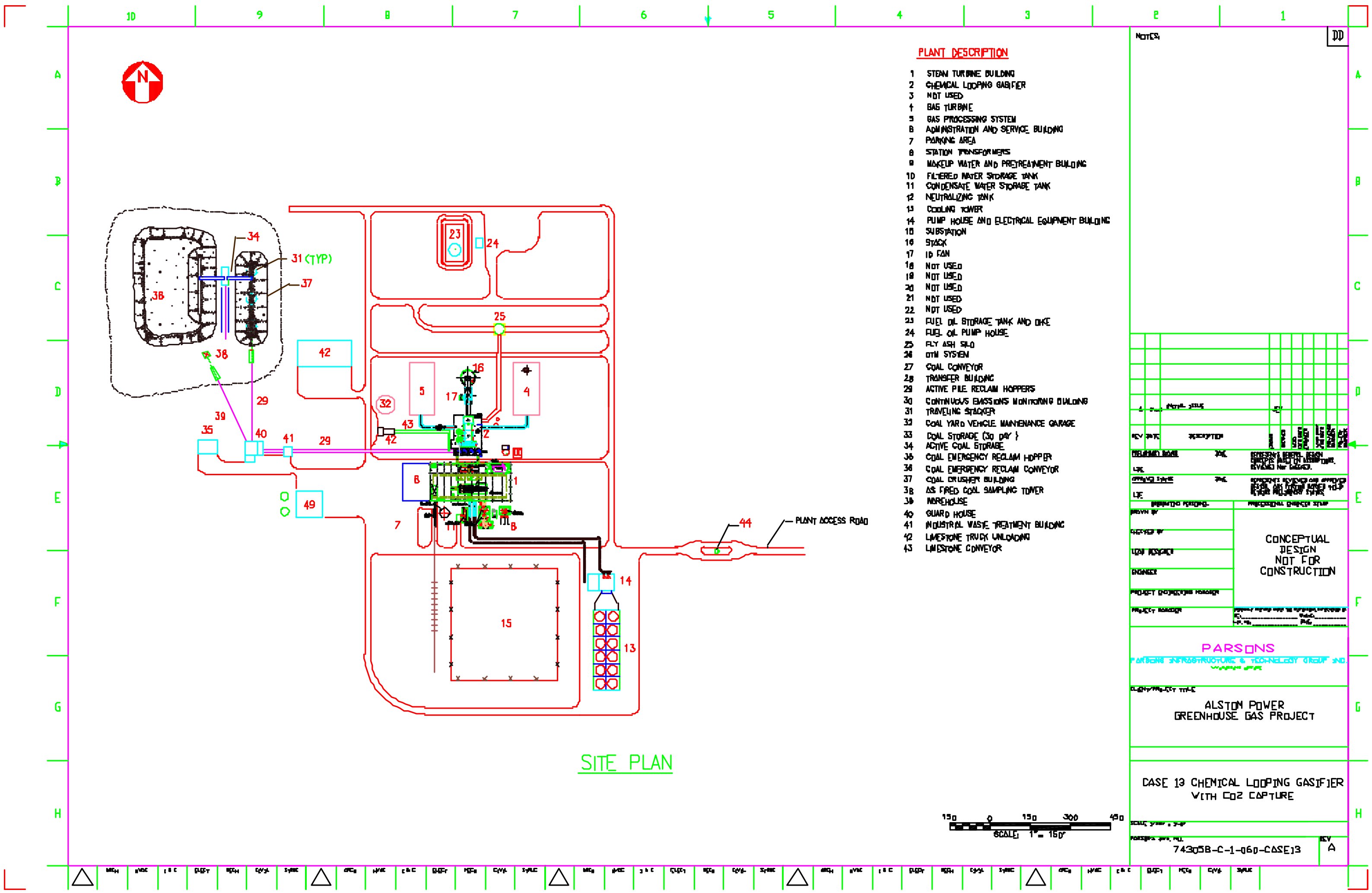

F03-E-2218
SECTION "B-B"
SECTION "C-C"
NOTE:
REFERENCE DRAWINGS:
GREENHOUSE GAS PROJECT
PROPOSAL DRAWING
F03-E-2218
ALSTOM



9.2.13.2. Case-13 Gas Processing System Equipment



9.2.13.3. Case-13 Site Plan



9.3. Appendix III – Detailed Investment Cost Breakdowns and Operating and Maintenance Costs

Appendix III provides detailed investment cost breakdowns and operating & maintenance costs for each of the thirteen plants studied. The costs are presented consecutively, starting with Case-1 and ending with Case-13. The costs are grouped into four separate areas: Boiler Island, Air Separation, Gas Processing, and Balance of Plant. Some of the cases do not have equipment in all four areas.

9.3.1. Case-1 Investment Costs and Operating and Maintenance Costs

Table 9.3. 1: Case-1 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc. **Report Date:** 4/18/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture

Net Output Power, kW 193,037 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost ($x1000)	Total Cost $/kW
1	FUEL & SORBENT HANDLING	6,760	1,657	3,765	-	-	12,182	1,463	548	14,193	74
2	FUEL & SORBENT PREP. & FEED	3,588	202	1,039	-	-	4,829	580	217	5,626	29
3	FEEDWATER & MISC. BOP SYSTEMS	11,721	-	5,757	-	-	17,478	2,097	786	20,361	105
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									75,658	392
4.2	Open										
4.4	Open										
4.4 - 4.9	Boiler BOP (Fluidizing Air Fans)	2,502	-	718	-	-	3,220	386	145	3,751	19
	SUBTOTAL 4	2,502	-	718	-	-	3,220	386	145	79,409	411
5	FLUE GAS CLEANUP	3,873	-	3,664	-	-	7,537	905	339	8,781	45
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	-	-	-	0	0	-	-	-	-	0
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	6,851	415	4,155	-	0	11,421	1,370	515	13,306	69
	SUBTOTAL 7	6,851	415	4,155	-	-	11,421	1,370	515	13,306	69
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	16,242	-	2,330	0	0	18,572	2,229	836	21,637	112
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,706	324	4,719	0	0	13,749	1,649	619	16,017	83
	SUBTOTAL 8	24,948	324	7,049	-	-	32,321	3,878	1,455	37,654	195
9	COOLING WATER SYSTEM	3,767	2,146	4,246	-	-	10,159	1,219	457	11,835	61
10	ASH/SPENT SORBENT HANDLING SYSTEMS	4,499	157	3,204	-	-	7,860	943	354	9,157	47
11	ACCESSORY ELECTRIC PLANT	5,545	2,366	7,661	-	-	15,572	1,869	700	18,141	94
12	INSTRUMENTATION & CONTROL	4,731	-	4,837	-	-	9,568	1,047	430	11,045	57
13	IMPROVEMENT TO SITE	1,389	798	2,355	-	-	4,542	545	204	5,291	27
14	BUILDINGS & STRUCTURES	-	7,307	7,290	-	-	14,597	1,750	658	17,005	88
	TOTAL COST	**80,174**	**15,372**	**55,740**	**-**	**-**	**151,286**	**18,052**	**6,808**	**251,804**	**1,304**

Client: ALSTOM Power Inc. **Report Date:** 4/18/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture

Jet Output Power, kV 193,037 **Estimate Type:** Conceptual **ost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,179		565			1,744	209	78	2,031	11
1.2	Fuel Stackout and Reclaim	1,524		398			1,922	231	87	2,240	12
1.3	Fuel Conveyors	1,417		358			1,775	213	80	2,068	11
1.4	Other Fuel Handling	371		83			454	54	20	528	3
1.5	Sorbent Receive & Unload	86		27			113	14	5	132	1
1.6	Sorbent Stackout and Reclaim	1,389		267			1,656	199	74	1,929	10
1.7	Sorbent Conveyors	495	100	127			722	87	33	842	4
1.8	Other Sorbent Handling	299	66	165			530	64	24	618	3
1.9	Fuel & Sorbent Hnd. Foundations		1,491	1,775			3,266	392	147	3,805	20
	SUBTOTAL. 1	**6,760**	**1,657**	**3,765**	**-**	**-**	**12,182**	**1,463**	**548**	**14,193**	**74**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	652		133			785	94	35	914	5
2.2	Fuel Conveyor to Storage	2,086		477			2,563	308	115	2,986	15
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	567		124			691	83	31	805	4
2.6	Sorbent Storage & Feed	283		113			396	48	18	462	2
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		202	192			394	47	18	459	2
	SUBTOTAL. 2	**3,588**	**202**	**1,039**	**-**	**-**	**4,829**	**580**	**217**	**5,626**	**29**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	3,591		1,360			4,951	594	223	5,768	30
3.2	Water Makeup & Pretreating	1,468		452			1,920	230	86	2,236	12
3.3	Other Feeddwater Subsystems	2,053		903			2,956	355	133	3,444	18
3.4	Service Water System	282		150			432	52	19	503	3
3.5	Other Boiler Plant Systems	1,974		1,783			3,757	451	169	4,377	23
3.6	FO Supply System & Nat. Gas	88		112			200	24	9	233	1
3.7	Waste Treatment Equipment	1,075		543			1,618	194	73	1,885	10
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	1,190		454			1,644	197	74	1,915	10
	SUBTOTAL. 3	**11,721**	**-**	**5,757**	**-**	**-**	**17,478**	**2,097**	**786**	**20,361**	**105**
4	FLUIDIZED BED BOILER						-			-	-
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			75,658	392
4.2	Open						-			-	-
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	365		105			470	56	21	547	3
4.5	Primary Air System (Fans)	1,343		385			1,728	207	78	2,013	10
4.6	Secondary Air System (Fans)	794		228			1,022	123	46	1,191	6
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	**2,502**	**-**	**718**	**-**	**-**	**3,220**	**386**	**145**	**79,409**	**411**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/18/2003

TOTAL PLANT COST SUMMARY

Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture

et Output Power, kV 193,037 **Estimate Type:** Conceptual **st Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories	3,500		3,289			6,789	815	305	7,909	41
5.4	Other Particulate Removal Materials	373		375			748	90	34	872	5
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.9	Open									-	-
	SUBTOTAL. 5	**3,873**	**-**	**3,664**	**-**	**-**	**7,537**	**905**	**339**	**8,781**	**45**
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator						-			-	-
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	**-**	**-**	**-**	**-**	**-**	**-**				
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator						-			-	-
7.2	ID Fans	1,408		404			1,812	217	82	2,111	11
7.3	Ductwork	1,818		1,309			3,127	375	141	3,643	19
7.4	Stack	3,625		1,997			5,622	675	253	6,550	34
7.9	Duct & Stack Foundations		415	445			860	103	39	1,002	5
	SUBTOTAL. 7	**6,851**	**415**	**4,155**	**-**	**-**	**11,421**	**1,370**	**515**	**13,306**	**69**
8	STEAM TURBINE GENERATOR						-				-
8.1	Steam TG & Accessories	16,242		2,330			18,572	2,229	836	21,637	112
8.2	Turbine Plant Auxiliaries	112		226			338	40	15	393	2
8.3	Condenser & Auxiliaries	2,880		694			3,574	429	161	4,164	22
8.4	Steam Piping	5,714		3,277			8,991	1,079	405	10,475	54
8.9	TG Foundations		324	522			846	101	38	985	5
	SUBTOTAL. 8	**24,948**	**324**	**7,049**	**-**	**-**	**32,321**	**3,878**	**1,455**	**37,654**	**195**
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,642		1,336			3,978	477	179	4,634	24
9.2	Circulating Water Pumps	665		55			720	86	32	838	4
9.3	Circulating Water System Auxiliaries	178		22			200	24	9	233	1
9.4	Circulating Water Piping		1,314	1,286			2,600	312	117	3,029	16
9.5	Make-up Water System	151		197			348	42	16	406	2
9.6	Component Cooling Water System	131		101			232	28	10	270	1
9.9	Circ. Water System Foundations Structures		832	1,249			2,081	250	94	2,425	13
	SUBTOTAL. 9	**3,767**	**2,146**	**4,246**	**-**	**-**	**10,159**	**1,219**	**457**	**11,835**	**61**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	331		961			1,292	155	58	1,505	8
10.7	Ash Transport & Feed Equipment	4,168		2,068			6,236	748	281	7,265	38
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		157	175			332	40	15	387	2
	SUBTOTAL. 10	**4,499**	**157**	**3,204**	**-**	**-**	**7,860**	**943**	**354**	**9,157**	**47**

Client: ALSTOM Power Inc. **Report Date:** 4/18/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture

Net Output Power, kW 193,037 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	674		91			765	92	34	891	5
115.2	Station Service Equipment	1,497		411			1,908	229	86	2,223	12
11.3	Switchgear & Motor Control	1,722		244			1,966	236	88	2,290	12
11.4	Conduit & Cable Tray		1,038	3,115			4,153	498	187	4,838	25
11.5	Wire & Cable		1,226	3,282			4,508	541	203	5,252	27
11.6	Protective Equipment	81		230			311	37	14	362	2
11.7	Standby Equipment	602		11			613	74	28	715	4
11.8	Main Power Transformer	969		38			1,007	121	45	1,173	6
11.9	Electrical Foundations		102	239			341	41	15	397	2
	SUBTOTAL. 11	**5,545**	**2,366**	**7,661**	**-**	**-**	**15,572**	**1,869**	**700**	**18,141**	**94**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	240		120			360	43	16	419	2
12.7	Distributed Control System Equipment	2,662		388			3,050	266	137	3,453	18
12.8	Instrument Wiring & tubing	1,145		3,034			4,179	501	188	4,868	25
12.9	Other I & C Equipment	684		1,295			1,979	237	89	2,305	12
	SUBTOTAL. 12	**4,731**	**-**	**4,837**	**-**	**-**	**9,568**	**1,047**	**430**	**11,045**	**57**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	393			416	50	19	485	3
13.2	Site Improvement		775	810			1,585	190	71	1,846	10
13.3	Site Facilities	1,389		1,152			2,541	305	114	2,960	15
	SUBTOTAL. 13	**1,389**	**798**	**2,355**	**-**	**-**	**4,542**	**545**	**204**	**5,291**	**27**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,188	1,940			4,128	495	186	4,809	25
14.2	Turbine Building		3,257	3,060			6,317	758	284	7,359	38
14.3	Administration Building		415	442			857	103	39	999	5
14.4	Circulation Water Pumphouse		89	71			160	19	7	186	1
14.5	Water Treatment Building		289	240			529	63	24	616	3
14.6	Machine Shop		370	251			621	74	28	723	4
14.7	Warehouse		251	254			505	61	23	589	3
14.8	Other Buildings & Structures		154	132			286	34	13	333	2
14.9	Waste Treating Building & Structure		294	900			1,194	143	54	1,391	7
	SUBTOTAL. 14	**-**	**7,307**	**7,290**	**-**	**-**	**14,597**	**1,750**	**658**	**17,005**	**88**

Table 9.3. 2: Case-1 Overall Power Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.

INITIAL & ANNUAL O&M EXPENSES

Cost Base: Jul-03

Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture

Net Plant Heat Rate (Btu/kWh): 9,611

Net Power Output (kW): 193,037

Capacity Factor (%): 80

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90 $/hour
Operating Labor Burden:	30.00 %
Labor O-H change Rate:	25.00 %

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	7.0	7.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	10.0	10.0

	Annual Cost $ / year	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	3,518,892	18.23
Maintenance Labor Costs (calc'd)	1,007,216	5.22
Administrative & Support Labor (calc'd)	1,131,527	5.86
TOTAL FIXED OPERATING COSTS	***5,657,635***	***29.31***

	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $ / year	$/kWh-net
Maintenance Material Cost (calc'd)					1,208,659	0.00089
Consumables						
Water (1000 gallons)		2,864	1.00		836,288	0.00062
Chemicals						
MU & WT Chem. (lbs.)	415,979	13,866	0.16	66,557	647,820	0.00048
Limestone (ton)	10,435	347.8	10.00	104,350	1,015,576	0.00075
Formic Acid (lbs.)			0.60			
Ammonia, NH3 (ton)			220			
Subtotal Chemicals				170,907	1,663,396	0.0012
Other Consumables						
Supplemental Fuel (MBtu)						
SCR Catalyst Replacement (MBtu)						
Emissions Penalties						
Subtotal Other						
Waste Disposal						
Fly Ash & Bottom Ash (ton)		804.3	8.00		1,878,845	0.0014
Subtotal Solid Waste Disposal					1,878,845	0.0014
By-Products & Emissions						
Gypsum (ton)						
Subtotal By-Products						
TOTAL VARIABLE OPERATING COST					***5.587.188***	***0.0041***

9.3.2. Case-2 Investment Costs and Operating and Maintenance Costs

Table 9.3. 3: Case-2 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 6/20/2003

TOTAL PLANT COST SUMMARY

Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture

Net Output Power, kW 134,514 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Account Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING	6,647	1,630	3,703	-	-	11,980	1,436	539	13,955	104
2	FUEL & SORBENT PREP. & FEED	3,524	198	1,021	-	-	4,743	569	214	5,526	41
3	FEEDWATER & MISC. BOP SYSTEMS	12,319	-	5,978	-	-	18,297	2,196	824	21,317	158
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									50,036	372
4.2	Air Separation Unit (ASU)									64,000	476
4.4	Open										
4.4 - 4.9	Boiler BOP (Fluidizing Air Fans)	170	-	49	-	-	219	26	10	255	2
	SUBTOTAL 4	170	-	49	-	-	219	26	10	114,291	850
5	FLUE GAS CLEANUP										
5.1-5.8	Miscellaneous	1,328	-	1,260	-	-	2,588	311	116	3,015	22
5.9	Gas Processing System (GPS)									57,122	425
	SUBTOTAL 5	1,328	-	1,260	-	-	2,588	311	116	60,137	447
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	-	-	-	0	0	-	-	-	-	0
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	514	10	305	-	0	829	99	37	965	7
	SUBTOTAL 7	514	10	305	-	-	829	99	37	965	7
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	16,296	-	2,338	0	0	18,634	2,236	839	21,709	161
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,734	325	4,734	0	0	13,793	1,655	620	16,068	119
	SUBTOTAL 8	25,030	325	7,072	-	-	32,427	3,891	1,459	37,777	281
9	COOLING WATER SYSTEM	3,777	2,152	4,258	-	-	10,187	1,223	459	11,869	88
10	ASH/SPENT SORBENT HANDLING SYSTEMS	3,935	137	2,801	-	-	6,873	825	347	8,045	60
11	ACCESSORY ELECTRIC PLANT	7,444	2,544	8,736	-	-	18,724	2,247	844	21,815	162
12	INSTRUMENTATION & CONTROL	4,475	-	4,577	-	-	9,052	1,086	407	10,545	78
13	IMPROVEMENT TO SITE	1,392	800	2,360	-	-	4,552	547	205	5,304	39
14	BUILDINGS & STRUCTURES	-	7,323	7,306	-	-	14,628	1,756	659	17,043	127
	TOTAL COST	**70,555**	**15,119**	**49,426**	**-**	**-**	**135,099**	**16,212**	**6,120**	**328,589**	**2,443**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 6/20/2003

TOTAL PLANT COST SUMMARY

Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture

Net Output Power, kW 134,514 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Account Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,160		555			1,715	206	77	1,998	15
1.2	Fuel Stackout and Reclaim	1,499		392			1,891	227	85	2,203	16
1.3	Fuel Conveyors	1,393		352			1,745	209	79	2,033	15
1.4	Other Fuel Handling	365		82			447	54	20	521	4
1.5	Sorbent Receive & Unload	84		27			111	13	5	129	1
1.6	Sorbent Stackout and Reclaim	1,365		262			1,627	195	73	1,895	14
1.7	Sorbent Conveyors	487	99	125			711	85	32	828	6
1.8	Other Sorbent Handling	294	65	162			521	62	23	606	5
1.9	Fuel & Sorbent Hnd. Foundations		1,466	1,746			3,212	385	145	3,742	28
	SUBTOTAL. 1	**6,647**	**1,630**	**3,703**	**-**	**-**	**11,980**	**1,436**	**539**	**13,955**	**104**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	640		131			771	93	35	899	7
2.2	Fuel Conveyor to Storage	2,049		469			2,518	302	113	2,933	22
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	557		121			678	81	31	790	6
2.6	Sorbent Storage & Feed	278		111			389	47	18	454	3
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		198	189			387	46	17	450	3
	SUBTOTAL. 2	**3,524**	**198**	**1,021**	**-**	**-**	**4,743**	**569**	**214**	**5,526**	**41**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	3,591		1,360			4,951	594	223	5,768	43
3.2	Water Makeup & Pretreating	1,798		553			2,351	282	106	2,739	20
3.3	Other Feeddwater Subsystems	2,053		903			2,956	355	133	3,444	26
3.4	Service Water System	345		183			528	63	24	615	5
3.5	Other Boiler Plant Systems	1,934		1,747			3,681	442	166	4,289	32
3.6	FO Supply System & Nat. Gas	89		112			201	24	9	234	2
3.7	Waste Treatment Equipment	1,316		664			1,980	238	89	2,307	17
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	1,193		456			1,649	198	74	1,921	14
	SUBTOTAL. 3	**12,319**	**-**	**5,978**	**-**	**-**	**18,297**	**2,196**	**824**	**21,317**	**158**
4	FLUIDIZED BED BOILER						-			-	-
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			50,036	372
4.2	Air Separation Unit (ASU)						-			64,000	476
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	34		10			44	5	2	51	0
4.5	Primary Air System (Fans)	136		39			175	21	8	204	2
4.6	Secondary Air System (Fans)						-			-	-
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	**170**	**-**	**49**	**-**	**-**	**219**	**26**	**10**	**114,291**	**850**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 6/20/2003

TOTAL PLANT COST SUMMARY

Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture

Net Output Power, kW 134,514 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Account Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories	1,150		1,081			2,231	268	100	2,599	19
5.4	Other Particulate Removal Materials	178		179			357	43	16	416	3
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.1-5.6		**1,328**	**-**	**1,260**	**-**	**-**	**2,588**	**311**	**116**	**3,015**	**22**
5.9	Gas Processing System (GPS)									57,122	425
	SUBTOTAL. 5	**1,328**	**-**	**1,260**	**-**	**-**	**2,588**	**311**	**116**	**60,137**	**447**
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator						-			-	-
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	**-**	**-**	**-**	**-**	**-**	**-**				
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator						-			-	-
7.2	ID Fans	172		49			221	26	10	257	2
7.3	Ductwork	342		246			588	71	26	685	5
7.4	Stack										
7.9	Duct & Stack Foundations		10	10			20	2	1	23	0
	SUBTOTAL. 7	**514**	**10**	**305**	**-**	**-**	**829**	**99**	**37**	**965**	**7**
8	STEAM TURBINE GENERATOR						-				-
8.1	Steam TG & Accessories	16,296		2,338			18,634	2,236	839	21,709	161
8.2	Turbine Plant Auxiliaries	112		226			338	41	15	394	3
8.3	Condenser & Auxiliaries	2,889		696			3,585	430	161	4,176	31
8.4	Steam Piping	5,733		3,288			9,021	1,082	406	10,509	78
8.9	TG Foundations		325	524			849	102	38	989	7
	SUBTOTAL. 8	**25,030**	**325**	**7,072**	**-**	**-**	**32,427**	**3,891**	**1,459**	**37,777**	**281**
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,650		1,340			3,990	479	180	4,649	35
9.2	Circulating Water Pumps	667		56			723	87	33	843	6
9.3	Circulating Water System Auxiliaries	178		22			200	24	9	233	2
9.4	Circulating Water Piping		1,317	1,289			2,606	313	117	3,036	23
9.5	Make-up Water System	151		197			348	42	16	406	3
9.6	Component Cooling Water System	131		102			233	28	10	271	2
9.9	Circ. Water System Foundations Structures		835	1,252			2,087	250	94	2,431	18
	SUBTOTAL. 9	**3,777**	**2,152**	**4,258**	**-**	**-**	**10,187**	**1,223**	**459**	**11,869**	**88**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	290		840			1,130	136	51	1,317	10
10.7	Ash Transport & Feed Equipment	3,645		1,808			5,453	654	281	6,388	47
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		137	153			290	35	15	340	3
	SUBTOTAL. 10	**3,935**	**137**	**2,801**	**-**	**-**	**6,873**	**825**	**347**	**8,045**	**60**

Client: ALSTOM Power Inc. **Report Date:** 6/20/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture

Net Output Power, kW 134,514 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Account Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	676		91			767	92	35	894	7
115.2	Station Service Equipment	2,383		654			3,037	364	137	3,538	26
11.3	Switchgear & Motor Control	2,662		378			3,040	365	137	3,542	26
11.4	Conduit & Cable Tray		1,130	3,392			4,522	543	204	5,269	39
11.5	Wire & Cable		1,312	3,512			4,824	579	217	5,620	42
11.6	Protective Equipment	148		421			569	68	26	663	5
11.7	Standby Equipment	603		11			614	74	28	716	5
11.8	Main Power Transformer	972		38			1,010	121	45	1,176	9
11.9	Electrical Foundations		102	239			341	41	15	397	3
	SUBTOTAL. 11	**7,444**	**2,544**	**8,736**	**-**	**-**	**18,724**	**2,247**	**844**	**21,815**	**162**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	227		113			340	41	15	396	3
12.7	Distributed Control System Equipment	2,518		367			2,885	346	130	3,361	25
12.8	Instrument Wiring & tubing	1,083		2,871			3,954	474	178	4,606	34
12.9	Other I & C Equipment	647		1,226			1,873	225	84	2,182	16
	SUBTOTAL. 12	**4,475**	**-**	**4,577**	**-**	**-**	**9,052**	**1,086**	**407**	**10,545**	**78**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	394			417	50	19	486	4
13.2	Site Improvement		777	811			1,588	191	71	1,850	14
13.3	Site Facilities	1,392		1,155			2,547	306	115	2,968	22
	SUBTOTAL. 13	**1,392**	**800**	**2,360**	**-**	**-**	**4,552**	**547**	**205**	**5,304**	**39**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,193	1,944			4,137	496	186	4,819	36
14.2	Turbine Building		3,265	3,068			6,333	760	285	7,378	55
14.3	Administration Building		416	443			859	103	39	1,001	7
14.4	Circulation Water Pumphouse		89	72			161	19	7	187	1
14.5	Water Treatment Building		289	240			529	64	24	617	5
14.6	Machine Shop		371	251			622	75	28	725	5
14.7	Warehouse		251	254			505	61	23	589	4
14.8	Other Buildings & Structures		154	132			286	34	13	333	2
14.9	Waste Treating Building & Structure		295	902			1,196	144	54	1,394	10
	SUBTOTAL. 14	**-**	**7,323**	**7,306**	**-**	**-**	**14,628**	**1,756**	**659**	**17,043**	**127**

Table 9.3. 4: Case-2 Boiler and Balance of Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture

Net Plant Heat Rate (Btu/kWh): 13,548

Net Power Output (kW): 134,514

Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	7.0	7.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	10.0	10.0

	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $ / year	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)					3,518,892	26.16
Maintenance Labor Costs (calc'd)					829,868	6.17
Administrative & Support Labor (calc'd)					1,087,190	8.08
TOTAL FIXED OPERATING COSTS					***5,435,950***	***40.41***
Maintenance Material Cost (calc'd)					995,842	0.00106
Consumables						
Water (11000 gallons)		3,809	1.00		1,112,228	0.00118
Chemicals						
MU & WT Chem. (lbs.)	553,200	18,440	0.16	86,146	861,517	0.00091
Limestone (ton)	10,156	339	10.00	101,556	988,420	0.00105
Formic Acid (lbs.)			0.60			
Ammonia, NH3 (ton)			220			
Subtotal Chemicals				187,702	1,849,937	0.0020
Other						
Supplemental Fuel (MBtu)						
SCR Catalyst Replacement (MBtu)						
Emissions Penalties						
Subtotal Other						
Waste Disposal						
Fly Ash & Bottom Ash (ton)		783	8.00		1,828,387	0.00194
Subtotal Solid Waste Disposal					1,828,387	0.0019
By-Products & Emissions						
Gypsum (ton)						
Subtotal By-Products						
TOTAL VARIABLE OPERATING COST					***5,786,394***	**0.0061**

Table 9.3. 5: Case-2 Gas Processing System Investment Costs

ABB LUMMUS GLOBAL HOUSTON

Project : CO2 Plant - DOE | Location : GC - USA | Project start: | Rev. : 2.01
Job/Prop # 0-9484 | Plant : CO2 Case 2 | Mech.compl.:
Scope : EPC | Capacity :
Piece count: | Labor Prod.: | 11-Feb-03

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontract ($,000)	Total ($,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		8,952	139	5,670		5,809	10.2%
12000	Vessels / Filters		2,138	33	1,354		1,387	2.4%
12100	Towers / Internals		1,502	23	951		974	1.7%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		323	5	205		210	0.4%
14200	Compressors		17,842	277	11,300		11,577	20.3%
18000	Special Equipment		-	-	-		-	0.0%
	Sub-Total Equipment	44	**30,757**	**477**	**19,479**	**-**	**19,956**	**34.9%**
21000	Civil		46,135	715	1,753		2,468	4.3%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		10,765	167	877		1,043	1.8%
23000	Buildings		12,303	191	468		658	1.2%
30000	Piping		84,582	1,311	3,896		5,207	9.1%
40000	Electrical		43,572	675	1,558		2,234	3.9%
50000	Instruments		35,883	556	2,727		3,283	5.7%
61100	Insulation		23,068	358	584		942	1.6%
61200	Fireproofing		15,378	238	292		531	0.9%
61300	Painting		12,815	199	166		364	0.6%
	Sub-Total Commodities		**284,502**	**4,410**	**12,321**	**-**	**16,731**	**29.3%**
70000	**Construction Indirects**						**7,093**	**12.4%**
	Sub-Total Direct Cost		**315,259**	**4,887**	**31,800**	**-**	**43,780**	**76.6%**
	ASU TIC plant cost						-	0.0%
71000	Constr. Management						700	1.2%
80000	Home Office Engineering						4,488	7.9%
80000	Basic Engineering						600	1.1%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						800	1.4%
19300	Spare parts						1,300	2.3%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						100	0.2%
97000	Freight						954	1.7%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**52,722**	**92.3%**
91400	Escalation						1,500	2.6%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**54,222**	**94.9%**
	Contracters Fee						2,900	5.1%
	Grand Total						**57,122**	**100.0%**

Exclusions : Bonds,Taxes,Import duties , Hazerdous material handling & disposal, Capital spare parts, Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MCC.

Table 9.3. 6: Case-2 Gas Processing System Operating and Maintenance Costs

Operating Costs ($/yr)	**Variable Costs**	**Fixed Costs**
Chemical and Dessicant	52500	
Waste Handling	0	
Fuel Gas *	137921	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	1615980	
Contracted services	830000	
Column Total	2636401	306600
Grand Total (Fixed & Variable)	2943001	
* Based on $4/ MMBU and 7000 hours/ yr.		
** Included in overall facility operating cost		

Table 9.3. 7: Case-2 Air Separation Unit Operating and Maintenance Costs

Operating Cost ($/yr)	Variable Costs	Fixed Costs
Minor Consumables	20,000	
Cooling Water*	0	
Natural Gas***	357,253	
Prepurified Adsorbent**	0	
Operating Labor		2,111,335
Column Total	**377,253**	**2,111,335**
Grand Total (Fixed + Variable)	**2,488,588**	

* Cooling water is supplied by others; thus, major treatment chemicals are part of this supply

** Prepurified adsorbent is included in the plant and is typically not replaced

***Based on $\$4.0/10^6$ Btu and 7008 hours/year

9.3.3. Case-3 Investment Costs and Operating and Maintenance Costs

Table 9.3. 8: Case-3 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 3 - 1x200 gr. MW O2-Fired CFB w/ASU & Flue Gas Sequestration

Net Output Power, kW 135,351 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING	6,647	1,630	3,703	-	-	11,980	1,436	539	13,955	103
2	FUEL & SORBENT PREP. & FEED	3,524	198	1,021	-	-	4,743	569	214	5,526	41
3	FEEDWATER & MISC. BOP SYSTEMS	12,129	-	5,902	-	-	18,031	2,165	811	21,007	155
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									50,036	370
4.2	Air Separation Unit (ASU)									64,000	473
4.4	Open										
4.4 - 4.9	Boiler BoP (Fluidizing Air Fans)	170	-	49	-	-	219	26	10	255	2
	SUBTOTAL 4	170	-	49	-	-	219	26	10	114,291	844
5	FLUE GAS CLEANUP										
5.1-5.6	Miscellaneous	1,328	-	1,260	-	-	2,588	311	116	3,015	22
5.9	Gas Processing System (GPS)									49,519	366
	SUBTOTAL 5	1,328	-	1,260	-	-	2,588	311	116	52,534	388
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	-	-	-	0	0	-	-	-	-	0
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	514	10	305	-	0	829	99	37	965	7
	SUBTOTAL 7	514	10	305	-	-	829	99	37	965	7
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	16,296	-	2,338	0	0	18,634	2,236	839	21,709	160
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,734	325	4,734	0	0	13,793	1,655	620	16,068	119
	SUBTOTAL 8	25,030	325	7,072	-	-	32,427	3,891	1,459	37,777	279
9	COOLING WATER SYSTEM	3,777	2,152	4,258	-	-	10,187	1,223	459	11,869	88
10	ASH/SPENT SORBENT HANDLING SYSTEMS	3,935	137	2,801	-	-	6,873	825	309	8,007	59
11	ACCESSORY ELECTRIC PLANT	7,444	2,544	8,736	-	-	18,724	2,247	844	21,815	161
12	INSTRUMENTATION & CONTROL	4,475	-	4,577	-	-	9,052	1,086	407	10,545	78
13	IMPROVEMENT TO SITE	1,392	800	2,360	-	-	4,552	547	205	5,304	39
14	BUILDINGS & STRUCTURES	-	7,323	7,306	-	-	14,628	1,756	659	17,043	126
	TOTAL COST	**70,365**	**15,119**	**49,350**	**-**	**-**	**134,833**	**16,181**	**6,069**	**320,638**	**2,369**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 3 - 1x200 gr. MW O2-Fired CFB w/ASU & Flue Gas Sequestration

Net Output Power, kW 135,351 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,160		555			1,715	206	77	1,998	15
1.2	Fuel Stackout and Reclaim	1,499		392			1,891	227	85	2,203	16
1.3	Fuel Conveyors	1,393		352			1,745	209	79	2,033	15
1.4	Other Fuel Handling	365		82			447	54	20	521	4
1.5	Sorbent Receive & Unload	84		27			111	13	5	129	1
1.6	Sorbent Stackout and Reclaim	1,365		262			1,627	195	73	1,895	14
1.7	Sorbent Conveyors	487	99	125			711	85	32	828	6
1.8	Other Sorbent Handling	294	65	162			521	62	23	606	4
1.9	Fuel & Sorbent Hnd. Foundations		1,466	1,746			3,212	385	145	3,742	28
	SUBTOTAL. 1	**6,647**	**1,630**	**3,703**	**-**	**-**	**11,980**	**1,436**	**539**	**13,955**	**103**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	640		131			771	93	35	899	7
2.2	Fuel Conveyor to Storage	2,049		469			2,518	302	113	2,933	22
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	557		121			678	81	31	790	6
2.6	Sorbent Storage & Feed	278		111			389	47	18	454	3
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		198	189			387	46	17	450	3
	SUBTOTAL. 2	**3,524**	**198**	**1,021**	**-**	**-**	**4,743**	**569**	**214**	**5,526**	**41**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	3,591		1,360			4,951	594	223	5,768	43
3.2	Water Makeup & Pretreating	1,699		523			2,222	267	100	2,589	19
3.3	Other Feeddwater Subsystems	2,053		903			2,956	355	133	3,444	25
3.4	Service Water System	326		173			499	60	22	581	4
3.5	Other Boiler Plant Systems	1,934		1,747			3,681	442	166	4,289	32
3.6	FO Supply System & Nat. Gas	89		112			201	24	9	234	2
3.7	Waste Treatment Equipment	1,244		628			1,872	225	84	2,181	16
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	1,193		456			1,649	198	74	1,921	14
	SUBTOTAL. 3	**12,129**	**-**	**5,902**	**-**	**-**	**18,031**	**2,165**	**811**	**21,007**	**155**
4	FLUIDIZED BED BOILER						-			-	-
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			50,036	370
4.2	Air Separation Unit (ASU)						-			64,000	473
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	34		10			44	5	2	51	0
4.5	Primary Air System (Fans)	136		39			175	21	8	204	2
4.6	Secondary Air System (Fans)						-			-	-
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	**170**	**-**	**49**	**-**	**-**	**219**	**26**	**10**	**114,291**	**844**

Client: ALSTOM Power Inc. **Report Date:** 4/24/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 3 - 1x200 gr. MW O2-Fired CFB w/ASU & Flue Gas Sequestration

Net Output Power, kW 135,351 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories	1,150		1,081			2,231	268	100	2,599	19
5.4	Other Particulate Removal Materials	178		179			357	43	16	416	3
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.1-5.6	Miscellaneous	1,328	-	1,260	-	-	2,588	311	116	3,015	22
5.9	Gas Processing System (GPS)									49,519	366
	SUBTOTAL. 5	1,328	-	1,260	-	-	2,588	311	116	52,534	388
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator						-			-	-
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	**-**	**-**	**-**	**-**	**-**	**-**				
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator						-			-	-
7.2	ID Fans	172		49			221	26	10	257	2
7.3	Ductwork	342		246			588	71	26	685	5
7.4	Stack										-
7.9	Duct & Stack Foundations		10	10			20	2	1	23	0
	SUBTOTAL. 7	**514**	**10**	**305**	**-**	**-**	**829**	**99**	**37**	**965**	**7**
8	STEAM TURBINE GENERATOR						-				-
8.1	Steam TG & Accessories	16,296		2,338			18,634	2,236	839	21,709	160
8.2	Turbine Plant Auxiliaries	112		226			338	41	15	394	3
8.3	Condenser & Auxiliaries	2,889		696			3,585	430	161	4,176	31
8.4	Steam Piping	5,733		3,288			9,021	1,082	406	10,509	78
8.9	TG Foundations		325	524			849	102	38	989	7
	SUBTOTAL. 8	**25,030**	**325**	**7,072**	**-**	**-**	**32,427**	**3,891**	**1,459**	**37,777**	**279**
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,650		1,340			3,990	479	180	4,649	34
9.2	Circulating Water Pumps	667		56			723	87	33	843	6
9.3	Circulating Water System Auxiliaries	178		22			200	24	9	233	2
9.4	Circulating Water Piping		1,317	1,289			2,606	313	117	3,036	22
9.5	Make-up Water System	151		197			348	42	16	406	3
9.6	Component Cooling Water System	131		102			233	28	10	271	2
9.9	Circ. Water System Foundations Structures		835	1,252			2,087	250	94	2,431	18
	SUBTOTAL. 9	**3,777**	**2,152**	**4,258**	**-**	**-**	**10,187**	**1,223**	**459**	**11,869**	**88**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	290		840			1,130	136	51	1,317	10
10.7	Ash Transport & Feed Equipment	3,645		1,808			5,453	654	245	6,352	47
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		137	153			290	35	13	338	2
	SUBTOTAL. 10	**3,935**	**137**	**2,801**	**-**	**-**	**6,873**	**825**	**309**	**8,007**	**59**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 3 - 1x200 gr. MW O2-Fired CFB w/ASU & Flue Gas Sequestration

Net Output Power, kW 135,351 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	676		91			767	92	35	894	7
115.2	Station Service Equipment	2,383		654			3,037	364	137	3,538	26
11.3	Switchgear & Motor Control	2,662		378			3,040	365	137	3,542	26
11.4	Conduit & Cable Tray		1,130	3,392			4,522	543	204	5,269	39
11.5	Wire & Cable		1,312	3,512			4,824	579	217	5,620	42
11.6	Protective Equipment	148		421			569	68	26	663	5
11.7	Standby Equipment	603		11			614	74	28	716	5
11.8	Main Power Transformer	972		38			1,010	121	45	1,176	9
11.9	Electrical Foundations		102	239			341	41	15	397	3
	SUBTOTAL. 11	**7,444**	**2,544**	**8,736**	**-**	**-**	**18,724**	**2,247**	**844**	**21,815**	**161**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	227		113			340	41	15	396	3
12.7	Distributed Control System Equipment	2,518		367			2,885	346	130	3,361	25
12.8	Instrument Wiring & tubing	1,083		2,871			3,954	474	178	4,606	34
12.9	Other I & C Equipment	647		1,226			1,873	225	84	2,182	16
	SUBTOTAL. 12	**4,475**	**-**	**4,577**	**-**	**-**	**9,052**	**1,086**	**407**	**10,545**	**78**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	394			417	50	19	486	4
13.2	Site Improvement		777	811			1,588	191	71	1,850	14
13.3	Site Facilities	1,392		1,155			2,547	306	115	2,968	22
	SUBTOTAL. 13	**1,392**	**800**	**2,360**	**-**	**-**	**4,552**	**547**	**205**	**5,304**	**39**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,193	1,944			4,137	496	186	4,819	36
14.2	Turbine Building		3,265	3,068			6,333	760	285	7,378	55
14.3	Administration Building		416	443			859	103	39	1,001	7
14.4	Circulation Water Pumphouse		89	72			161	19	7	187	1
14.5	Water Treatment Building		289	240			529	64	24	617	5
14.6	Machine Shop		371	251			622	75	28	725	5
14.7	Warehouse		251	254			505	61	23	589	4
14.8	Other Buildings & Structures		154	132			286	34	13	333	2
14.9	Waste Treating Building & Structure		295	902			1,196	144	54	1,394	10
	SUBTOTAL. 14	**-**	**7,323**	**7,306**	**-**	**-**	**14,628**	**1,756**	**659**	**17,043**	**126**

Table 9.3. 9: Case-3 Boiler and Balance of Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 3 - 1x200 gr. MW O2-Fired CFB w/ASU & Flue Gas Sequestration

Net Plant Heat Rate (Btu/kWh): 13,492

Net Power Output (kW): 135,351

Capacity Factor (%): 80

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	7.0	7.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	10.0	10.0

	Annual Cost $ / year	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	3,518,892	26.00
Maintenance Labor Costs (calc'd)	828,476	6.12
Administrative & Support Labor (calc'd)	1,295,484	9.57
TOTAL FIXED OPERATING COSTS	***5,642,852***	***41.69***
Maintenance Material Cost (calc'd)	994,171	0.0010

Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $ / year	Annual Unit Cost $/kW-net
Water (1000 gallons)		3,517	1.00		1,026,964	0.0011
Chemicals						
MU & WT Chem. (lbs.)	510,750	17,025	0.16	79,536	795,408	0.0008
Limestone (ton)	10,156	338.5	10.00	101,556	988,420	0.0010
Formic Acid (lbs.)			0.60			
Ammonia, NH3 (ton)			220			
Subtotal Chemicals				181,092	1,783,828	0.0019
Other						
Supplemental Fuel (MBtu)						
SCR Catalyst Replacement (MBtu)						
Emissions Penalties						
Subtotal Other						
Waste Disposal						
Fly Ash & Bottom Ash (ton)		782.7	8.00		1,828,387	0.0019
Subtotal Solid Waste Disposal					1,828,387	0.0019
By-Products & Emissions						
Gypsum (ton)						
Subtotal By-Products						
TOTAL VARIABLE OPERATING COST					***5,633,350***	***0.0059***

Table 9.3. 10: Case-3 Gas Processing System Investment Costs

ABB LUMMUS GLOBAL HOUSTON

Project	CO2 Plant - DOE	Location : GC - USA	Project start:	Rev. : 00
Job/Prop #	0-9484	Plant : CO2 Case 3	Mech.compl.:	
Scope	EPC	Capacity :		
Piece count:		Labor Prod.:		1-Nov-02

Acc't Code	Description	Pieces	Direct Manhours	Labor ($.000)	Material ($.000)	Subcontract ($.000)	Total ($.000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		3,919	61	2,482		2,543	5.1%
12000	Vessels / Filters		986	15	625		640	1.3%
12100	Towers / Internals		1,099	17	696		713	1.4%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		87	1	55		57	0.1%
14200	Compressors		17,368	269	11,000		11,269	22.8%
18000	Special Equipment		4,018	62	2,545		2,607	5.3%
	Sub-Total Equipment	27	27,478	426	17,403	-	17,829	36.0%
21000	Civil		41,217	639	1,566		2,205	4.5%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		9,617	149	783		932	1.9%
23000	Buildings		10,991	170	418		588	1.2%
30000	Piping		75,564	1,171	3,481		4,652	9.4%
40000	Electrical		38,927	603	1,392		1,996	4.0%
50000	Instruments		32,058	497	2,436		2,933	5.9%
61100	Insulation		20,608	319	522		842	1.7%
61200	Fireproofing		13,739	213	261		474	1.0%
61300	Painting		11,449	177	148		325	0.7%
	Sub-Total Commodities		254,170	3,940	11,007	-	14,947	30.2%
70000	**Construction Indirects**						6,337	12.8%
	Sub-Total Direct Cost		281,648	4,366	28,410	-	39,112	79.0%
	ASU TIC plant cost						-	0.0%
71000	Constr. Management						600	1.2%
80000	Home Office Engineering						2,754	5.6%
80000	Basic Engineering						400	0.8%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						700	1.4%
19300	Spare parts						1,200	2.4%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						100	0.2%
97000	Freight						852	1.7%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						45,719	92.3%
91400	Escalation						1,300	2.6%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						47,019	95.0%
	Contracters Fee						2,500	5.0%
	Grand Total						49,519	100.0%

Exclusions : Bonds,Taxes,Import duties , Hazerdous material handling & disposal, Capital spare parts, Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MCC.

Table 9.3.11: Case-3 Gas Processing System Operating and Maintenance Costs

Operating Costs ($/yr)	Variable Costs	Fixed Costs
Chemical and Dessicant	38066.66667	
Waste Handling	0	
Fuel Gas *	139160	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	1615980	
Contracted services	830000	
Column Total	2623207	306600
Grand Total (Fixed & Variable)	2929807	
* Based on $7/ MMBU and 7000 hours/ yr.		
** Included in overall facility operating cost		

Table 9.3. 12: Case-3 Air Separation Unit Operating and Maintenance Costs

Operating Cost ($/yr)	Variable Costs	Fixed Costs
Minor Consumables	20,000	
Cooling Water*	0	
Natural Gas***	357,253	
Prepurified Adsorbent**	0	
Operating Labor		2,111,335
Column Total	**377,253**	**2,111,335**
Grand Total (Fixed + Variable)	**2,488,588**	

* Cooling water is supplied by others; thus, major treatment chemicals are part of this supply

** Prepurified adsorbent is included in the plant and is typically not replaced

***Based on $4.0/10^6 Btu and 7008 hours/year

9.3.4. Case-4 Investment Costs and Operating and Maintenance Costs

Table 9.3.13: Case-4 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc. **Report Date:** 4/24/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 4 - 1x200 gr. MW O2-Fired CMB w/ASU & CO2 Capture

Net Output Power, kW 132,168 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					$	$/kW
1	FUEL & SORBENT HANDLING	6,648	1,628	3,701	-	-	11,977	1,438	537	13,952	106
2	FUEL & SORBENT PREP. & FEED	3,525	199	1,022	-	-	4,746	570	214	5,530	42
3	FEEDWATER & MISC. BOP SYSTEMS	12,330	-	5,982	-	-	18,312	2,198	823	21,333	161.41
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									58,349	441
4.2	Air Separation Unit (ASU)									64,000	484
4.4	Open										
4.4 - 4.9	Boiler BoP (Fluidizing Air Fans)	528	-	151	-	-	679	82	30	791	6
	SUBTOTAL 4	528	-	151	-	-	679	82	30	123,140	932
5	FLUE GAS CLEANUP										
5.1-5.6	Miscellaneous	1,371	-	1,301	-	-	2,672	320	121	3,113	24
5.9	Gas Processing System (GPS)									56,898	430
	SUBTOTAL 5	1,371	-	1,301	-	-	2,672	320	121	60,011	454
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	-	-	-	0	0	-	-	-	-	0
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	518	10	307	-	0	835	100	37	972	7
	SUBTOTAL 7	518	10	307	-	-	835	100	37	972	7
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	16,304	-	2,339	0	0	18,643	2,237	839	21,719	164
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,738	325	4,735	0	0	13,798	1,656	620	16,074	122
	SUBTOTAL 8	25,042	325	7,074	-	-	32,441	3,893	1,459	37,793	286
9	COOLING WATER SYSTEM	3,778	2,153	4,260	-	-	10,191	1,224	459	11,874	90
10	ASH/SPENT SORBENT HANDLING SYSTEMS	3,935	137	2,801	-	-	6,873	825	347	8,045	61
11	ACCESSORY ELECTRIC PLANT	7,454	2,549	8,753	-	-	18,756	2,250	845	21,851	165
12	INSTRUMENTATION & CONTROL	4,477	-	4,578	-	-	9,055	1,087	407	10,549	80
13	IMPROVEMENT TO SITE	1,392	800	2,361	-	-	4,553	547	205	5,305	40
14	BUILDINGS & STRUCTURES	-	7,324	7,308	-	-	14,631	1,757	659	17,047	129
	TOTAL COST	**70,998**	**15,125**	**49,599**	**-**	**-**	**135,721**	**16,291**	**6,143**	**337,402**	**2,553**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 4 - 1x200 gr. MW O2-Fired CMB w/ASU & CO2 Capture

Net Output Power, kW 132,168 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					$	$/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,159		555			1,714	206	77	1,997	15
1.2	Fuel Stackout and Reclaim	1,497		391			1,888	227	85	2,200	17
1.3	Fuel Conveyors	1,392		352			1,744	209	78	2,031	15
1.4	Other Fuel Handling	364		81			445	53	20	518	4
1.5	Sorbent Receive & Unload	85		27			112	13	5	130	1
1.6	Sorbent Stackout and Reclaim	1,368		263			1,631	196	73	1,900	14
1.7	Sorbent Conveyors	488	99	126			713	86	32	831	6
1.8	Other Sorbent Handling	295	65	162			522	63	23	608	5
1.9	Fuel & Sorbent Hnd. Foundations		1,464	1,744			3,208	385	144	3,737	28
	SUBTOTAL. 1	**6,648**	**1,628**	**3,701**	**-**	**-**	**11,977**	**1,438**	**537**	**13,952**	**106**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	640		131			771	92	35	898	7
2.2	Fuel Conveyor to Storage	2,047		469			2,516	302	113	2,931	22
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	559		122			681	82	31	794	6
2.6	Sorbent Storage & Feed	279		111			390	47	18	455	3
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		199	189			388	47	17	452	3
	SUBTOTAL. 2	**3,525**	**199**	**1,022**	**-**	**-**	**4,746**	**570**	**214**	**5,530**	**42**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	3,591		1,360			4,951	594	223	5,768	44
3.2	Water Makeup & Pretreating	1,804		555			2,359	283	106	2,748	21
3.3	Other Feeddwater Subsystems	2,053		903			2,956	355	133	3,444	26
3.4	Service Water System	347		184			531	64	24	619	5
3.5	Other Boiler Plant Systems	1,932		1,745			3,677	441	165	4,283	32
3.6	FO Supply System & Nat. Gas	89		112			201	24	9	234	2
3.7	Waste Treatment Equipment	1,321		667			1,988	239	89	2,316	18
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	1,193		456			1,649	198	74	1,921	15
	SUBTOTAL. 3	**12,330**	**-**	**5,982**	**-**	**-**	**18,312**	**2,198**	**823**	**21,333**	**161**
4	FLUIDIZED BED BOILER						-			-	-
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			58,349	441
4.2	Air Separation Unit (ASU)						-			64,000	484
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	35		10			45	5	2	52	0
4.5	Primary Air System (Fans)	140		40			180	22	8	210	2
4.6	Secondary Air System (Fans)	353		101			454	55	20	529	4
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	**528**	**-**	**151**	**-**	**-**	**679**	**82**	**30**	**123,140**	**932**

Client: ALSTOM Power Inc. **Report Date:** 4/24/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 4 - 1x200 gr. MW O2-Fired CMB w/ASU & CO2 Capture

Net Output Power, kW 132,168 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories	1,187		1,116			2,303	276	104	2,683	20
5.4	Other Particulate Removal Materials	184		185			369	44	17	430	3
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.1-5.6	Miscellaneous	1,371	-	1,301	-	-	2,672	320	121	3,113	24
5.9	Gas Processing System (GPS)									56,898	430
	SUBTOTAL. 5	**1,371**	**-**	**1,301**	**-**	**-**	**2,672**	**320**	**121**	**60,011**	**454**
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator						-			-	-
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	**-**	**-**	**-**	**-**	**-**	**-**				
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator						-			-	-
7.2	ID Fans	176		51			227	27	10	264	2
7.3	Ductwork	342		246			588	71	26	685	5
7.4	Stack										-
7.9	Duct & Stack Foundations		10	10			20	2	1	23	0
	SUBTOTAL. 7	**518**	**10**	**307**	**-**	**-**	**835**	**100**	**37**	**972**	**7**
8	STEAM TURBINE GENERATOR						-				-
8.1	Steam TG & Accessories	16,304		2,339			18,643	2,237	839	21,719	164
8.2	Turbine Plant Auxiliaries	112		226			338	41	15	394	3
8.3	Condenser & Auxiliaries	2,890		696			3,586	430	161	4,177	32
8.4	Steam Piping	5,736		3,289			9,025	1,083	406	10,514	80
8.9	TG Foundations		325	524			849	102	38	989	7
	SUBTOTAL. 8	**25,042**	**325**	**7,074**	**-**	**-**	**32,441**	**3,893**	**1,459**	**37,793**	**286**
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,651		1,340			3,991	479	180	4,650	35
9.2	Circulating Water Pumps	667		56			723	87	33	843	6
9.3	Circulating Water System Auxiliaries	178		22			200	24	9	233	2
9.4	Circulating Water Piping		1,318	1,290			2,608	313	117	3,038	23
9.5	Make-up Water System	151		197			348	42	16	406	3
9.6	Component Cooling Water System	131		102			233	28	10	271	2
9.9	Circ. Water System Foundations Structures		835	1,253			2,088	251	94	2,433	18
	SUBTOTAL. 9	**3,778**	**2,153**	**4,260**	**-**	**-**	**10,191**	**1,224**	**459**	**11,874**	**90**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	290		840			1130	136	51	1,317	10
10.7	Ash Transport & Feed Equipment	3645		1808			5453	654	281	6,388	48
10.8	Misc. Ash Handling Equipment						0			-	-
10.9	Ash/Spent Sorbent Foundations		137	153			290	35	15	340	3
	SUBTOTAL. 10	**3,935**	**137**	**2,801**	**-**	**-**	**6,873**	**825**	**347**	**8,045**	**61**

Client: ALSTOM Power Inc. **Report Date:** 4/24/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 4 - 1x200 gr. MW O2-Fired CMB w/ASU & CO2 Capture

Net Output Power, kW 132,168 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	676		92			768	92	35	895	7
115.2	Station Service Equipment	2,387		655			3,042	365	137	3,544	27
11.3	Switchgear & Motor Control	2,667		378			3,045	365	137	3,547	27
11.4	Conduit & Cable Tray		1,133	3,399			4,532	544	204	5,280	40
11.5	Wire & Cable		1,314	3,519			4,833	580	218	5,631	43
11.6	Protective Equipment	149		422			571	68	26	665	5
11.7	Standby Equipment	603		11			614	74	28	716	5
11.8	Main Power Transformer	972		38			1,010	121	45	1,176	9
11.9	Electrical Foundations		102	239			341	41	15	397	3
	SUBTOTAL. 11	**7,454**	**2,549**	**8,753**	**-**	**-**	**18,756**	**2,250**	**845**	**21,851**	**165**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	227		113			340	41	15	396	3
12.7	Distributed Control System Equipment	2,519		368			2,887	346	130	3,363	25
12.8	Instrument Wiring & tubing	1,084		2,871			3,955	475	178	4,608	35
12.9	Other I & C Equipment	647		1,226			1,873	225	84	2,182	17
	SUBTOTAL. 12	**4,477**	**-**	**4,578**	**-**	**-**	**9,055**	**1,087**	**407**	**10,549**	**80**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	394			417	50	19	486	4
13.2	Site Improvement		777	812			1,589	191	71	1,851	14
13.3	Site Facilities	1,392		1,155			2,547	306	115	2,968	22
	SUBTOTAL. 13	**1,392**	**800**	**2,361**	**-**	**-**	**4,553**	**547**	**205**	**5,305**	**40**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,193	1,945			4,138	497	186	4,821	36
14.2	Turbine Building		3,266	3,069			6,335	760	285	7,380	56
14.3	Administration Building		416	443			859	103	39	1,001	8
14.4	Circulation Water Pumphouse		89	72			161	19	7	187	1
14.5	Water Treatment Building		289	240			529	64	24	617	5
14.6	Machine Shop		371	251			622	75	28	725	5
14.7	Warehouse		251	254			505	61	23	589	4
14.8	Other Buildings & Structures		154	132			286	34	13	333	3
14.9	Waste Treating Building & Structure		295	902			1,196	144	54	1,394	11
	SUBTOTAL. 14	**-**	**7,324**	**7,308**	**-**	**-**	**14,631**	**1,757**	**659**	**17,047**	**129**

Table 9.3.14: Case-4 Boiler and Balance of Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 4 - 1x200 gr. MW O2-Fired CMB w/ASU & CO2 Capture

Net Plant Heat Rate (Btu/kWh): 13,894
Net Power Output (kW): 132,168
Capacity Factor (%): 80

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	7.0	7.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	10.0	10.0

	Annual Cost $ / year	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	3,518,892	26.62
Maintenance Labor Costs (calc'd)	873,760	6.61
Administrative & Support Labor (calc'd)	1,098,163	8.31
TOTAL FIXED OPERATING COSTS	***5,490,815***	***41.54***
Maintenance Material Cost (calc'd)	1,048,512	0.0011

Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $ / year	Annual Unit Cost $/kW-net
Water (11000 gallons)		3,829	1.00		1,118,068	0.0012
Chemicals						
MU & WT Chem. (lbs.)	556,002	18,533	0.16	86,583	865,862	0.0009
Limestone (ton)	10,198	341.2	10.00	101,984	996,304	0.0011
Formic Acid (lbs.)			0.60			
Ammonia, NH3 (ton)			220			
Subtotal Chemicals				188,567	1,862,166	0.0020
Other						
Supplemental Fuel (MBtu)						
SCR Catalyst Replacement (MBtu)						
Emissions Penalties						
Subtotal Other						
Waste Disposal						
Fly Ash & Bottom Ash (ton)		788.9	8.00		1,842,870	0.0020
Subtotal Solid Waste Disposal					1,842,870	0.0020
By-Products & Emissions						
Gypsum (ton)						
Subtotal By-Products						
TOTAL VARIABLE OPERATING COST					***5,871,616***	***0.0063***

Table 9.3.15: Case-4 Gas Processing System Investment Costs

ABB LUMMUS GLOBAL HOUSTON

Project : CO2 Plant - DOE | Location : GC - USA | Project start: | Rev. : 4.02
Job/Prop # : 0-9484 | Plant : CO2 Case 4 | Mech.compl.:
Scope : EPC | Capacity :
Piece count: | Labor Prod.: | 11-Feb-03

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontract ($,000)	Total ($,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		8,956	139	5,672		5,811	10.2%
12000	Vessels / Filters		2,138	33	1,354		1,387	2.4%
12100	Towers / Internals		1,502	23	951		974	1.7%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		323	5	205		210	0.4%
14200	Compressors		17,684	274	11,200		11,474	20.2%
18000	Special Equipment		-	-	-		-	0.0%
	Sub-Total Equipment	44	**30,603**	**474**	**19,382**	**-**	**19,856**	**34.9%**
21000	Civil		45,904	712	1,744		2,456	4.3%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		10,711	166	872		1,038	1.8%
23000	Buildings		12,241	190	465		655	1.2%
30000	Piping		84,158	1,304	3,876		5,181	9.1%
40000	Electrical		43,354	672	1,551		2,223	3.9%
50000	Instruments		35,703	553	2,713		3,267	5.7%
61100	Insulation		22,952	356	581		937	1.6%
61200	Fireproofing		15,301	237	291		528	0.9%
61300	Painting		12,751	198	165		362	0.6%
	Sub-Total Commodities		**283,077**	**4,388**	**12,259**	**-**	**16,647**	**29.3%**
70000	**Construction Indirects**						**7,058**	**12.4%**
	Sub-Total Direct Cost		**313,680**	**4,862**	**31,641**	**-**	**43,561**	**76.6%**
	ASU TIC plant cost						-	0.0%
71000	Constr. Management						700	1.2%
80000	Home Office Engineering						4,488	7.9%
80000	Basic Engineering						600	1.1%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						800	1.4%
19300	Spare parts						1,300	2.3%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						100	0.2%
97000	Freight						949	1.7%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**52,498**	**92.3%**
91400	Escalation						1,500	2.6%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**53,998**	**94.9%**
	Contracters Fee						2,900	5.1%
	Grand Total						**56,898**	**100.0%**

Exclusions : Bonds,Taxes,Import duties , Hazerdous material handling & disposal, Capital spare parts, Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MCC.

Table 9.3.16: Case-4 Gas Processing System Operating and Maintenance Costs

Operating Costs ($/yr)	Variable Costs	Fixed Costs
Chemical and Dessicant	52500	
Waste Handling	0	
Fuel Gas *	137921	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	1615980	
Contracted services	830000	
Column Total	2636401	306600
Grand Total (Fixed & Variable)	2943001	

* Based on $4/ MMBU and 7000 hours/ yr.
** Included in overall facility operating cost

Table 9.3. 17: Case-4 Air Separation Unit Operating and Maintenance Costs

Operating Cost ($/yr)	Variable Costs	Fixed Costs
Minor Consumables	20,000	
Cooling Water*	0	
Natural Gas***	357,253	
Prepurified Adsorbent**	0	
Operating Labor		2,111,335
Column Total	**377,253**	**2,111,335**
Grand Total (Fixed + Variable)	**2,488,588**	

* Cooling water is supplied by others; thus, major treatment chemicals are part of this supply

** Prepurified adsorbent is included in the plant and is typically not replaced

***Based on $4.0/10^6$ Btu and 7008 hours/year

9.3.5. Case-5 Investment Costs and Operating and Maintenance Costs

Table 9.3.18: Case-5 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 5 - 1x200 gr. MW Air-Fired CFB w/Carbonate Regeneration Process & CO2 Capture

Net Output Power, kW 161,184 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cos	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING	6,838	1,650	3,764	-	-	12,252	1,471	551	14,274	89
2	FUEL & SORBENT PREP. & FEED	3,604	213	1,056	-	-	4,873	585	220	5,678	35
3	FEEDWATER & MISC. BOP SYSTEMS	11,767	-	5,761	-	-	17,528	2,103	789	20,420	126.69
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									61,101	379
4.2	Open										
4.4	Open										
4.4 - 4.9	Boiler BoP (Fluidizing Air Fans)	1,349	-	387	-	-	1,736	208	78	2,022	13
	SUBTOTAL 4	1,349	-	387	-	-	1,736	208	78	63,123	392
5	FLUE GAS CLEANUP	-	-	-	-	-	-	-	-		
5.1-5.6	Miscellaneous										
5.9	Gas Processing System (GPS)									51,382	319
	SUBTOTAL 5	-	-	-	-	-	-	-	-	51,382	319
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	-	-	-	0	0	-	-	-	-	0
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	5,161	45	3,039	-	0	8,245	989	371	9,605	60
	SUBTOTAL 7	5,161	45	3,039	-	-	8,245	989	371	9,605	60
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	15,923	-	2,285	0	0	18,208	2,185	819	21,212	132
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,540	318	4,628	0	0	13,486	1,619	607	15,712	97
	SUBTOTAL 8	24,463	318	6,913	-	-	31,694	3,804	1,426	36,924	229
9	COOLING WATER SYSTEM	3,701	2,109	4,173	-	-	9,983	1,198	449	11,630	72
10	ASH/SPENT SORBENT HANDLING SYSTEMS	3,684	129	2,624	-	-	6,437	773	288	7,498	47
11	ACCESSORY ELECTRIC PLANT	6,036	1,897	6,534	-	-	14,467	1,736	653	16,856	105
12	INSTRUMENTATION & CONTROL	4,583	-	4,687	-	-	9,270	1,113	417	10,800	67
13	IMPROVEMENT TO SITE	1,373	789	2,327	-	-	4,489	538	202	5,229	32
14	BUILDINGS & STRUCTURES	-	7,217	7,205	-	-	14,433	1,730	650	16,813	104
	TOTAL COST	**72,559**	**14,367**	**48,470**	**-**	**-**	**135,407**	**16,248**	**6,094**	**270,232**	**1,677**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 5 - 1x200 gr. MW Air-Fired CFB w/Carbonate Regeneration Process & CO2 Capture

Net Output Power, kW 161,184 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,167		559			1,726	207	78	2,011	12
1.2	Fuel Stackout and Reclaim	1,508		394			1,902	228	86	2,216	14
1.3	Fuel Conveyors	1,402		355			1,757	211	79	2,047	13
1.4	Other Fuel Handling	367		82			449	54	20	523	3
1.5	Sorbent Receive & Unload	91		29			120	14	5	139	1
1.6	Sorbent Stackout and Reclaim	1,464		281			1,745	210	79	2,034	13
1.7	Sorbent Conveyors	523	106	134			763	92	34	889	6
1.8	Other Sorbent Handling	316	69	174			559	67	25	651	4
1.9	Fuel & Sorbent Hnd. Foundations		1,475	1,756			3,231	388	145	3,764	23
	SUBTOTAL. 1	**6,838**	**1,650**	**3,764**	**-**	**-**	**12,252**	**1,471**	**551**	**14,274**	**89**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	644		132			776	93	35	904	6
2.2	Fuel Conveyor to Storage	2,062		472			2,534	304	114	2,952	18
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	599		130			729	88	33	850	5
2.6	Sorbent Storage & Feed	299		119			418	50	19	487	3
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		213	203			416	50	19	485	3
	SUBTOTAL. 2	**3,604**	**213**	**1,056**	**-**	**-**	**4,873**	**585**	**220**	**5,678**	**35**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	3,591		1,360			4,951	594	223	5,768	36
3.2	Water Makeup & Pretreating	1,515		466			1,981	238	89	2,308	14
3.3	Other Feeddwater Subsystems	2,053		903			2,956	355	133	3,444	21
3.4	Service Water System	291		154			445	53	20	518	3
3.5	Other Boiler Plant Systems	1,949		1,760			3,709	445	167	4,321	27
3.6	FO Supply System & Nat. Gas	87		110			197	24	9	230	1
3.7	Waste Treatment Equipment	1,109		560			1,669	200	75	1,944	12
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	1,172		448			1,620	194	73	1,887	12
	SUBTOTAL. 3	**11,767**	**-**	**5,761**	**-**	**-**	**17,528**	**2,103**	**789**	**20,420**	**127**
4	FLUIDIZED BED BOILER						-			-	-
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			61,101	379
4.2	Open						-			-	-
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	365		105			470	56	21	547	3
4.5	Primary Air System (Fans)	931		267			1,198	144	54	1,396	9
4.6	Secondary Air System (Fans)	53		15			68	8	3	79	0
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	**1,349**	**-**	**387**	**-**	**-**	**1,736**	**208**	**78**	**63,123**	**392**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 5 - 1x200 gr. MW Air-Fired CFB w/Carbonate Regeneration Process & CO2 Capture

Net Output Power, kW 161,184 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories						-			-	-
5.4	Other Particulate Removal Materials						-			-	-
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.1-5.6	Miscellaneous										
5.9	Gas Processing System (GPS)									51,382	319
	SUBTOTAL. 5	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**51,382**	**319**
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator						-			-	-
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	**-**	**-**	**-**	**-**	**-**	**-**				
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator						-			-	-
7.2	ID Fans	446		128			574	69	26	669	4
7.3	Ductwork	1,575		1,133			2,708	325	122	3,155	20
7.4	Stack	3,140		1,730			4,870	584	219	5,673	35
7.9	Duct & Stack Foundations		45	48			93	11	4	108	1
	SUBTOTAL. 7	**5,161**	**45**	**3,039**	**-**	**-**	**8,245**	**989**	**371**	**9,605**	**60**
8	STEAM TURBINE GENERATOR						-				-
8.1	Steam TG & Accessories	15,923		2,285			18,208	2,185	819	21,212	132
8.2	Turbine Plant Auxiliaries	110		221			331	40	15	386	2
8.3	Condenser & Auxiliaries	2,829		682			3,511	421	158	4,090	25
8.4	Steam Piping	5,601		3,212			8,813	1,058	397	10,268	64
8.9	TG Foundations		318	513			831	100	37	968	6
	SUBTOTAL. 8	**24,463**	**318**	**6,913**	**-**	**-**	**31,694**	**3,804**	**1,426**	**36,924**	**229**
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,597		1,313			3,910	469	176	4,555	28
9.2	Circulating Water Pumps	653		55			708	85	32	825	5
9.3	Circulating Water System Auxiliaries	175		22			197	24	9	230	1
9.4	Circulating Water Piping		1,291	1,264			2,555	307	115	2,977	18
9.5	Make-up Water System	148		193			341	41	15	397	2
9.6	Component Cooling Water System	128		99			227	27	10	264	2
9.9	Circ. Water System Foundations Structures		818	1,227			2,045	245	92	2,382	15
	SUBTOTAL. 9	**3,701**	**2,109**	**4,173**	**-**	**-**	**9,983**	**1,198**	**449**	**11,630**	**72**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	271		787			1,058	127	48	1,233	8
10.7	Ash Transport & Feed Equipment	3,413		1,694			5,107	613	230	5,950	37
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		129	143			272	33	10	315	2
	SUBTOTAL. 10	**3,684**	**129**	**2,624**	**-**	**-**	**6,437**	**773**	**288**	**7,498**	**47**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 5 - 1x200 gr. MW Air-Fired CFB w/Carbonate Regeneration Process & CO2 Capture

Net Output Power, kW 161,184 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	664		90			754	91	34	879	5
115.2	Station Service Equipment	1,753		481			2,234	268	101	2,603	16
11.3	Switchgear & Motor Control	1,959		278			2,237	268	101	2,606	16
11.4	Conduit & Cable Tray		832	2,496			3,328	399	150	3,877	24
11.5	Wire & Cable		965	2,584			3,549	426	160	4,135	26
11.6	Protective Equipment	113		321			434	52	20	506	3
11.7	Standby Equipment	595		11			606	73	27	706	4
11.8	Main Power Transformer	952		38			990	119	45	1,154	7
11.9	Electrical Foundations		100	235			335	40	15	390	2
	SUBTOTAL. 11	**6,036**	**1,897**	**6,534**	**-**	**-**	**14,467**	**1,736**	**653**	**16,856**	**105**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	232		116			348	42	16	406	3
12.7	Distributed Control System Equipment	2,579		376			2,955	355	133	3,443	21
12.8	Instrument Wiring & tubing	1,109		2,940			4,049	486	182	4,717	29
12.9	Other I & C Equipment	663		1,255			1,918	230	86	2,234	14
	SUBTOTAL. 12	**4,583**	**-**	**4,687**	**-**	**-**	**9,270**	**1,113**	**417**	**10,800**	**67**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	388			411	49	19	479	3
13.2	Site Improvement		766	800			1,566	188	70	1,824	11
13.3	Site Facilities	1,373		1,139			2,512	301	113	2,926	18
	SUBTOTAL. 13	**1,373**	**789**	**2,327**	**-**	**-**	**4,489**	**538**	**202**	**5,229**	**32**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,162	1,917			4,079	489	184	4,752	29
14.2	Turbine Building		3,209	3,015			6,224	747	280	7,251	45
14.3	Administration Building		412	439			851	102	38	991	6
14.4	Circulation Water Pumphouse		88	71			159	19	7	185	1
14.5	Water Treatment Building		286	238			524	63	24	611	4
14.6	Machine Shop		367	249			616	74	28	718	4
14.7	Warehouse		249	252			501	60	23	584	4
14.8	Other Buildings & Structures		152	131			283	34	13	330	2
14.9	Waste Treating Building & Structure		292	893			1,196	142	53	1,391	9
	SUBTOTAL. 14	**-**	**7,217**	**7,205**	**-**	**-**	**14,433**	**1,730**	**650**	**16,813**	**104**

Table 9.3.19: Case-5 Boiler and Balance of Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc. — **INITIAL & ANNUAL O&M EXPENSES** — Cost Base: Jul-03

Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers — **Case 5** - 1x200 gr. MW Air-Fired CFB w/Carbonate Regeneration Process & CO_2 Capture

Net Plant Heat Rate (Btu/kWh): 11,307

Net Power Output (kW): 161,184

Capacity Factor (%): 80

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	7.0	7.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	10.0	10.0

	Annual Cost $ / year	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	3,518,892	21.83
Maintenance Labor Costs (calc'd)	875,400	5.43
Administrative & Support Labor (calc'd)	1,098,573	6.82
TOTAL FIXED OPERATING COSTS	***5,492,865***	***34.08***

Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $ / year	Annual Unit Cost $/kW-net
Maintenance Material Cost (calc'd)					1,050,480	0.0009
Water (11000 gallons)		2,992	1.00		873,664	0.0008
Chemicals						
MU & WT Chem. (lbs.)	434,506	14,484	0.16	67,663	676,692	0.0006
Limestone (ton)	11,341	378.0	10.00	113,404	1,103,760	0.0010
Formic Acid (lbs.)			0.60			
Ammonia, NH3 (ton)			220			
Subtotal Chemicals				181,070	1,780,452	0.0016
Other						
Supplemental Fuel (MBtu)						
SCR Catalyst Replacement (MBtu)						
Emissions Penalties						
Subtotal Other						
Waste Disposal						
Fly Ash & Bottom Ash (ton)		823.4	8.00		1,923,462	0.0017
Subtotal Solid Waste Disposal					1,923,462	0.0017
By-Products & Emissions						
Gypsum (ton)						
Subtotal By-Products						
TOTAL VARIABLE OPERATING COST					***5,628,059***	***0.0050***

Table 9.3.20: Case-5 Gas Processing System Investment Costs

ABB LUMMUS GLOBAL HOUSTON

Project : CO2 Plant - DOE	Location : GC - USA	Project start:	Rev. : 00
Job/Prop # 0-9484	**Plant : CO2 Case 5**	Mech.compl.:	
Scope : EPC	Capacity :		
Piece count:	Labor Prod.:		1-Dec-02

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontract ($,000)	Total ($,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		8,735	135	5,532		5,668	11.0%
12000	Vessels / Filters		1,281	20	811		831	1.6%
12100	Towers / Internals		-	-	-		-	0.0%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		261	4	165		169	0.3%
14200	Compressors		14,842	230	9,400		9,630	18.7%
18000	Special Equipment		3,923	61	2,484		2,545	5.0%
	Sub-Total Equipment	22	**29,041**	**450**	**18,393**	**-**	**18,843**	**36.7%**
21000	Civil		43,561	675	1,655		2,331	4.5%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		10,164	158	828		985	1.9%
23000	Buildings		11,616	180	441		621	1.2%
30000	Piping		79,863	1,238	3,679		4,916	9.6%
40000	Electrical		41,141	638	1,471		2,109	4.1%
50000	Instruments		33,881	525	2,575		3,100	6.0%
61100	Insulation		21,781	338	552		889	1.7%
61200	Fireproofing		14,520	225	276		501	1.0%
61300	Painting		12,100	188	156		344	0.7%
	Sub-Total Commodities		**268,629**	**4,164**	**11,633**	**-**	**15,797**	**30.7%**
70000	**Construction Indirects**						**6,698**	**13.0%**
	Sub-Total Direct Cost		**297,670**	**4,614**	**30,026**	**-**	**41,337**	**80.5%**
	ASU TIC plant cost						-	0.0%
71000	Constr. Management						600	1.2%
80000	Home Office Engineering						2,244	4.4%
80000	Basic Engineering						300	0.6%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						700	1.4%
19300	Spare parts						1,200	2.3%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						100	0.2%
97000	Freight						901	1.8%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**47,382**	**92.2%**
91400	Escalation						1,400	2.7%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**48,782**	**94.9%**
	Contracters Fee						2,600	5.1%
	Grand Total						**51,382**	**100.0%**

Exclusions : Bonds,Taxes,Import duties , Hazerdous material handling & disposal, Capital spare parts, Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MCC.

Table 9.3.21: Case-5 Gas Processing System Operating and Maintenance Costs

Operating Costs ($/yr)	**Variable Costs**	**Fixed Costs**
Chemical and Dessicant	51910	
Waste Handling	0	
Fuel Gas *	131600	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	1615980	
Contracted services	830000	
Column Total	2629490	306600
Grand Total (Fixed & Variable)	2936090	

* Based on $4/ MMBU and 7000 hours/ yr.
** Included in overall facility operating cost

9.3.6. Case-6 Investment Costs and Operating and Maintenance Costs

Table 9.3.22: Case-6 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 6 - 1x200 gr. MW O2-Fired CMB w/OTM & CO2 Capture

Net Output Power, kW 197,435 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING	7,617	1,865	4,241	-	-	13,723	1,647	617	15,987	81
2	FUEL & SORBENT PREP. & FEED	4,068	228	1,178	-	-	5,474	657	246	6,377	32
3	FEEDWATER & MISC. BOP SYSTEMS	13,505	-	6,570	-	-	20,075	2,408	903	23,386	118
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									71,018	360
4.2	Oxygen Transport Membrane (OTM)									105,000	532
4.4	Open										
4.4 - 4.9	Boiler BoP (Fluidizing Air Fans)	723	-	208	-	-	931	112	41	1,084	5
	SUBTOTAL 4	723	-	208	-	-	931	112	41	177,102	897
	FLUE GAS CLEANUP										
5.1-5.6	Miscellaneous	1,947	-	1,848	-	-	3,795	456	171	4,422	22
5.9	Gas Processing System (GPS)									67,783	343
	SUBTOTAL 5	3,894	-	3,696	-	-	7,590	912	342	76,627	388
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	24,173	102	695	49	0	25,019	2,502	-	27,521	139
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	10,656	48	3,383	109	0	14,195	1,514	212	15,921	81
	SUBTOTAL 7	10,656	48	3,383	109	-	14,195	1,514	212	15,921	81
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	17,757	-	2,548	0	0	20,305	2,437	914	23,656	120
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	9,492	351	5,146	0	0	14,989	1,798	673	17,460	88
	SUBTOTAL 8	27,249	351	7,694	-	-	35,294	4,235	1,587	41,116	208
9	COOLING WATER SYSTEM	4,125	2,320	4,632	-	-	11,077	1,329	499	12,905	65
10	ASH/SPENT SORBENT HANDLING SYSTEMS	4,039	141	2,877	-	-	7,057	847	317	8,221	42
11	ACCESSORY ELECTRIC PLANT	9,513	3,457	11,848	-	-	24,818	2,979	1,116	28,913	146
12	INSTRUMENTATION & CONTROL	4,793	-	4,902	-	-	9,695	1,164	436	11,295	57
13	IMPROVEMENT TO SITE	1,465	842	2,483	-	-	4,790	576	216	5,582	28
14	BUILDINGS & STRUCTURES	-	7,723	7,698	-	-	15,421	1,851	694	17,966	91
	TOTAL COST	**115,820**	**17,077**	**62,105**	**158.00**	**-**	**195,159**	**22,733**	**7,055**	**468,919**	**2,375**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 6 - 1x200 gr. MW O2-Fired CMB w/OTM & CO2 Capture

Net Output Power, kW 197,435 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					$	$/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,328		636			1,964	236	88	2,288	12
1.2	Fuel Stackout and Reclaim	1,716		448			2,164	260	97	2,521	13
1.3	Fuel Conveyors	1,595		403			1,998	240	90	2,328	12
1.4	Other Fuel Handling	417		93			510	61	23	594	3
1.5	Sorbent Receive & Unload	97		31			128	15	6	149	1
1.6	Sorbent Stackout and Reclaim	1,567		301			1,868	224	84	2,176	11
1.7	Sorbent Conveyors	559	113	144			816	98	37	951	5
1.8	Other Sorbent Handling	338	74	186			598	72	27	697	4
1.9	Fuel & Sorbent Hnd. Foundations		1,678	1,999			3,677	441	165	4,283	22
	SUBTOTAL. 1	**7,617**	**1,865**	**4,241**	**-**	**-**	**13,723**	**1,647**	**617**	**15,987**	**81**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	739		151			890	107	40	1,037	5
2.2	Fuel Conveyor to Storage	2,367		542			2,909	349	131	3,389	17
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	642		140			782	94	35	911	5
2.6	Sorbent Storage & Feed	320		128			448	54	20	522	3
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		228	217			445	53	20	518	3
	SUBTOTAL. 2	**4,068**	**228**	**1,178**	**-**	**-**	**5,474**	**657**	**246**	**6,377**	**32**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	3,805		1,369			5,174	621	233	6,028	31
3.2	Water Makeup & Pretreating	1,880		578			2,458	295	111	2,864	15
3.3	Other Feeddwater Subsystems	2,053		903			2,956	355	133	3,444	17
3.4	Service Water System	361		192			553	66	25	644	3
3.5	Other Boiler Plant Systems	2,279		2,058			4,337	520	195	5,052	26
3.6	FO Supply System & Nat. Gas	121		153			274	33	12	319	2
3.7	Waste Treatment Equipment	1,376		694			2,070	248	93	2,411	12
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	1,630		623			2,253	270	101	2,624	13
	SUBTOTAL. 3	**13,505**	**-**	**6,570**	**-**	**-**	**20,075**	**2,408**	**903**	**23,386**	**118**
4	FLUIDIZED BED BOILER						-			-	-
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			71,018	360
4.2	Oxygen Transport Membrane (OTM)						-			105,000	532
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	52		15			67	8	3	78	0
4.5	Primary Air System (Fans)	181		52			233	28	10	271	1
4.6	Secondary Air System (Fans)	490		141			631	76	28	735	4
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	**723**	**-**	**208**	**-**	**-**	**931**	**112**	**41**	**177,102**	**897**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 6 - 1x200 gr. MW O2-Fired CMB w/OTM & CO2 Capture

Net Output Power, kW 197,435 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories	1,686		1,585			3,271	393	147	3,811	19
5.4	Other Particulate Removal Materials	261		263			524	63	24	611	3
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.1-5.6	Miscellaneous	1,947		1,848			3,795	456	171	4,422	22
5.9	Gas Processing System (GPS)									67,783	343
	SUBTOTAL. 5	**3,894**	**-**	**3,696**	**-**	**-**	**7,590**	**912**	**342**	**76,627**	**388**
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator	24,173		595	42		24,810	2,481		27,291	138
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations		102	100	7		209	21		230	1
	SUBTOTAL. 6	**24,173**	**102**	**695**	**49**	**-**	**25,019**	**2,502**		**27,521**	**139**
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator	7,807		1,561	109		9,477	948		10,425	53
7.2	ID Fans	319		91			410	49	18	477	2
7.3	Ductwork	1,690		1,216			2,906	349	131	3,386	17
7.4	Stack	840		463			1,302	156	59	1,517	8
7.9	Duct & Stack Foundations		48	52			100	12	4	116	1
	SUBTOTAL. 7	**10,656**	**48**	**3,383**	**109**	**-**	**14,195**	**1,514**	**212**	**15,921**	**81**
8	STEAM TURBINE GENERATOR						-				-
8.1	Steam TG & Accessories	17,757		2,548			20,305	2,437	914	23,656	120
8.2	Turbine Plant Auxiliaries	121		245			366	44	16	426	2
8.3	Condenser & Auxiliaries	3,124		753			3,877	465	174	4,516	23
8.4	Steam Piping	6,247		3,582			9,829	1,179	442	11,450	58
8.9	TG Foundations		351	566			917	110	41	1,068	5
	SUBTOTAL. 8	**27,249**	**351**	**7,694**	**-**	**-**	**35,294**	**4,235**	**1,587**	**41,116**	**208**
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,856		1,444			4,300	516	194	5,010	25
9.2	Circulating Water Pumps	719		60			779	93	35	907	5
9.3	Circulating Water System Auxiliaries	192		24			216	26	10	252	1
9.4	Circulating Water Piping		1,420	1,390			2,810	337	126	3,273	17
9.5	Make-up Water System	163		213			376	45	17	438	2
9.6	Component Cooling Water System	195		151			346	42	16	404	2
9.9	Circ. Water System Foundations Structures		900	1,350			2,250	270	101	2,621	13
	SUBTOTAL. 9	**4,125**	**2,320**	**4,632**	**-**	**-**	**11,077**	**1,329**	**499**	**12,905**	**65**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	297		863			1,160	139	52	1,351	7
10.7	Ash Transport & Feed Equipment	3,742		1,857			5,599	672	252	6,523	33
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		141	157			298	36	13	347	2
	SUBTOTAL. 10	**4,039**	**141**	**2,877**	**-**	**-**	**7,057**	**847**	**317**	**8,221**	**42**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers
Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 6 - 1x200 gr. MW O2-Fired CMB w/OTM & CO2 Capture

Net Output Power, kW 197,435 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	719		97			816	98	37	951	5
115.2	Station Service Equipment	3,265		896			4,161	499	187	4,847	25
11.3	Switchgear & Motor Control	3,647		518			4,165	500	187	4,852	25
11.4	Conduit & Cable Tray		1,549	4,649			6,198	744	279	7,221	37
11.5	Wire & Cable		1,798	4,813			6,611	793	297	7,701	39
11.6	Protective Equipment	199		564			763	92	34	889	5
11.7	Standby Equipment	635		12			647	78	29	754	4
11.8	Main Power Transformer	1,048		41			1,089	131	49	1,269	6
11.9	Electrical Foundations		110	258			368	44	17	429	2
	SUBTOTAL. 11	**9,513**	**3,457**	**11,848**	**-**	**-**	**24,818**	**2,979**	**1,116**	**28,913**	**146**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	243		121			364	44	16	424	2
12.7	Distributed Control System Equipment	2,697		394			3,091	371	139	3,601	18
12.8	Instrument Wiring & tubing	1,160		3,074			4,234	508	191	4,933	25
12.9	Other I & C Equipment	693		1,313			2,006	241	90	2,337	12
	SUBTOTAL. 12	**4,793**	**-**	**4,902**	**-**	**-**	**9,695**	**1,164**	**436**	**11,295**	**57**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		25	414			439	53	20	512	3
13.2	Site Improvement		817	854			1,671	201	75	1,947	10
13.3	Site Facilities	1,465		1,215			2,680	322	121	3,123	16
	SUBTOTAL. 13	**1,465**	**842**	**2,483**	**-**	**-**	**4,790**	**576**	**216**	**5,582**	**28**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,311	2,050			4,361	523	196	5,080	26
14.2	Turbine Building		3,482	3,272			6,754	810	304	7,868	40
14.3	Administration Building		430	459			889	107	40	1,036	5
14.4	Circulation Water Pumphouse		92	74			166	20	7	193	1
14.5	Water Treatment Building		300	249			549	66	25	640	3
14.6	Machine Shop		384	260			644	77	29	750	4
14.7	Warehouse		260	263			523	63	24	610	3
14.8	Other Buildings & Structures		159	137			296	36	13	345	2
14.9	Waste Treating Building & Structure		305	934			1,239	149	56	1,444	7
	SUBTOTAL. 14	**-**	**7,723**	**7,698**	**-**	**-**	**15,421**	**1,851**	**694**	**17,966**	**91**

Table 9.3.23: Case-6 Boiler and Balance of Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Case 6 - 1x200 gr. MW O2-Fired CMB w/OTM & CO2 Capture

Cost Base: Jul-03
Net Plant Heat Rate (Btu/kWh): 11,380
Net Power Output (kW): 197,435
Capacity Factor (%): 80

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	7.0	7.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	10.0	10.0

	Annual Cost $ / year	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	3,518,892	17.82
Maintenance Labor Costs (calc'd)	1,184,544	6.00
Administrative & Support Labor (calc'd)	1,175,859	5.96
TOTAL FIXED OPERATING COSTS	***5,879,295***	***29.78***

	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $ / year	Annual Unit Cost $/kW-net
Maintenance Material Cost (calc'd)					1,421,453	0.0010
Consumables						
Water (1000 gallons)		4,055	1.00		1,184,060	0.0009
Chemicals						
MU & WT Chem. (lbs.)	588,917	19,631	0.16	91,708	917,160	0.0007
Limestone (ton)	12,610	420.3	10.00	126,101	1,227,276	0.0009
Formic Acid (lbs.)			0.60			
Ammonia, NH3 (ton)			220			
Subtotal Chemicals				217,809	2,144,436	0.0015
Other						
Supplemental Fuel (MBtu)						
SCR Catalyst Replacement (MBtu)						
Emissions Penalties						
Subtotal Other						
Waste Disposal						
Fly Ash & Bottom Ash (ton)		971.9	8.00		2,270,358	0.0016
Subtotal Solid Waste Disposal					2,270,358	0.0016
By-Products & Emissions						
Gypsum (ton)						
Subtotal By-Products						
TOTAL VARIABLE OPERATING COST					***7,020,308***	***0.0051***

Table 9.3.24: Case-6 Gas Processing System Investment Costs

ABB LUMMUS GLOBAL HOUSTON

Project	CO2 Plant - DOE	Location	: GC - USA	Project start:	Rev. : 00
Job/Prop #	0-9484	Plant	: CO2 Case 6	Mech.compl.:	
Scope	EPC	Capacity	:		
Piece count:		Labor Prod.:			21-Feb-03

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontrac ($,000)	Total ($,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		11,800	183	7,473		7,656	11.3%
12000	Vessels / Filters		2,443	38	1,547		1,585	2.3%
12100	Towers / Internals		1,927	30	1,220		1,250	1.8%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		390	6	247		253	0.4%
14200	Compressors		20,495	318	12,980		13,298	19.6%
18000	Special Equipment		-	-	-		-	0.0%
	Sub-Total Equipment	45	**37,055**	**574**	**23,468**	**-**	**24,042**	**35.5%**
21000	Civil		55,582	862	2,112		2,974	4.4%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		12,969	201	1,056		1,257	1.9%
23000	Buildings		14,822	230	563		793	1.2%
30000	Piping		101,900	1,579	4,694		6,273	9.3%
40000	Electrical		52,494	814	1,877		2,691	4.0%
50000	Instruments		43,230	670	3,286		3,956	5.8%
61100	Insulation		27,791	431	704		1,135	1.7%
61200	Fireproofing		18,527	287	352		639	0.9%
61300	Painting		15,439	239	199		439	0.6%
	Sub-Total Commodities		**342,755**	**5,313**	**14,843**	**-**	**20,156**	**29.7%**
70000	**Construction Indirects**						**8,546**	**12.6%**
	Sub-Total Direct Cost		**379,810**	**5,887**	**38,311**	**-**	**52,744**	**77.8%**
	ASU TIC plant cost						-	0.0%
71000	Constr. Management						800	1.2%
80000	Home Office Engineering						4,590	6.8%
80000	Basic Engineering						600	0.9%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						1,000	1.5%
19300	Spare parts						1,600	2.4%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						100	0.1%
97000	Freight						1,149	1.7%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**62,583**	**92.3%**
91400	Escalation						1,800	2.7%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**64,383**	**95.0%**
	Contracters Fee						3,400	5.0%
	Grand Total						**67,783**	**100.0%**

Exclusions : Bonds,Taxes,Import duties , Hazerdous material handling & disposal, Capital spare parts,
Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MCC.

Table 9.3.25: Case-6 Gas Processing System Operating and Maintenance Costs

Operating Costs ($/yr)	Variable Costs	Fixed Costs
Chemical and Dessicant	27685	
Waste Handling	0	
Fuel Gas *	122122	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	2033490	
Contracted services	830000	
Column Total	3013297	306600
Grand Total (Fixed & Variable)	3319897	

* Based on $4/ MMBU and 7000 hours/ yr.
** Included in overall facility operating cost

9.3.7. Case-7 Investment Costs and Operating and Maintenance Costs

Table 9.3.26: Case-7 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 7 - 1X200 gr. MW- CMB Chemical Looping w/ CO2 Capture

Net Output Power, kW 164,484 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING	6,521	1,629	3,686	-	-	11,836	1,419	534	13,789	84
2	FUEL & SORBENT PREP. & FEED	3,487	184	996	-	-	4,667	560	209	5,436	33
3	FEEDWATER & MISC. BOP SYSTEMS	11,989	-	5,853	-	-	17,842	2,140	804	20,786	126
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									66,442	404
4.2	Open										
4.4	Open										
4.4 - 4.9	Boiler BoP (Fluidizing Air Fans)	1,331	-	382	-	-	1,713	205	77	1,995	12
	SUBTOTAL 4	1,331	-	382	-	-	1,713	205	77	68,437	416
5	FLUE GAS CLEANUP	-	-	-	-	-	-	-	-		
5.9	Gas Processing System (GPS)									55,117	335
	SUBTOTAL 5	-	-	-	-	-	-	-	-	55,117	335
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	-	-	-	0	0	-	-	-	-	0
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	2,726	42	1,688	-	0	4,456	535	201	5,191	32
	SUBTOTAL 7	2,726	42	1,688	-	-	4,456	535	201	5,191	32
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	15,929	-	2,285	0	0	18,214	2,186	820	21,220	129
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,543	318	4,631	0	0	13,492	1,619	607	15,718	96
	SUBTOTAL 8	24,472	318	6,916	-	-	31,706	3,805	1,427	36,938	225
9	COOLING WATER SYSTEM	3,702	2,109	4,174	-	-	9,985	1,199	449	11,633	71
10	ASH/SPENT SORBENT HANDLING SYSTEMS	3,526	123	2,511	-	-	6,160	739	278	7,177	44
11	ACCESSORY ELECTRIC PLANT	5,850	1,809	6,232	-	-	13,891	1,668	626	16,185	98
12	INSTRUMENTATION & CONTROL	4,601	-	4,706	-	-	9,307	1,117	420	10,844	66
13	IMPROVEMENT TO SITE	1,373	789	2,327	-	-	4,489	538	203	5,230	32
14	BUILDINGS & STRUCTURES	-	7,219	7,205	-	-	14,424	1,731	650	16,805	102
	TOTAL COST	**69,578**	**14,222**	**46,676**	**-**	**-**	**130,476**	**15,656**	**5,878**	**273,568**	**1,663**

Client: ALSTOM Power Inc. **Report Date:** 4/24/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 7 - 1X200 gr. MW- CMB Chemical Looping w/ CO2 Capture

Net Output Power, kW 164,484 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,169		560			1,729	207	78	2,014	12
1.2	Fuel Stackout and Reclaim	1,511		395			1,906	229	86	2,221	14
1.3	Fuel Conveyors	1,404		355			1,759	211	79	2,049	12
1.4	Other Fuel Handling	367		82			449	54	20	523	3
1.5	Sorbent Receive & Unload	78		25			103	12	5	120	1
1.6	Sorbent Stackout and Reclaim	1,267		244			1,511	181	68	1,760	11
1.7	Sorbent Conveyors	452	92	116			660	79	30	769	5
1.8	Other Sorbent Handling	273	60	150			483	58	22	563	3
1.9	Fuel & Sorbent Hnd. Foundations		1,477	1,759			3,236	388	146	3,770	23
	SUBTOTAL. 1	**6,521**	**1,629**	**3,686**	**-**	**-**	**11,836**	**1,419**	**534**	**13,789**	**84**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	646		132			778	93	35	906	6
2.2	Fuel Conveyor to Storage	2,066		473			2,539	305	114	2,958	18
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	517		113			630	76	28	734	4
2.6	Sorbent Storage & Feed	258		103			361	43	16	420	3
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		184	175			359	43	16	418	3
	SUBTOTAL. 2	**3,487**	**184**	**996**	**-**	**-**	**4,667**	**560**	**209**	**5,436**	**33**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	3,591		1,360			4,951	594	223	5,768	35
3.2	Water Makeup & Pretreating	1,627		501			2,128	255	96	2,479	15
3.3	Other Feeddwater Subsystems	2,053		903			2,956	355	133	3,444	21
3.4	Service Water System	313		166			479	57	22	558	3
3.5	Other Boiler Plant Systems	1,953		1,764			3,717	446	167	4,330	26
3.6	FO Supply System & Nat. Gas	87		110			197	24	9	230	1
3.7	Waste Treatment Equipment	1,192		601			1,793	215	81	2,089	13
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	1,173		448			1,621	194	73	1,888	11
	SUBTOTAL. 3	**11,989**	**-**	**5,853**	**-**	**-**	**17,842**	**2,140**	**804**	**20,786**	**126**
4	FLUIDIZED BED BOILER						-			-	-
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			66,442	404
4.2	Open						-			-	-
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	365		105			470	56	21	547	3
4.5	Primary Air System (Fans)	914		262			1,176	141	53	1,370	8
4.6	Secondary Air System (Fans)	52		15			67	8	3	78	0
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	**1,331**	**-**	**382**	**-**	**-**	**1,713**	**205**	**77**	**68,437**	**416**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 7 - 1X200 gr. MW- CMB Chemical Looping w/ CO2 Capture

Net Output Power, kW 164,484 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories						-			-	-
5.4	Other Particulate Removal Materials						-			-	-
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.9	Gas Processing System (GPS)									55,117	335
	SUBTOTAL. 5	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**55,117**	**335**
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator						-			-	-
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	**-**	**-**	**-**	**-**	**-**	**-**				
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator						-			-	-
7.2	ID Fans	408		117			525	63	24	612	4
7.3	Ductwork	1,478		1,064			2,542	305	114	2,961	18
7.4	Stack	840		462			1,302	157	59	1,517	9
7.9	Duct & Stack Foundations		42	45			87	10	4	101	1
	SUBTOTAL. 7	**2,726**	**42**	**1,688**	**-**	**-**	**4,456**	**535**	**201**	**5,191**	**32**
8	STEAM TURBINE GENERATOR						-				-
8.1	Steam TG & Accessories	15,929		2,285			18,214	2,186	820	21,220	129
8.2	Turbine Plant Auxiliaries	110		222			332	40	15	387	2
8.3	Condenser & Auxiliaries	2,829		682			3,511	421	158	4,090	25
8.4	Steam Piping	5,604		3,214			8,818	1,058	397	10,273	62
8.9	TG Foundations		318	513			831	100	37	968	6
	SUBTOTAL. 8	**24,472**	**318**	**6,916**	**-**	**-**	**31,706**	**3,805**	**1,427**	**36,938**	**225**
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,597		1,313			3,910	469	176	4,555	28
9.2	Circulating Water Pumps	653		55			708	85	32	825	5
9.3	Circulating Water System Auxiliaries	175		22			197	24	9	230	1
9.4	Circulating Water Piping		1,291	1,264			2,555	307	115	2,977	18
9.5	Make-up Water System	148		193			341	41	15	397	2
9.6	Component Cooling Water System	129		99			228	27	10	265	2
9.9	Circ. Water System Foundations Structures		818	1,228			2,046	246	92	2,384	14
	SUBTOTAL. 9	**3,702**	**2,109**	**4,174**	**-**	**-**	**9,985**	**1,199**	**449**	**11,633**	**71**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	260		753			1,013	122	46	1,181	7
10.7	Ash Transport & Feed Equipment	3,266		1,621			4,887	586	220	5,693	35
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		123	137			260	31	12	303	2
	SUBTOTAL. 10	**3,526**	**123**	**2,511**	**-**	**-**	**6,160**	**739**	**278**	**7,177**	**44**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 7 - 1X200 gr. MW- CMB Chemical Looping w/ CO2 Capture

Net Output Power, kW 164,484 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	665		90			755	91	34	880	5
115.2	Station Service Equipment	1,667		457			2,124	255	96	2,475	15
11.3	Switchgear & Motor Control	1,862		264			2,126	255	96	2,477	15
11.4	Conduit & Cable Tray		791	2,373			3,164	380	142	3,686	22
11.5	Wire & Cable		918	2,457			3,375	405	152	3,932	24
11.6	Protective Equipment	108		307			415	50	19	484	3
11.7	Standby Equipment	595		11			606	73	27	706	4
11.8	Main Power Transformer	953		38			991	119	45	1,155	7
11.9	Electrical Foundations		100	235			335	40	15	390	2
	SUBTOTAL. 11	**5,850**	**1,809**	**6,232**	**-**	**-**	**13,891**	**1,668**	**626**	**16,185**	**98**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	233		117			350	42	16	408	2
12.7	Distributed Control System Equipment	2,589		378			2,967	356	134	3,457	21
12.8	Instrument Wiring & tubing	1,114		2,951			4,065	488	183	4,736	29
12.9	Other I & C Equipment	665		1,260			1,925	231	87	2,243	14
	SUBTOTAL. 12	**4,601**	**-**	**4,706**	**-**	**-**	**9,307**	**1,117**	**420**	**10,844**	**66**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	388			411	49	19	479	3
13.2	Site Improvement		766	800			1,566	188	71	1,825	11
13.3	Site Facilities	1,373		1,139			2,512	301	113	2,926	18
	SUBTOTAL. 13	**1,373**	**789**	**2,327**	**-**	**-**	**4,489**	**538**	**203**	**5,230**	**32**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,162	1,917			4,079	490	184	4,753	29
14.2	Turbine Building		3,210	3,015			6,225	747	280	7,252	44
14.3	Administration Building		412	439			851	102	38	991	6
14.4	Circulation Water Pumphouse		88	71			159	19	7	185	1
14.5	Water Treatment Building		287	238			525	63	24	612	4
14.6	Machine Shop		367	249			616	74	28	718	4
14.7	Warehouse		249	252			501	60	23	584	4
14.8	Other Buildings & Structures		152	131			283	34	13	330	2
14.9	Waste Treating Building & Structure		292	893			1,185	142	53	1,380	8
	SUBTOTAL. 14	**-**	**7,219**	**7,205**	**-**	**-**	**14,424**	**1,731**	**650**	**16,805**	**102**

Table 9.3.27: Case-7 Boiler and Balance of Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc. — **INITIAL & ANNUAL O&M EXPENSES** — Cost Base: Jul-03

Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Case 7 - 1x200 gr. MW CMB Chemical Looping w/CO2 Capture

Net Plant Heat Rate (Btu/kWh): 11,051

Net Power Output (kW): 164,484

Capacity Factor (%): 80

						Annual Cost	Annual Unit Cost
OPERATING & MAINTENANCE LABOR							
Operating Labor							
Operating Labor Rate (Base):	30.90 $/hour						
Operating Labor Burden:	30.00 %						
Labor O-H change Rate:	25.00 %						
Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant					
Skilled Operator	1.0	1.0					
Operator	7.0	7.0					
Foreman	1.0	1.0					
Lab Tech's, etc.	1.0	1.0					
TOTAL O.J.'s	10.0	10.0					
						$ / year	$/kW-net
Annual Operating Labor Costs (calc'd)						3,518,892	21.39
Maintenance Labor Costs (calc'd)						873,805	5.31
Administrative & Support Labor (calc'd)						1,098,174	6.68
TOTAL FIXED OPERATING COSTS						***5,490,871***	***33.38***
Maintenance Material Cost (calc'd)						1,048,566	0.0009
Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost			
Water (11000 gallons)		3,311	1.00			966,812	0.0008
Chemicals							
MU & WT Chem. (lbs.)	480,763	16,025	0.16	74,866		748,688	0.0006
Limestone (ton)	9,049	301.6	10.00	90,486		880,672	0.0008
Formic Acid (lbs.)			0.60				
Ammonia, NH3 (ton)			220				
Subtotal Chemicals				165,352		1,629,360	0.0014
Other							
Supplemental Fuel (MBtu)							
SCR Catalyst Replacement (MBtu)							
Emissions Penalties							
Subtotal Other							
Waste Disposal							
Fly Ash & Bottom Ash (ton)		760.6	8.00			1,776,762	0.0015
Subtotal Solid Waste Disposal						1,776,762	0.0015
By-Products & Emissions							
Gypsum (ton)							
Subtotal By-Products							
TOTAL VARIABLE OPERATING COST						***5,421,500***	***0.0047***

Table 9.3.28: Case-7 Gas Processing System Investment Costs

ABB LUMMUS GLOBAL HOUSTON

ABB

Project : CO2 Plant - DOE	Location : GC - USA	Project start:	Rev. : 00
Job/Prop # : 0-9484	**Plant : CO2 Case 7**	Mech.compl.:	
Scope : EPC	Capacity :		
Piece count:	Labor Prod.:		1-Dec-02

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontract ($,000)	Total ($,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		8,072	125	5,112		5,237	9.5%
12000	Vessels / Filters		1,353	21	857		878	1.6%
12100	Towers / Internals		1,099	17	696		713	1.3%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		377	6	239		244	0.4%
14200	Compressors		15,947	247	10,100		10,347	18.8%
18000	Special Equipment		3,648	57	2,311		2,367	4.3%
	Sub-Total Equipment	31	**30,496**	**473**	**19,314**	-	**19,787**	**35.9%**
21000	Civil		45,744	709	1,738		2,447	4.4%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		10,674	165	869		1,035	1.9%
23000	Buildings		12,198	189	464		653	1.2%
30000	Piping		83,864	1,300	3,863		5,163	9.4%
40000	Electrical		43,203	670	1,545		2,215	4.0%
50000	Instruments		35,579	551	2,704		3,255	5.9%
61100	Insulation		22,872	355	579		934	1.7%
61200	Fireproofing		15,248	236	290		526	1.0%
61300	Painting		12,707	197	164		361	0.7%
	Sub-Total Commodities		**282,089**	**4,372**	**12,216**	-	**16,589**	**30.1%**
70000	**Construction Indirects**						**7,033**	**12.8%**
	Sub-Total Direct Cost		**312,585**	**4,845**	**31,530**	-	**43,409**	**78.8%**
	ASU TIC plant cost						-	0.0%
71000	Constr. Management						700	1.3%
80000	Home Office Engineering						3,162	5.7%
80000	Basic Engineering						400	0.7%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						800	1.5%
19300	Spare parts						1,300	2.4%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						100	0.2%
97000	Freight						946	1.7%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**50,817**	**92.2%**
91400	Escalation						1,500	2.7%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**52,317**	**94.9%**
	Contracters Fee						2,800	5.1%
	Grand Total						**55,117**	**100.0%**

Exclusions : Bonds,Taxes,Import duties , Hazerdous material handling & disposal, Capital spare parts,
Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MCC.

Table 9.3.29: Case-7 Gas Processing System Operating and Maintenance Costs

Operating Costs ($/yr)	**Variable Costs**	**Fixed Costs**
Chemical and Dessicant	38067	
Waste Handling	0	
Fuel Gas *	109200	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	1615980	
Contracted services	830000	
Column Total	2593247	306600
Grand Total (Fixed & Variable)	2899847	

* Based on $4/ MMBU and 7000 hours/ yr.
** Included in overall facility operating cost

9.3.8. Case-8 Investment Costs and Operating and Maintenance Costs

Table 9.3.30: Case-8 Overall Power Plant Investment Costs

		Case-8	
Act Number	Account Title	($x1000)	$/kW
1	Coal Receiving and Handling	15,310	58
2	Coal Preparation and Feed	14,661	56
3	Feedwater & Miscellaneous BOP Systems	15,540	59
4	Gasifier & Accessories	116,250	442
4a	Air Separation Unit	33,992	129
5a	Shift, Gas Cooling, Humidification and Acid Gas Removal	31,232	119
5b	CO2 Compression, Purification, and Liquefaction	0	0
6	Combustion Turbine & Auxiliaries	49,623	189
7	Heat Recovery Boiler & Stack	21,418	81
8	Steam Turbine Generator	22,289	85
9	Cooling Water System	16,030	61
10	Slag Recovery & Handling	23,186	88
11	Accessory Electric Plant	23,672	90
12	I&C	11,371	43
13	Site Improvements	5,309	20
14	Buildings & Structures	11,848	45
	Total Plant Cost	411,731	1,565

Table 9.3.31: Case-8 Overall Power Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Case 8 Built & Operating IGCC w/o CO_2 Capture

Cost Base: Jul-03
Net Plant Heat Rate (Btu/kWh): 9,069
Net Power Output (kW): 263,087
Capacity Factor (%): 80

GASIFIER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	9.0	9.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	12.0	12.0

	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)					4,222,670	16.05
Maintenance Labor Costs (calc'd)					3,921,569	14.91
Administrative & Support Labor (calc'd)					2,036,060	7.74
TOTAL FIXED OPERATING COSTS					***10,180,299***	***38.70***
Maintenace Materisl Cost (calc'd)					4,705,882	0.0026
Consumables						
Water (1000 gallons)		2,926	1.00		854,392	0.0005
Chemicals						
Makeup Chemicals & Catalysts					705,882	0.0004
Subtotal Chemicals					705,882	0.0004
Waste Disposal						
Slag Disposal		631.38	8.00		1,474,904	0.000800
Catalyst Disposal					4,706	0.000003
Subtotal Solid Waste Disposal					1,479,610	0.000803
TOTAL VARIABLE OPERATING COST					***7,745,766***	***0.0042***

9.3.9. Case-9 Investment Costs and Operating and Maintenance Costs

Table 9.3.32: Case-9 Overall Power Plant Investment Costs

		Case-9	
Act Number	Account Title	($x1000)	$/kW
1	Coal Receiving and Handling	16,961	74
2	Coal Preparation and Feed	16,243	70
3	Feedwater & Miscellaneous BOP Systems	15,554	67
4	Gasifier & Accessories	114,358	496
4a	Air Separation Unit	36,951	160
5a	Shift, Gas Cooling, Humidification and Acid Gas Removal	62,875	273
5b	CO2 Compression, Purification, and Liquefaction	52,418	227
6	Combustion Turbine & Auxiliaries	49,670	215
7	Heat Recovery Boiler & Stack	21,439	93
8	Steam Turbine Generator	22,014	96
9	Cooling Water System	16,526	72
10	Slag Recovery & Handling	25,204	109
11	Accessory Electric Plant	23,562	102
12	I&C	11,381	49
13	Site Improvements	5,314	23
14	Buildings & Structures	11,859	51
	Total Plant Cost	502,330	2,179

Table 9.3.33: Case-9 Overall Power Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Cost Base: Jul-03

Case 9 Built & Operating IGCC w/ CO_2 Capture

Net Plant Heat Rate (Btu/kW 11,467

Net Power Output (kV 230,515

Built & Operating IGCC w/ CO_2 Capture

Capacity Factor (% 80

GASIFIER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	11.0	11.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	14.0	14.0

	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)					4,926,449	21.37
Maintenance Labor Costs (calc'd)					4,784,487	20.76
Administrative & Support Labor (calc'd)					2,427,734	10.53
TOTAL FIXED OPERATING COSTS					***12,138,670***	***52.66***
Maintenace Materisl Cost (calc'd)					5,741,384	0.0036
Consumables						
Water (1000 gallons)		3,000	1.00		876,000	0.0005
Chemicals						
Makeup Chemicals & Catalysts					941,176	0.0006
Subtotal Chemicals					941,176	0.0006
Waste Disposal						
Slag Disposal		699.48	8.00		1,633,985	0.001011
Catalyst Disposal					9,412	0.000006
Subtotal Solid Waste Disposal					1,643,397	0.001017
TOTAL VARIABLE OPERATING COST					***9,201,958***	***0.0057***

9.3.10. Case-10 Investment Costs and Operating and Maintenance Costs

Table 9.3.34: Case-10 Overall Power Plant Investment Costs

		Case-10	
Act Number	Account Title	($x1000)	$/kW
1	Coal Receiving and Handling	15,619	66
2	Coal Preparation and Feed	13,595	58
3	Feedwater & Miscellaneous BOP Systems	11,921	51
4	Gasifier & Accessories	62,842	267
4a	Air Separation Unit	32,357	138
5a	Shift, Gas Cooling, Humidification and Acid Gas Removal	36,189	154
5b	CO2 Compression, Purification, and Liquefaction	0	0
6	Combustion Turbine & Auxiliaries	53,307	227
7	Heat Recovery Boiler & Stack	18,479	79
8	Steam Turbine Generator	19,193	82
9	Cooling Water System	12,197	52
10	Slag Recovery & Handling	16,843	72
11	Accessory Electric Plant	22,953	98
12	I&C	10,749	46
13	Site Improvements	4,753	20
14	Buildings & Structures	10,471	45
	Total Plant Cost	341,468	1,451

Table 9.3.35: Case-10 Overall Power Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Case10 Commercially Offered IGCC w/o CO_2 Capture

Cost Base: Jul-03

Net Plant Heat Rate (Btu/kWh): 9,884
Net Power Output (kW): 235,294
Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor		
Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	9.0	9.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	12.0	12.0

	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)					4,222,670	17.95
Maintenance Labor Costs (calc'd)					3,252,342	13.82
Administrative & Support Labor (calc'd)					1,868,753	7.94
TOTAL FIXED OPERATING COSTS					***9,343,766***	***39.71***
Maintenace Materisl Cost (calc'd)					3,902,811	0.0024
Consumables						
Water (1000 gallons)		2,926	1.00		854,392	0.0005
Chemicals						
Makeup Chemicals & Catalysts					705,882	0.0004
Subtotal Chemicals					705,882	0.0004
Waste Disposal						
Slag Disposal		613.008	8.00		1,431,987	0.000868
Catalyst Disposal					4,706	0.000003
Subtotal Solid Waste Disposal					1,436,693	0.000871
TOTAL VARIABLE OPERATING COST					***6,899,778***	***0.0042***

9.3.11. Case-11 Investment Costs and Operating and Maintenance Costs

Table 9.3.36: Case-11 Overall Power Plant Investment Costs

		Case-11	
Act Number	Account Title	($x1000)	$/kW
1	Coal Receiving and Handling	16,795	84
2	Coal Preparation and Feed	14,619	73
3	Feedwater & Miscellaneous BOP Systems	11,976	60
4	Gasifier & Accessories	62,692	312
4a	Air Separation Unit	34,387	171
5a	Shift, Gas Cooling, Humidification and Acid Gas Removal	48,581	242
5b	CO2 Compression, Purification, and Liquefaction	49,587	247
6	Combustion Turbine & Auxiliaries	55,476	276
7	Heat Recovery Boiler & Stack	18,563	92
8	Steam Turbine Generator	18,924	94
9	Cooling Water System	12,254	61
10	Slag Recovery & Handling	17,994	90
11	Accessory Electric Plant	22,838	114
12	I&C	11,461	57
13	Site Improvements	5,066	25
14	Buildings & Structures	11,164	56
	Total Plant Cost	412,377	2,052

Table 9.3.37: Case-11 Overall Power Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

INITIAL & ANNUAL O&M EXPENSES

Case 11 Commercially Offered IGCC w/o CO_2 Capture

Cost Base: Jul-03

Net Plant Heat Rate (Btu/kWh): 12,441
Net Power Output (kW): 201,004
Capacity Factor (%): 80

BOILER ISLAND AND BALANCE OF PLANT O&M COSTS

OPERATING & MAINTENANCE LABOR

Operating Labor		
Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	11.0	11.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	14.0	14.0

	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	4,926,449	24.51
Maintenance Labor Costs (calc'd)	3,927,722	19.54
Administrative & Support Labor (calc'd)	2,213,543	11.01
TOTAL FIXED OPERATING COSTS	***11,067,713***	***55.06***
Maintenace Materisl Cost (calc'd)	4,713,266	0.0033

Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $	Annual Unit Cost $/kW-net
Water (1000 gallons)		3,000	1.00		876,000	0.0006
Chemicals						
Makeup Chemicals & Catalysts					941,176	0.0007
Subtotal Chemicals					941,176	0.0007
Waste Disposal						
Slag Disposal		701.664	8.00		1,639,087	0.001164
Catalyst Disposal					941,176	0.000668
Subtotal Solid Waste Disposal					2,580,264	0.001832
TOTAL VARIABLE OPERATING COST					***9,110,706***	***0.0065***

9.3.12. Case-12 Investment Costs and Operating and Maintenance Costs

Table 9.3.38: Case-12 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 12 - CMB Chemical Looping Gasification w/o CO2 Capture

Net Output Power, kW 265,146 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					$	$/kW
1	FUEL & SORBENT HANDLING	7,893	1,972	4,461	-	-	14,326	1,718	644	16,688	63
2	FUEL & SORBENT PREP. & FEED	4,220	222	1,206	-	-	5,648	677	254	6,579	25
3	FEEDWATER & MISC. BOP SYSTEMS	5,035	-	2,458	-	-	7,493	899	340	8,732	33
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									51,257	193
4.2	Open										
4.4	Open										
4.4 - 4.9	Boiler BoP (Fluidizing Air Fans)	1,331	-	382	-	-	1,713	205	77	1,995	8
	SUBTOTAL 4	1,331	-	382	-	-	1,713	205	77	53,252	201
5	FLUE GAS CLEANUP	-	-	-	-	-	-	-	-		
5.9	Gas Processing System (GPS)									-	-
	SUBTOTAL 5	-	-	-	-	-	-	-	-	-	-
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	92,732	-	1,824	0	0	94,556	5,533	1,651	102,332	386
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	16,509	42	4,452	-	0	20,973	2,517	944	24,434	92
	SUBTOTAL 7	16,509	42	4,452	-	-	20,973	2,517	944	24,434	92
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	6,690	-	960	0	0	7,650	918	344	8,912	34
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	3,588	134	1,945	0	0	5,667	680	255	6,602	25
	SUBTOTAL 8	10,278	134	2,905	-	-	13,317	1,598	599	15,514	59
9	COOLING WATER SYSTEM	1,667	949	1,880	-	-	4,496	538	202	5,236	20
10	ASH/SPENT SORBENT HANDLING SYSTEMS	4,267	149	3,038	-	-	7,454	895	335	8,684	33
11	ACCESSORY ELECTRIC PLANT	8,189	2,536	8,726	-	-	19,451	2,336	875	22,662	85
12	INSTRUMENTATION & CONTROL	4,601	-	4,706	-	-	9,307	1,117	419	10,843	41
13	IMPROVEMENT TO SITE	1,373	789	2,327	-	-	4,489	538	203	5,230	20
14	BUILDINGS & STRUCTURES	-	7,219	7,205	-	-	14,424	1,731	650	16,805	63
	TOTAL COST	**158,095**	**14,012**	**45,570**	**-**	**-**	**217,647**	**20,302**	**7,193**	**296,991**	**1,120**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 12 - CMB Chemical Looping Gasification w/o CO2 Capture

Net Output Power, kW 265,146 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,415		677			2,092	251	94	2,437	9
1.2	Fuel Stackout and Reclaim	1,828		478			2,306	277	104	2,687	10
1.3	Fuel Conveyors	1,699		430			2,129	255	96	2,480	9
1.4	Other Fuel Handling	445		100			545	65	24	634	2
1.5	Sorbent Receive & Unload	95		30			125	15	6	146	1
1.6	Sorbent Stackout and Reclaim	1,534		295			1,829	219	82	2,130	8
1.7	Sorbent Conveyors	547	111	140			798	96	36	930	4
1.8	Other Sorbent Handling	330	73	182			585	70	26	681	3
1.9	Fuel & Sorbent Hnd. Foundations		1,788	2,129			3,917	470	176	4,563	17
	SUBTOTAL. 1	**7,893**	**1,972**	**4,461**	**-**	**-**	**14,326**	**1,718**	**644**	**16,688**	**63**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	782		160			942	113	42	1,097	4
2.2	Fuel Conveyor to Storage	2,500		572			3,072	369	138	3,579	13
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	626		137			763	91	34	888	3
2.6	Sorbent Storage & Feed	312		125			437	52	20	509	2
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		222	212			434	52	20	506	2
	SUBTOTAL. 2	**4,220**	**222**	**1,206**	**-**	**-**	**5,648**	**677**	**254**	**6,579**	**25**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	1,508		571			2,079	250	94	2,423	9
3.2	Water Makeup & Pretreating	683		210			893	107	40	1,040	4
3.3	Other Feeddwater Subsystems	862		379			1,241	149	58	1,448	5
3.4	Service Water System	131		70			201	24	9	234	1
3.5	Other Boiler Plant Systems	820		741			1,561	187	70	1,818	7
3.6	FO Supply System & Nat. Gas	37		46			83	10	4	97	0
3.7	Waste Treatment Equipment	501		253			754	90	34	878	3
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	493		188			681	82	31	794	3
	SUBTOTAL. 3	**5,035**	**-**	**2,458**	**-**	**-**	**7,493**	**899**	**340**	**8,732**	**33**
4	FLUIDIZED BED BOILER						-			-	-
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			51,257	193
4.2	Open						-			-	-
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	365		105			470	56	21	547	2
4.5	Primary Air System (Fans)	914		262			1,176	141	53	1,370	5
4.6	Secondary Air System (Fans)	52		15			67	8	3	78	0
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	**1,331**	**-**	**382**	**-**	**-**	**1,713**	**205**	**77**	**53,252**	**201**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 12 - CMB Chemical Looping Gasification w/o CO2 Capture

Net Output Power, kW 265,146 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories						-			-	-
5.4	Other Particulate Removal Materials						-			-	-
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.9	Gas Processing System (GPS)									-	-
	SUBTOTAL. 5	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator	38,400		1,824			40,224	5,533	1,627	47,384	179
6.2	Fuel Compressor & CT Accessories	54,332					54,332			54,332	205
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-		24	616	2
	SUBTOTAL. 6	**92,732**	**-**	**1,824**	**-**	**-**	**94,556**	**5,533**	**1,651**	**102,332**	**386**
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator	11,645		1,603			13,248	1,590	596	15,434	58
7.2	ID Fans	408		117			525	63	24	612	2
7.3	Ductwork	1,478		1,064			2,542	305	114	2,961	11
7.4	Stack	2,978		1,623			4,571	549	206	5,326	20
7.9	Duct & Stack Foundations		42	45			87	10	4	101	0
	SUBTOTAL. 7	**16,509**	**42**	**4,452**	**-**	**-**	**20,973**	**2,517**	**944**	**24,434**	**92**
8	STEAM TURBINE GENERATOR						-				-
8.1	Steam TG & Accessories	6,690		960			7,650	918	344	8,912	34
8.2	Turbine Plant Auxiliaries	-		-			-	-	-	-	-
8.3	Condenser & Auxiliaries	-					-	-	-	-	-
8.4	Steam Piping	3,588	134	1,945			5,667	680	255	6,602	25
8.9	TG Foundations		-	-			-	-		-	-
	SUBTOTAL. 8	**10,278**	**134**	**2,905**	**-**	**-**	**13,317**	**1,598**	**599**	**15,514**	**59**
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	1,169		591			1,760	211	79	2,050	8
9.2	Circulating Water Pumps	294		25			319	38	14	371	1
9.3	Circulating Water System Auxiliaries	79		10			89	11	4	104	0
9.4	Circulating Water Piping		581	569			1,150	138	52	1,340	5
9.5	Make-up Water System	67		87			154	18	7	179	1
9.6	Component Cooling Water System	58		45			103	12	5	120	0
9.9	Circ. Water System Foundations Structures		368	553			921	110	41	1,072	4
	SUBTOTAL. 9	**1,667**	**949**	**1,880**	**-**	**-**	**4,496**	**538**	**202**	**5,236**	**20**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	315		911			1,226	147	55	1,428	5
10.7	Ash Transport & Feed Equipment	3,952		1,961			5,913	710	266	6,889	26
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		149	166			315	38	14	367	1
	SUBTOTAL. 10	**4,267**	**149**	**3,038**	**-**	**-**	**7,454**	**895**	**335**	**8,684**	**33**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 4/24/2003

TOTAL PLANT COST SUMMARY

Case 12 - CMB Chemical Looping Gasification w/o CO2 Capture

Net Output Power, kW 265,146 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	930		126			1,056	127	48	1,231	5
115.2	Station Service Equipment	2,334		640			2,974	357	134	3,465	13
11.3	Switchgear & Motor Control	2,607		370			2,977	357	134	3,468	13
11.4	Conduit & Cable Tray		1,108	3,322			4,430	532	199	5,161	19
11.5	Wire & Cable		1,288	3,440			4,728	567	213	5,508	21
11.6	Protective Equipment	151		430			581	70	26	677	3
11.7	Standby Equipment	833		16			849	102	38	989	4
11.8	Main Power Transformer	1,334		53			1,387	166	62	1,615	6
11.9	Electrical Foundations		140	329			469	58	21	548	2
	SUBTOTAL. 11	**8,189**	**2,536**	**8,726**	**-**	**-**	**19,451**	**2,336**	**875**	**22,662**	**85**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	233		117			350	42	16	408	2
12.7	Distributed Control System Equipment	2,589		378			2,967	356	133	3,456	13
12.8	Instrument Wiring & tubing	1,114		2,951			4,065	488	183	4,736	18
12.9	Other I & C Equipment	665		1,260			1,925	231	87	2,243	8
	SUBTOTAL. 12	**4,601**	**-**	**4,706**	**-**	**-**	**9,307**	**1,117**	**419**	**10,843**	**41**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	388			411	49	19	479	2
13.2	Site Improvement		766	800			1,566	188	71	1,825	7
13.3	Site Facilities	1,373		1,139			2,512	301	113	2,926	11
	SUBTOTAL. 13	**1,373**	**789**	**2,327**	**-**	**-**	**4,489**	**538**	**203**	**5,230**	**20**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,162	1,917			4,079	490	184	4,753	18
14.2	Turbine Building		3,210	3,015			6,225	747	280	7,252	27
14.3	Administration Building		412	439			851	102	38	991	4
14.4	Circulation Water Pumphouse		88	71			159	19	7	185	1
14.5	Water Treatment Building		287	238			525	63	24	612	2
14.6	Machine Shop		367	249			616	74	28	718	3
14.7	Warehouse		249	252			501	60	23	584	2
14.8	Other Buildings & Structures		152	131			283	34	13	330	1
14.9	Waste Treating Building & Structure		292	893			1,185	142	53	1,380	5
	SUBTOTAL. 14	**-**	**7,219**	**7,205**	**-**	**-**	**14,424**	**1,731**	**650**	**16,805**	**63**

Table 9.3.39: Case-12 Overall Power Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
INITIAL & ANNUAL O&M EXPENSES
Cost Base: Jul-03

Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers
Case 12 - CMB Chemical Looping Gasification w/o CO2 Capture

Net Plant Heat Rate (Btu/kWh): 8,248
Net Power Output (kW): 265,146
Capacity Factor (%): 80

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	9.0	9.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	12.0	12.0

	Annual Cost $ / year	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	4,222,670	15.93
Maintenance Labor Costs (calc'd)	2,828,717	10.67
Administrative & Support Labor (calc'd)	1,762,847	6.65
TOTAL FIXED OPERATING COSTS	***8,814,235***	***33.24***
Maintenance Material Cost (calc'd)	3,394,461	0.0018

Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $ / year	Annual Unit Cost $/kW-net
Water (1000 gallons)		2,926	1.00		854,392	0.0005
Chemicals						
MU & WT Chem. (lbs.)	480,763	16,025	0.16	74,866	748,688	0.0004
Limestone (ton)	9,049	364.4	10.00	90,486	1,064,048	0.0006
Formic Acid (lbs.)			0.60			
Ammonia, NH3 (ton)			220			
Subtotal Chemicals				165,352	1,812,736	0.0010
Other						
Supplemental Fuel (MBtu)						
SCR Catalyst Replacement (MBtu)						
Emissions Penalties						
Subtotal Other						
Waste Disposal						
Fly Ash & Bottom Ash (ton)		925.4	8.00		2,161,734	0.0012
Subtotal Solid Waste Disposal					2,161,734	0.0012
By-Products & Emissions						
Gypsum (ton)						
Subtotal By-Products						
TOTAL VARIABLE OPERATING COST					***8,223,323***	***0.0044***

9.3.13. Case-13 Investment Costs and Operating and Maintenance Costs

Table 9.3.40: Case-13 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc. **Report Date:** 7/22/2003
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

TOTAL PLANT COST SUMMARY

Case 13 - CMB Chemical Looping Gasification w/ CO2 Capture

Net Output Power, kW 256,830 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					$	$/kW
1	FUEL & SORBENT HANDLING	8,523	2,129	4,816	-	-	15,468	1,858	695	18,021	70
2	FUEL & SORBENT PREP. & FEED	4,557	240	1,302	-	-	6,099	732	274	7,105	28
3	FEEDWATER & MISC. BOP SYSTEMS	5,841	-	2,852	-	-	8,693	1,043	391	10,127	39
4	CHEMICAL LOOPING GASIFIER										
4.1	Chemical Looping Gasifier, w/o Bhse & Accessories									62,708	244
4.2	Open										
4.4	Open										
4.4 - 4.9	Gasifier BoP (Fluidizing Air Fans)	578	-	166	-	-	744	87	33	864	3
	SUBTOTAL 4	578	-	166	-	-	744	87	33	63,572	248
5	FLUE GAS CLEANUP	-	-	-	-	-	-	-	-		
5.9	Gas Processing System (GPS)									64,880	253
	SUBTOTAL 5	-	-	-	-	-	-	-	-	64,880	253
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	57,934	-	8,775	0	0	66,709	8,005	3,002	77,716	303
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	16,478	42	4,452	-	0	20,972	2,517	944	24,433	95
	SUBTOTAL 7	16,478	42	4,452	-	-	20,972	2,517	944	24,433	95
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	7,761	-	1,113	0	0	8,874	1,065	399	10,338	40
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	4,162	155	2,256	0	0	6,573	788	295	7,656	30
	SUBTOTAL 8	11,923	155	3,369	-	-	15,447	1,853	694	17,994	70
9	COOLING WATER SYSTEM	1,667	949	1,880	-	-	4,496	538	202	5,236	20
10	ASH/SPENT SORBENT HANDLING SYSTEMS	4,608	161	3,281	-	-	8,050	968	362	9,380	37
11	ACCESSORY ELECTRIC PLANT	8,598	2,660	9,163	-	-	20,421	2,450	919	23,790	93
12	INSTRUMENTATION & CONTROL	4,601	4,706	-	-	-	9,307	1,117	419	10,843	42
13	IMPROVEMENT TO SITE	1,373	789	2,327	-	-	4,489	538	203	5,230	20
14	BUILDINGS & STRUCTURES	-	7,219	7,205	-	-	14,424	1,731	650	16,805	65
	TOTAL COST	**126,681**	**19,050**	**49,588**	**-**	**-**	**195,319**	**23,437**	**8,788**	**355,132**	**1,383**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 7/22/2003

TOTAL PLANT COST SUMMARY

Case 13 - CMB Chemical Looping Gasification w/ CO2 Capture

Net Output Power, kW 256,830 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					$	$/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,528		731			2,259	271	102	2,632	10
1.2	Fuel Stackout and Reclaim	1,974		516			2,490	299	112	2,901	11
1.3	Fuel Conveyors	1,835		464			2,299	276	103	2,678	10
1.4	Other Fuel Handling	480		108			588	71	26	685	3
1.5	Sorbent Receive & Unload	102		32			134	16	6	156	1
1.6	Sorbent Stackout and Reclaim	1,656		318			1,974	237	89	2,300	9
1.7	Sorbent Conveyors	591	120	152			863	104	39	1,006	4
1.8	Other Sorbent Handling	357	78	196			631	76	28	735	3
1.9	Fuel & Sorbent Hnd. Foundations		1,931	2,299			4,230	508	190	4,928	19
	SUBTOTAL. 1	**8,523**	**2,129**	**4,816**	**-**	**-**	**15,468**	**1,858**	**695**	**18,021**	**70**
2	FUEL & SORBENT PREP. & FEED						-			-	-
2.1	Coal Crushing & Drying	844		172			1,016	122	46	1,184	5
2.2	Fuel Conveyor to Storage	2,700		618			3,318	398	149	3,865	15
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	676		148			824	99	37	960	4
2.6	Sorbent Storage & Feed	337		135			472	57	21	550	2
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		240	229			469	56	21	546	2
	SUBTOTAL. 2	**4,557**	**240**	**1,302**	**-**	**-**	**6,099**	**732**	**274**	**7,105**	**28**
3	FEEDWATER & MISC. BOP SYSTEMS						-			-	-
3.1	Feedwater System	1,750		663			2,413	289	109	2,811	11
3.2	Water Makeup & Pretreating	793		244			1,037	124	47	1,208	5
3.3	Other Feeddwater Subsystems	1,000		440			1,440	173	65	1,678	7
3.4	Service Water System	152		81			233	28	10	271	1
3.5	Other Boiler Plant Systems	951		859			1,810	217	81	2,108	8
3.6	FO Supply System & Nat. Gas	42		54			96	12	4	112	0
3.7	Waste Treatment Equipment	581		293			874	105	39	1,018	4
3.8	Misc. Eqip. (Cranes, AirComp.Comm.)	572		218			790	95	36	921	4
	SUBTOTAL. 3	**5,841**	**-**	**2,852**	**-**	**-**	**8,693**	**1,043**	**391**	**10,127**	**39**
4	CHEMICAL LOOPING GASIFIER										
4.1	Chemical Looping Gasifier, w/o Bhse & Accessories						-			62,708	244
4.2	Open						-			-	-
4.3	Open						-			-	-
4.4	Gasifier BoP (Fluidizing Air Fans)						-			-	-
4.5	Primary Air System (Fans)						-			-	-
4.6	Secondary Air System (Fans)	578		166			744	87	33	864	3
4.7	Major Component Rigging						-				-
4.8	Gasifier Foundation						-			-	-
	SUBTOTAL. 4	**578**	**-**	**166**	**-**	**-**	**744**	**87**	**33**	**63,572**	**248**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 7/22/2003

TOTAL PLANT COST SUMMARY

Case 13 - CMB Chemical Looping Gasification w/ CO2 Capture

Net Output Power, kW 256,830 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories						-			-	-
5.4	Other Particulate Removal Materials						-			-	-
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.9	Gas Processing System (GPS)									64,880	253
	SUBTOTAL. 5	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**64,880**	**253**
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator	44,200		1,908			46,108	5,533	2,075	53,716	209
6.2	Fuel Compressor & CT Accessories	13,734		6,867			20,601	2,472	927	24,000	93
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	**57,934**	**-**	**8,775**	**-**	**-**	**66,709**	**8,005**	**3,002**	**77,716**	**303**
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator	11,645		1,603			13,248	1,590	596	15,434	60
7.2	ID Fans	408		117			525	63	24	612	2
7.3	Ductwork	1,478		1,064			2,542	305	114	2,961	12
7.4	Stack	2,947		1,623			4,570	549	206	5,325	21
7.9	Duct & Stack Foundations		42	45			87	10	4	101	0
	SUBTOTAL. 7	**16,478**	**42**	**4,452**	**-**	**-**	**20,972**	**2,517**	**944**	**24,433**	**95**
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	7,761		1,113			8,874	1,065	399	10,338	40
8.2	Turbine Plant Auxiliaries	54		108			162	19	7	188	1
8.3	Condenser & Auxiliaries	1,378		332			1,710	205	77	1,992	8
8.4	Steam Piping	2,730		1,566			4,296	515	193	5,004	19
8.9	TG Foundations		155	250			405	49	18	472	2
	SUBTOTAL. 8	**11,923**	**155**	**3,369**	**-**	**-**	**15,447**	**1,853**	**694**	**17,994**	**70**
9	COOLING WATER SYSTEM										
9.1	Cooling Towers	1,169		591			1,760	211	79	2,050	8
9.2	Circulating Water Pumps	294		25			319	38	14	371	1
9.3	Circulating Water System Auxiliaries	79		10			89	11	4	104	0
9.4	Circulating Water Piping		581	569			1,150	138	52	1,340	5
9.5	Make-up Water System	67		87			154	18	7	179	1
9.6	Component Cooling Water System	58		45			103	12	5	120	0
9.9	Circ. Water System Foundations Structures		368	553			921	110	41	1,072	4
	SUBTOTAL. 9	**1,667**	**949**	**1,880**	**-**	**-**	**4,496**	**538**	**202**	**5,236**	**20**
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	340		984			1,324	159	60	1,543	6
10.7	Ash Transport & Feed Equipment	4,268		2,118			6,386	768	287	7,441	29
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		161	179			340	41	15	396	2
	SUBTOTAL. 10	**4,608**	**161**	**3,281**	**-**	**-**	**8,050**	**968**	**362**	**9,380**	**37**

Client: ALSTOM Power Inc.
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers

Report Date: 7/22/2003

TOTAL PLANT COST SUMMARY

Case 13 - CMB Chemical Looping Gasification w/ CO2 Capture

Net Output Power, kW 256,830 **Estimate Type:** Conceptual **Cost Base:** Jul-03 ($x1000)

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost $	Total Cost $/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	977		132			1,109	133	50	1,292	5
11.2	Station Service Equipment	2,450		672			3,122	375	141	3,638	14
11.3	Switchgear & Motor Control	2,737		388			3,125	375	141	3,641	14
11.4	Conduit & Cable Tray		1,163	3,488			4,651	558	209	5,418	21
11.5	Wire & Cable		1,350	3,612			4,962	595	223	5,780	23
11.6	Protective Equipment	159		452			611	73	27	711	3
11.7	Standby Equipment	875		17			892	107	40	1,039	4
11.8	Main Power Transformer	1,400		56			1,456	175	66	1,697	7
11.9	Electrical Foundations		147	346			493	59	22	574	2
	SUBTOTAL. 11	**8,598**	**2,660**	**9,163**	**-**	**-**	**20,421**	**2,450**	**919**	**23,790**	**93**
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	233	117				350	42	16	408	2
12.7	Distributed Control System Equipment	2,589	378				2,967	356	133	3,456	13
12.8	Instrument Wiring & tubing	1,114	2,951				4,065	488	183	4,736	18
12.9	Other I & C Equipment	665	1,260				1,925	231	87	2,243	9
	SUBTOTAL. 12	**4,601**	**4,706**	**-**	**-**	**-**	**9,307**	**1,117**	**419**	**10,843**	**42**
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	388			411	49	19	479	2
13.2	Site Improvement		766	800			1,566	188	71	1,825	7
13.3	Site Facilities	1,373		1,139			2,512	301	113	2,926	11
	SUBTOTAL. 13	**1,373**	**789**	**2,327**	**-**	**-**	**4,489**	**538**	**203**	**5,230**	**20**
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,162	1,917			4,079	490	184	4,753	19
14.2	Turbine Building		3,210	3,015			6,225	747	280	7,252	28
14.3	Administration Building		412	439			851	102	38	991	4
14.4	Circulation Water Pumphouse		88	71			159	19	7	185	1
14.5	Water Treatment Building		287	238			525	63	24	612	2
14.6	Machine Shop		367	249			616	74	28	718	3
14.7	Warehouse		249	252			501	60	23	584	2
14.8	Other Buildings & Structures		152	131			283	34	13	330	1
14.9	Waste Treating Building & Structure		292	893			1,185	142	53	1,380	5
	SUBTOTAL. 14	**-**	**7,219**	**7,205**	**-**	**-**	**14,424**	**1,731**	**650**	**16,805**	**65**

Table 9.3.41: Case-13 Overall Power Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.
INITIAL & ANNUAL O&M EXPENSES
Cost Base: Jul-03

Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers
Case 13 - CMB Chemical Looping Gasification w/ CO2 Capture

Net Plant Heat Rate (Btu/kWh): 8,248
Net Power Output (kW): 256,830
Capacity Factor (%): 80

OPERATING & MAINTENANCE LABOR
Operating Labor

Operating Labor Rate (Base):	30.90	$/hour
Operating Labor Burden:	30.00	%
Labor O-H change Rate:	25.00	%

Operating Labor Requirements (O.J.) per shift	1 unit/mod.	Total Plant
Skilled Operator	1.0	1.0
Operator	11.0	11.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	14.0	14.0

	Annual Cost $ / year	Annual Unit Cost $/kW-net
Annual Operating Labor Costs (calc'd)	4,926,449	19.18
Maintenance Labor Costs (calc'd)	2,764,531	10.76
Administrative & Support Labor (calc'd)	1,922,745	7.49
TOTAL FIXED OPERATING COSTS	***9,613,725***	***37.43***
Maintenance Material Cost (calc'd)	3,317,437	0.0018

Consumables	Consumption Initial	Consumption Per Day	Unit Cost	Initial Cost	Annual Cost $ / year	Annual Unit Cost $/kW-net
Water (1000 gallons)		3,000	1.00		876,000	0.0005
Chemicals						
MU & WT Chem. (lbs.)	480,763	16,025	0.16	74,866	748,688	0.0004
Limestone (ton)	9,049	394.2	10.00	90,486	1,151,064	0.0006
Formic Acid (lbs.)			0.60			
Ammonia, NH3 (ton)			220			
Subtotal Chemicals				165,352	1,899,752	0.0011
Other						
Supplemental Fuel (MBtu)						
SCR Catalyst Replacement (MBtu)						
Emissions Penalties						
Subtotal Other						
Waste Disposal						
Fly Ash & Bottom Ash (ton)		1,001.1	8.00		2,338,570	0.0013
Subtotal Solid Waste Disposal					2,338,570	0.0013
By-Products & Emissions						
Gypsum (ton)						
Subtotal By-Products						
TOTAL VARIABLE OPERATING COST					***8,431,759***	***0.0047***

Table 9.3.42: Case-13 Gas Processing System Investment Costs

ABB LUMMUS GLOBAL HOUSTON

Project : CO2 Plant - DOE | Location : GC - USA | Project start: | Rev. : 00
Job/Prop # 0-9484 | **Plant : CO2 Case 13** | Mech.compl.:
Scope : EPC | Capacity :
Piece count: | Labor Prod.: | 1-Dec-02

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontract ($,000)	Total ($,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		9,502	147	6,018		6,165	9.5%
12000	Vessels / Filters		1,592	25	1,008		1,033	1.6%
12100	Towers / Internals		1,294	20	819		839	1.3%
12200	Reactors		-	-	-		-	0.0%
13000	Tanks		-	-	-		-	0.0%
14100	Pumps		443	7	281		288	0.4%
14200	Compressors		18,772	291	11,889		12,180	18.8%
18000	Special Equipment		4,295	67	2,720		2,787	4.3%
	Sub-Total Equipment	**31**	**35,898**	**556**	**22,736**	**-**	**23,292**	**35.9%**
21000	Civil		53,847	835	2,046		2,881	4.4%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		12,564	195	1,023		1,218	1.9%
23000	Buildings		14,359	223	546		768	1.2%
30000	Piping		98,720	1,530	4,547		6,077	9.4%
40000	Electrical		50,856	788	1,819		2,607	4.0%
50000	Instruments		41,881	649	3,183		3,832	5.9%
61100	Insulation		26,924	417	682		1,099	1.7%
61200	Fireproofing		17,949	278	341		619	1.0%
61300	Painting		14,958	232	193		425	0.7%
	Sub-Total Commodities		**332,059**	**5,147**	**14,380**	**-**	**19,527**	**30.1%**
70000	**Construction Indirects**						**8,279**	**12.8%**
	Sub-Total Direct Cost		**367,958**	**5,703**	**37,116**	**-**	**51,098**	**78.8%**
	ASU TIC plant cost						-	0.0%
71000	Constr. Management						824	1.3%
80000	Home Office Engineering						3,722	5.7%
80000	Basic Engineering						471	0.7%
95000	License fee	Excluded					-	0.0%
19400	Vendor Reps						942	1.5%
19300	Spare parts						1,530	2.4%
80000	Training cost	Excluded					-	0.0%
80000	Commissioning	Excluded					-	0.0%
19200	Catalyst & Chemicals						118	0.2%
97000	Freight						1,113	1.7%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**59,818**	**92.2%**
91400	Escalation						1,766	2.7%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**61,584**	**94.9%**
	Contracters Fee						3,296	5.1%
	Grand Total						**64,880**	**100.0%**

Exclusions : Bonds,Taxes,Import duties , Hazerdous material handling & disposal, Capital spare parts,
Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MCC.

Table 9.3.43: Case-13 Gas Processing System Operating and Maintenance Costs

Operating Costs ($/yr)	Variable Costs	Fixed Costs
Chemical and Dessicant	46675	
Waste Handling	0	
Fuel Gas *	133893	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	1981401	
Contracted services	1017687	
Column Total	3179656	306600
Grand Total (Fixed & Variable)	3486256	

* Based on $4/ MMBU and 7000 hours/ yr.
** Included in overall facility operating cost

9.4. Appendix IV: Economic Sensitivity Studies

Sensitivity analyses were conducted for all 13 case studies to determine the effect on COE of variation of selected base parameter values by ± 25 percent and CO_2 by-product selling price up to $20 per ton. These parameters (shaded in yellow in Table 4.1.1) are capacity factor, EPC price, coal price, CO_2 credit sell price, equity rate, corporate tax rate, and the discount rate for cost of capital. The base parameter values represent the point where all the sensitivity curves intersect (point 0, 0). Sensitivity analysis results tables and "spider plots" for all Cases (1–13) are provided in this appendix.

9.4.1. Case 1 – Air-Fired CFB without CO_2 Capture

Results for the Case 1 COE sensitivity study are shown in Figure 9.4.1 and summarized in Table 9.4.1. The levelized COE for the base parameter values is 4.5 cents per kWh. Levelized COE ranges from a low of 3.9 to a high of 5.5 cents per kWh.

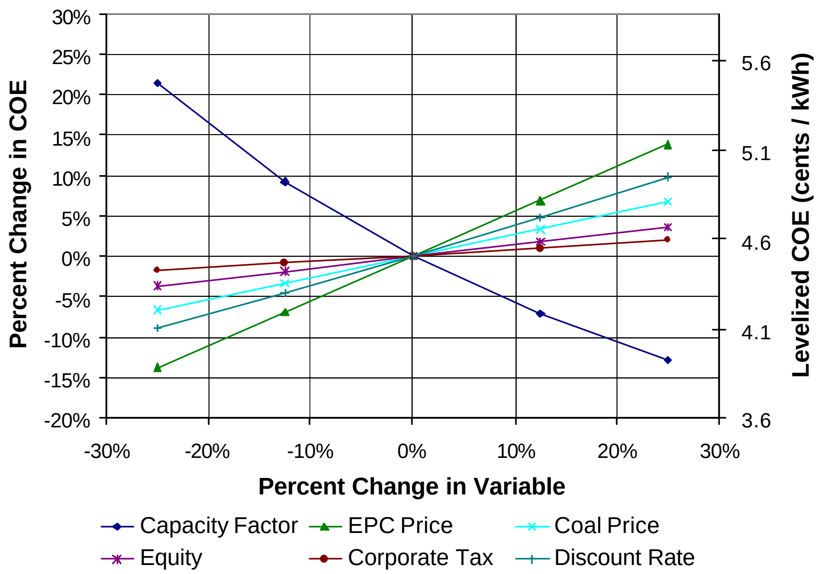


Figure 9.4.1: Case 1 - Air-Fired CFB without CO_2 Capture Economic Sensitivity Results

Table 9.4. 1: Case 1 – Air-Fired CFB without CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 1 - Air-Fired CFB without CO2 Capture												
Power Generation														
Net Output	kW	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	5,256	6,132	7,884	8,760	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5
Net Plant Heat Rate, HHV	Btu / kWh	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611
Net Generation	MWh / year	1,352,803	1,014,602	1,183,703	1,521,904	1,691,004	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803
Costs														
EPC Price	$ / kW	1,304	1,304	1,304	1,304	1,304	978	1,141	1,467	1,631	1,304	1,304	1,304	1,304
EPC Price	$1000s	**251,804**	251,804	251,804	251,804	251,804	188,853	220,329	283,280	314,755	251,804	251,804	251,804	251,804
Fixed O&M Costs	$1000 / year	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658
Fixed O&M Costs	$ / kW	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31
Variable O&M Costs	$1000 / year	5,587	4,190	4,889	6,286	6,984	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587
Variable O&M Costs	cents / kWh	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41
Total O&M Costs	cents / kWh	0.83	0.97	0.89	0.78	0.75	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427
CO2 Produced	lbm / kWh	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
CO2 Emitted	lbm / h	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427
CO2 Emitted	lbm / kWh	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		2.5	3.3	2.9	2.2	2.0	1.9	2.2	2.8	3.1	2.5	2.5	2.5	2.5
Fixed O&M		0.4	0.6	0.5	0.4	0.3	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	0.9	1.1	1.4	1.5
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		4.5	5.5	4.9	4.2	3.9	3.9	4.2	4.8	5.2	4.2	4.4	4.7	4.8

Parameter	Units	Case 1 - Air-Fired CFB without CO2 Capture											
Power Generation													
Net Output	kW	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5
Net Plant Heat Rate, HHV	Btu / kWh	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611
Net Generation	MWh / year	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803
Costs													
EPC Price	$ / kW	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304
EPC Price	$1000s	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804
Fixed O&M Costs	$1000 / year	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658
Fixed O&M Costs	$ / kW	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31
Variable O&M Costs	$1000 / year	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587
Variable O&M Costs	cents / kWh	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41
Total O&M Costs	cents / kWh	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83
Fuel Cost Calculation													
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions													
CO2 Produced	lbm / h	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427
CO2 Produced	lbm / kWh	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
CO2 Emitted	lbm / h	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427
CO2 Emitted	lbm / kWh	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
Financing Assumptions													
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13
Levelized Cost of Electricity (cents / kWh)													
Financial Component		2.3	2.4	2.6	2.7	2.4	2.5	2.5	2.6	2.1	2.3	2.7	2.9
Fixed O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		4.4	4.4	4.6	4.7	4.5	4.5	4.6	4.6	4.1	4.3	4.7	5.0

9.4.2. Case 2 – Oxygen-Fired CFB with ASU and CO_2 Capture

Results for the Case 2 COE sensitivity study are shown in Figure 9.4.2 and summarized in Table 9.4.2. The levelized COE for the base parameter values is 8.3 cents per kWh. Levelized COE ranges from a low of 5.6 to a high of 10.1 cents per kWh. CO_2 mitigation costs ranged from $12 to 62 per ton of CO_2 avoided (reference plant is Case 1) with the baseline value at $41 per ton of CO_2 avoided.

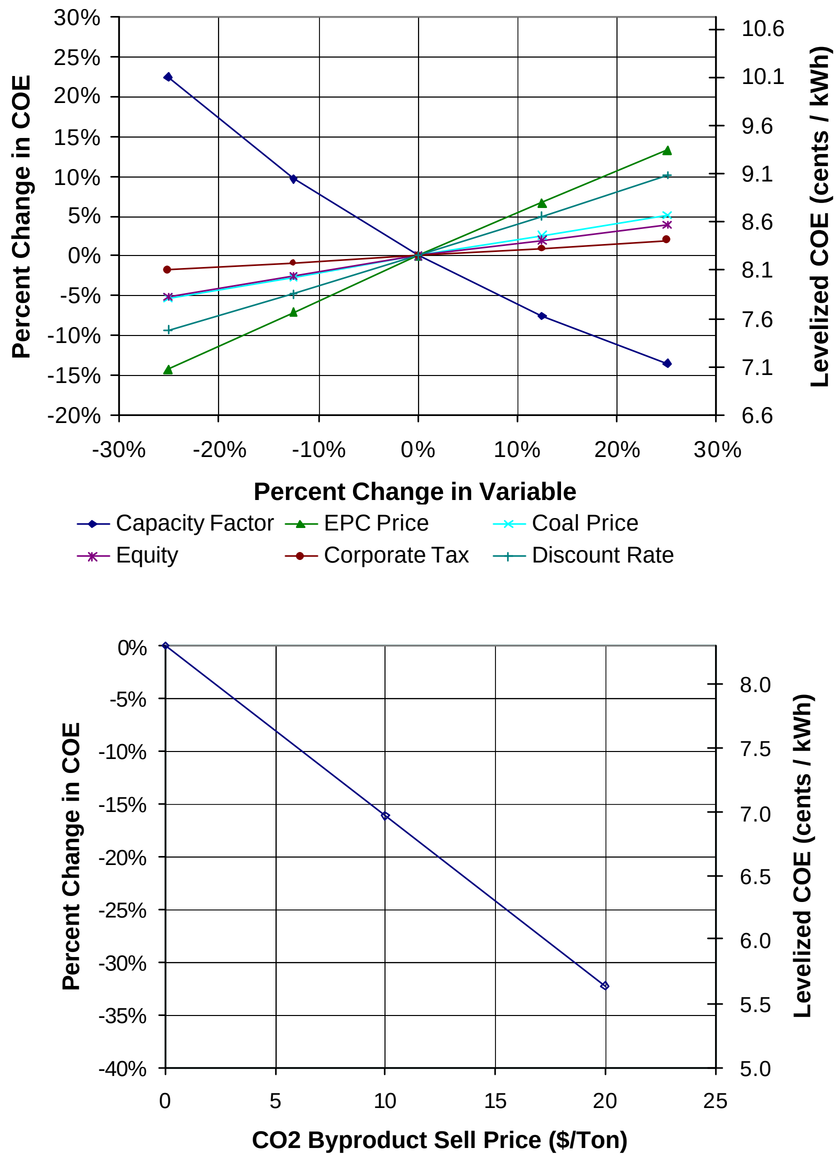


Figure 9.4.2: Case 2 - Oxygen-Fired CFB with ASU and CO_2 Capture Economic Sensitivity Results

Table 9.4. 2: Case 2 – Oxygen-Fired CFB with ASU and CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 2 - Oxygen-Fired CFB with ASU and CO2 Capture												
Power Generation														
Net Output	kW	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8
Net Plant Heat Rate, HHV	Btu / kWh	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760
Net Generation	MWh / year	928,020	696,015	812,018	1,044,023	1,160,025	928,020	928,020	928,020	928,020	928,020	928,020	928,020	928,020
Costs														
EPC Price	$ / kW	2,481	2,481	2,481	2,481	2,481	1,861	2,171	2,792	3,102	2,481	2,481	2,481	2,481
EPC Price	$1000s	**328,589**	328,589	328,589	328,589	328,589	246,442	287,515	369,663	410,736	328,589	328,589	328,589	328,589
Fixed O&M Costs	$1000 / year	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854
Fixed O&M Costs	$ / kW	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31
Variable O&M Costs	$1000 / year	8,820	6,615	7,718	9,923	11,025	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820
Variable O&M Costs	cents / kWh	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95
Total O&M Costs	cents / kWh	1.80	2.08	1.92	1.70	1.63	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Credits														
CO2 Price	$ / ton	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995
CO2 Produced	lbm / kWh	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85
CO2 Emitted	lbm / h	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618
CO2 Emitted	lbm / kWh	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		4.7	6.3	5.4	4.2	3.8	3.6	4.2	5.3	5.8	4.7	4.7	4.7	4.7
Fixed O&M		0.8	1.1	1.0	0.8	0.7	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Fuel		1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.3	1.5	1.9	2.2
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		8.3	10.1	9.0	7.6	7.1	7.1	7.7	8.8	9.3	7.8	8.0	8.5	8.7
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	41	62	50	34	29	28	35	47	53	36	39	43	46

Parameter	Units	Case 2 - Oxygen-Fired CFB with ASU and CO2 Capture													
Power Generation															
Net Output	kW	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423	132,423
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8	24.8
Net Plant Heat Rate, HHV	Btu / kWh	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760	13,760
Net Generation	MWh / year	928,020	928,020	928,020	928,020	928,020	928,020	928,020	928,020	928,020	928,020	928,020	928,020	928,020	928,020
Costs															
EPC Price	$ / kW	2,481	2,481	2,481	2,481	2,481	2,481	2,481	2,481	2,481	2,481	2,481	2,481	2,481	2,481
EPC Price	$1000s	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589
Fixed O&M Costs	$1000 / year	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854
Fixed O&M Costs	$ / kW	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31	59.31
Variable O&M Costs	$1000 / year	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820
Variable O&M Costs	cents / kWh	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95	0.95
Total O&M Costs	cents / kWh	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995
CO2 Produced	lbm / kWh	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85
CO2 Emitted	lbm / h	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618
CO2 Emitted	lbm / kWh	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		4.3	4.5	4.9	5.1	4.6	4.7	4.8	4.9	4.0	4.3	5.1	5.6	4.7	4.7
Fixed O&M		0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Fuel		1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.3	-2.7
Total		7.8	8.0	8.4	8.6	8.1	8.2	8.3	8.4	7.5	7.9	8.7	9.1	6.9	5.6
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	36	39	43	45	40	40	42	43	33	37	46	50	26	12

9.4.3. Case 3 – Oxygen-Fired CFB with ASU and Flue Gas Sequestration

Results for the Case 3 COE sensitivity study are shown in Figure 9.4.3 and summarized in Table 9.4.3. The levelized COE for the base parameter values is 8.0 cents per kWh. Levelized COE ranges from a low of 5.2 to a high of 9.8 cents per kWh. CO_2 mitigation costs ranged from $7 to 53 per ton of CO_2 avoided (reference plant is Case 1) with the baseline value at $35 per ton of CO_2 avoided.

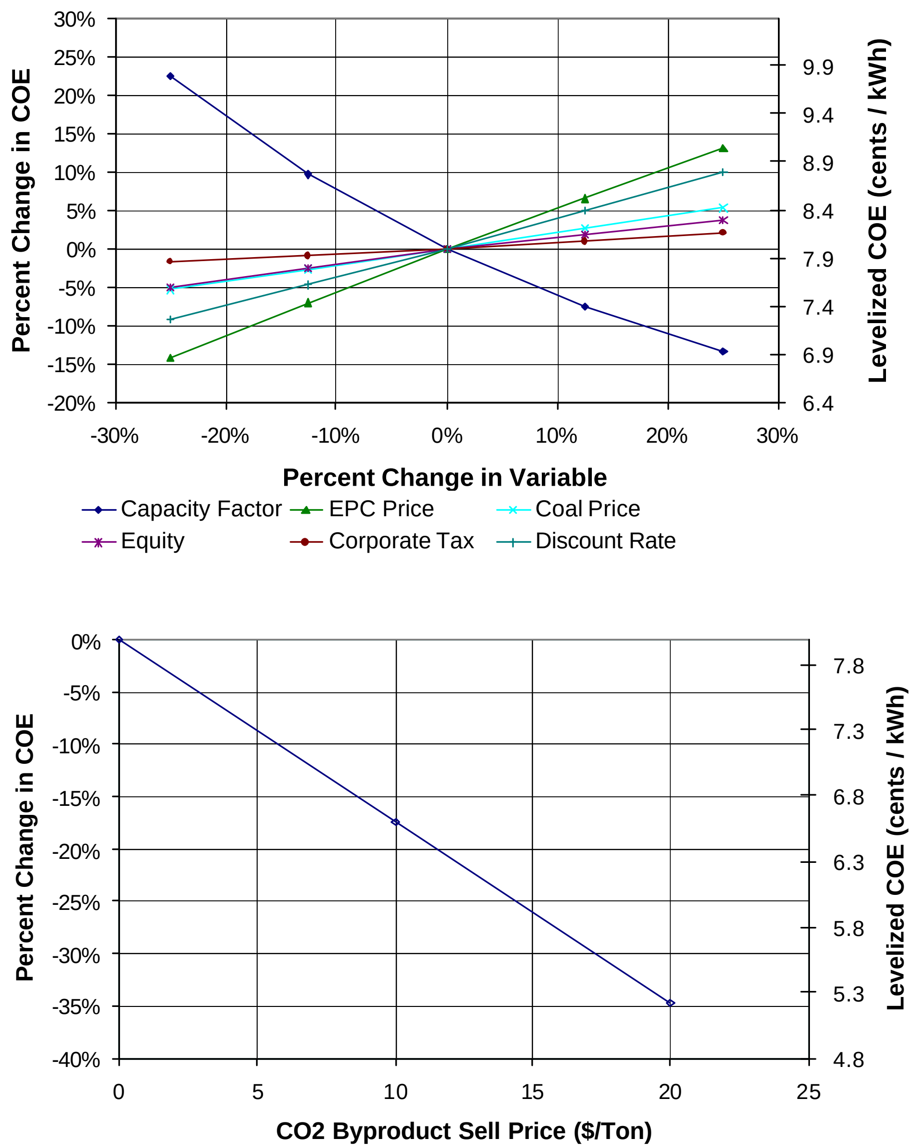


Figure 9.4.3: Case 3 - Oxygen-Fired CFB with ASU and Flue Gas Sequestration Economic Sensitivity Results

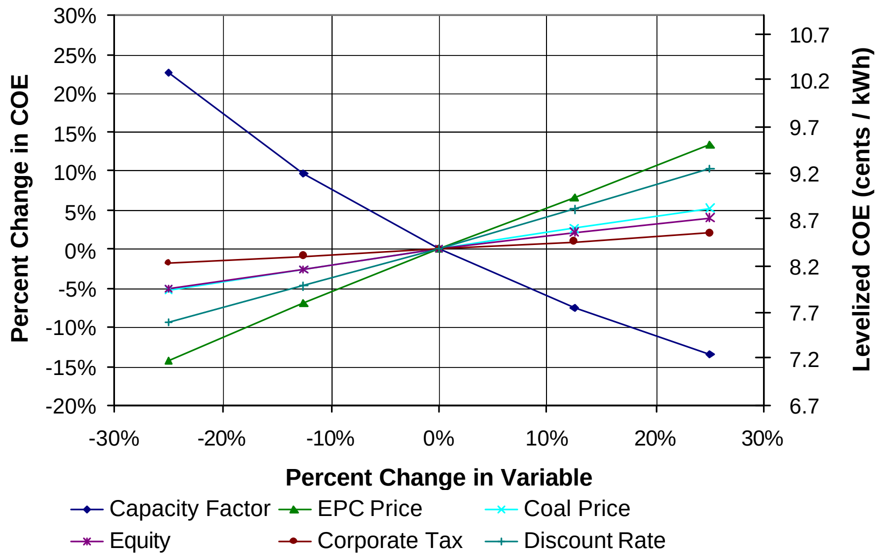
Table 9.4. 3: Case 3 – Oxygen-Fired CFB with ASU and Flue Gas Sequestration Sensitivity Analysis Results

Parameter	Units	Case 3 - Oxygen-Fired CFB with ASU and Flue Gas Sequestration												
Power Generation														
Net Output	kW	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3
Net Plant Heat Rate, HHV	Btu / kWh	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492
Net Generation	MWh / year	948,540	711,405	829,972	1,067,107	1,185,675	948,540	948,540	948,540	948,540	948,540	948,540	948,540	948,540
Costs														
EPC Price	$ / kW	2,369	2,369	2,369	2,369	2,369	1,777	2,073	2,665	2,961	2,369	2,369	2,369	2,369
EPC Price	$1000s	**320,638**	320,638	320,638	320,638	320,638	240,479	280,558	360,718	400,798	320,638	320,638	320,638	320,638
Fixed O&M Costs	$1000 / year	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061
Fixed O&M Costs	$ / kW	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55
Variable O&M Costs	$1000 / year	8,654	6,490	7,572	9,736	10,817	8,654	8,654	8,654	8,654	8,654	8,654	8,654	8,654
Variable O&M Costs	cents / kWh	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91
Total O&M Costs	cents / kWh	1.76	2.05	1.88	1.67	1.59	1.76	1.76	1.76	1.76	1.76	1.76	1.76	1.76
Credits														
CO2 Price	$ / ton	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466
CO2 Produced	lbm / kWh	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79
CO2 Emitted	lbm / h	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371
CO2 Emitted	lbm / kWh	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		4.5	6.0	5.2	4.0	3.6	3.4	4.0	5.0	5.6	4.5	4.5	4.5	4.5
Fixed O&M		0.8	1.1	1.0	0.8	0.7	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Fuel		1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.3	1.5	1.9	2.1
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		8.0	9.8	8.7	7.4	6.9	6.8	7.4	8.5	9.0	7.6	7.8	8.2	8.4
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	35	53	43	29	24	23	29	40	45	31	33	37	39

Parameter	Units	Case 3 - Oxygen-Fired CFB with ASU and Flue Gas Sequestration													
Power Generation															
Net Output	kW	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351	135,351
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3	25.3
Net Plant Heat Rate, HHV	Btu / kWh	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492	13,492
Net Generation	MWh / year	948,540	948,540	948,540	948,540	948,540	948,540	948,540	948,540	948,540	948,540	948,540	948,540	948,540	948,540
Costs															
EPC Price	$ / kW	2,369	2,369	2,369	2,369	2,369	2,369	2,369	2,369	2,369	2,369	2,369	2,369	2,369	2,369
EPC Price	$1000s	320,638	320,638	320,638	320,638	320,638	320,638	320,638	320,638	320,638	320,638	320,638	320,638	320,638	320,638
Fixed O&M Costs	$1000 / year	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061	8,061
Fixed O&M Costs	$ / kW	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55	59.55
Variable O&M Costs	$1000 / year	8,654	8,654	8,654	8,654	8,654	8,654	8,654	8,654	8,654	8,654	8,654	8,654	8,654	8,654
Variable O&M Costs	cents / kWh	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91
Total O&M Costs	cents / kWh	1.76	1.76	1.76	1.76	1.76	1.76	1.76	1.76	1.76	1.76	1.76	1.76	1.76	1.76
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466	377,466
CO2 Produced	lbm / kWh	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79
CO2 Emitted	lbm / h	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371	2,371
CO2 Emitted	lbm / kWh	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		4.1	4.3	4.7	4.8	4.4	4.5	4.6	4.7	3.8	4.2	4.9	5.3	4.5	4.5
Fixed O&M		0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Fuel		1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.4	-2.8
Total		7.6	7.8	8.1	8.3	7.8	7.9	8.1	8.1	7.2	7.6	8.4	8.8	6.6	5.2
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	31	33	36	38	33	34	36	36	27	31	39	43	21	7

9.4.4. Case 4 – Oxygen-Fired CMB with ASU and CO_2 Capture

Results for the Case 4 COE sensitivity study are shown in Figure 9.4.4 and summarized in Table 9.4.4. The levelized COE for the base parameter values is 8.4 cents per kWh. Levelized COE ranges from a low of 5.7 to a high of 10.3 cents per kWh. CO_2 mitigation costs ranged from $14 to 65 per ton of CO_2 avoided (reference plant is Case 1) with the baseline value at $43 per ton of CO_2 avoided.



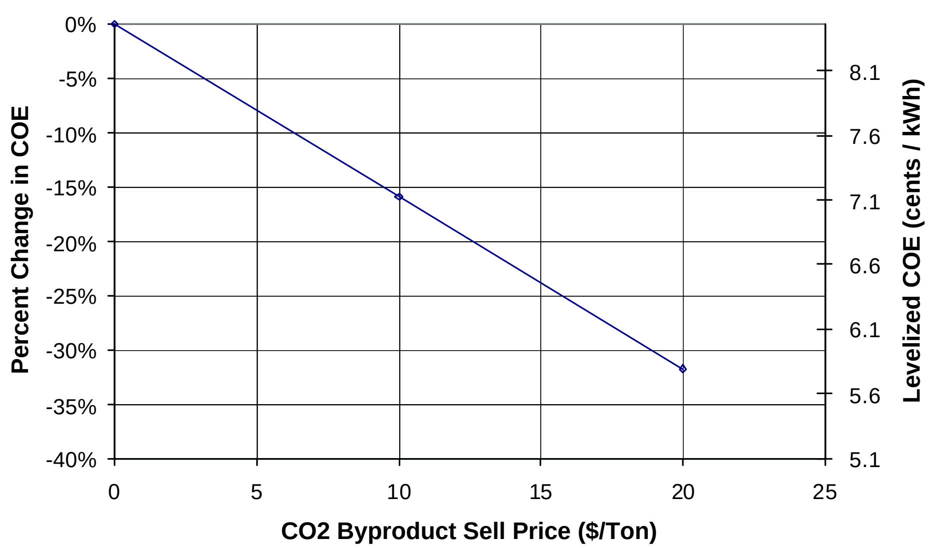


Figure 9.4.4: Case 4 - Oxygen-Fired CFB with ASU and CO_2 Capture Economic Sensitivity Results

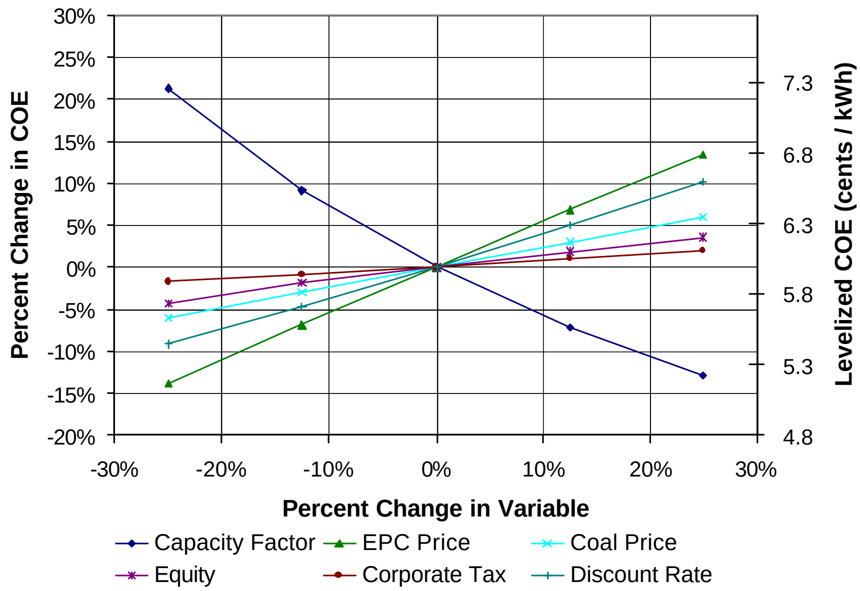
Table 9.4. 4: Case 4 – Oxygen-Fired CFB with ASU and CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 4 - Oxygen-Fired CMB with ASU and CO2 Capture												
Power Generation														
Net Output	kW	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6
Net Plant Heat Rate, HHV	Btu / kWh	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894
Net Generation	MWh / year	926,233	694,675	810,454	1,042,013	1,157,792	926,233	926,233	926,233	926,233	926,233	926,233	926,233	926,233
Costs														
EPC Price	$ / kW	2,553	2,553	2,553	2,553	2,553	1,915	2,234	2,872	3,191	2,553	2,553	2,553	2,553
EPC Price	$1000s	**337,402**	337,402	337,402	337,402	337,402	253,052	295,227	379,577	421,753	337,402	337,402	337,402	337,402
Fixed O&M Costs	$1000 / year	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899
Fixed O&M Costs	$ / kW	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77
Variable O&M Costs	$1000 / year	8,889	6,667	7,778	10,000	11,111	8,889	8,889	8,889	8,889	8,889	8,889	8,889	8,889
Variable O&M Costs	cents / kWh	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96
Total O&M Costs	cents / kWh	1.81	2.10	1.93	1.72	1.64	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81
Credits														
CO2 Price	$ / ton	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959
CO2 Produced	lbm / kWh	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87
CO2 Emitted	lbm / h	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579
CO2 Emitted	lbm / kWh	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		4.9	6.5	5.6	4.3	3.9	3.7	4.3	5.4	6.0	4.9	4.9	4.9	4.9
Fixed O&M		0.9	1.1	1.0	0.8	0.7	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Variable O&M		1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Fuel		1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.3	1.5	2.0	2.2
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		8.4	10.3	9.2	7.8	7.3	7.2	7.8	9.0	9.5	8.0	8.2	8.6	8.8
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	43	65	53	36	31	30	37	50	56	39	41	46	48

Parameter	Units	Case 4 - Oxygen-Fired CMB with ASU and CO2 Capture													
Power Generation															
Net Output	kW	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168	132,168
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6	24.6
Net Plant Heat Rate, HHV	Btu / kWh	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894	13,894
Net Generation	MWh / year	926,233	926,233	926,233	926,233	926,233	926,233	926,233	926,233	926,233	926,233	926,233	926,233	926,233	926,233
Costs															
EPC Price	$ / kW	2,553	2,553	2,553	2,553	2,553	2,553	2,553	2,553	2,553	2,553	2,553	2,553	2,553	2,553
EPC Price	$1000s	337,402	337,402	337,402	337,402	337,402	337,402	337,402	337,402	337,402	337,402	337,402	337,402	337,402	337,402
Fixed O&M Costs	$1000 / year	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899	7,899
Fixed O&M Costs	$ / kW	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77	59.77
Variable O&M Costs	$1000 / year	8,889	8,889	8,889	8,889	8,889	8,889	8,889	8,889	8,889	8,889	8,889	8,889	8,889	8,889
Variable O&M Costs	cents / kWh	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96	0.96
Total O&M Costs	cents / kWh	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959	379,959
CO2 Produced	lbm / kWh	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87	2.87
CO2 Emitted	lbm / h	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579	27,579
CO2 Emitted	lbm / kWh	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21	0.21
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		4.4	4.6	5.0	5.2	4.7	4.8	4.9	5.0	4.1	4.5	5.3	5.7	4.9	4.9
Fixed O&M		0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Variable O&M		1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Fuel		1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.3	-2.7
Total		8.0	8.2	8.6	8.8	8.3	8.3	8.5	8.6	7.6	8.0	8.8	9.3	7.1	5.7
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	39	41	45	47	42	43	44	45	35	39	48	53	29	14

9.4.5. Case 5 – Air-Fired CFB with Carbonate Regeneration and CO_2 Capture

Results for the Case 5 COE sensitivity study are shown in Figure 9.4.5 and summarized in Table 9.4.5. The levelized COE for the base parameter values is 5.9 cents per kWh. Levelized COE ranges from a low of 3.7 to a high of 7.2 cents per kWh. CO_2 mitigation costs ranged from $-8 to 27 per ton of CO_2 avoided (reference plant is Case 1) with the baseline value at $14 per ton of CO_2 avoided.



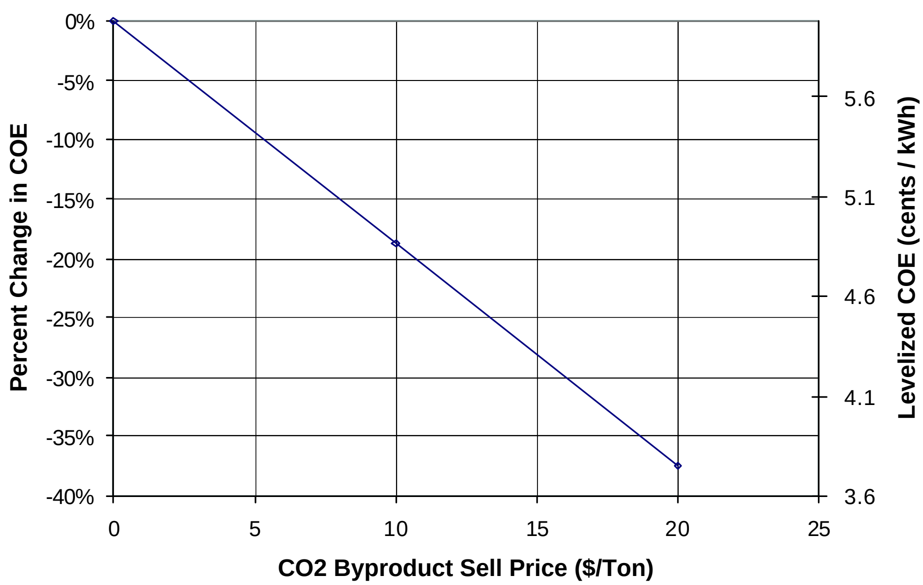


Figure 9.4.5: Case 5 - Air-Fired CFB with Carbonate Regeneration and CO_2 Capture Economic Sensitivity Results

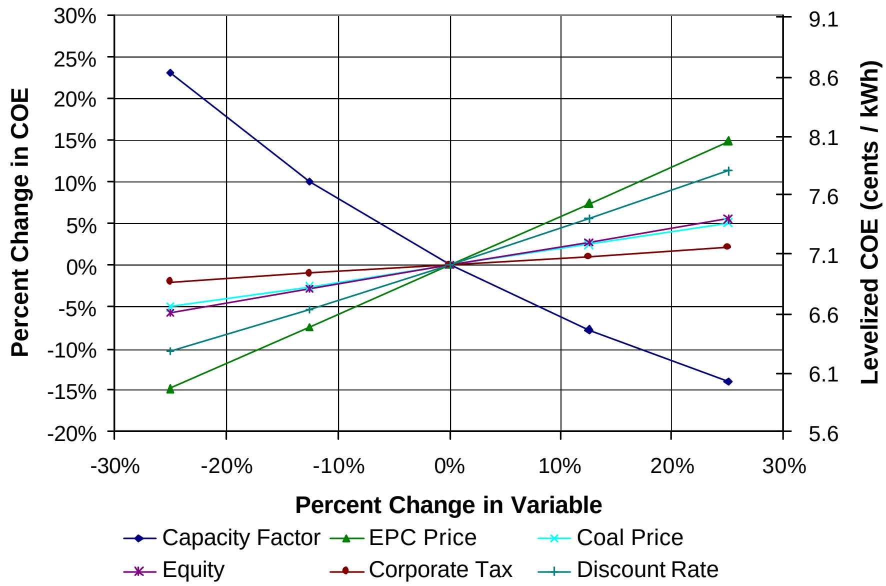
Table 9.4. 5: Case 5 – Air-Fired CFB with Carbonate Regeneration and CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 5 - Air-Fired CFB with Carbonate Regeneration and CO2 Capture												
Power Generation														
Net Output	kW	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2
Net Plant Heat Rate, HHV	Btu / kWh	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307
Net Generation	MWh / year	1,129,577	847,183	988,380	1,270,775	1,411,972	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577
Costs														
EPC Price	$ / kW	1,677	1,677	1,677	1,677	1,677	1,257	1,467	1,886	2,096	1,677	1,677	1,677	1,677
EPC Price	$1000s	**270,232**	270,232	270,232	270,232	270,232	202,674	236,453	304,011	337,790	270,232	270,232	270,232	270,232
Fixed O&M Costs	$1000 / year	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799
Fixed O&M Costs	$ / kW	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98
Variable O&M Costs	$1000 / year	8,264	6,198	7,231	9,298	10,331	8,264	8,264	8,264	8,264	8,264	8,264	8,264	8,264
Variable O&M Costs	cents / kWh	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73
Total O&M Costs	cents / kWh	1.25	1.42	1.32	1.19	1.14	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
Credits														
CO2 Price	$ / ton	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997
CO2 Produced	lbm / kWh	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23
CO2 Emitted	lbm / h	967	967	967	967	967	967	967	967	967	967	967	967	967
CO2 Emitted	lbm / kWh	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		3.3	4.4	3.8	2.9	2.6	2.5	2.9	3.7	4.1	3.3	3.3	3.3	3.3
Fixed O&M		0.5	0.7	0.6	0.5	0.4	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Variable O&M		0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Fuel		1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.1	1.2	1.6	1.8
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		5.9	7.2	6.5	5.5	5.2	5.1	5.5	6.4	6.7	5.6	5.8	6.1	6.3
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	14	27	20	10	7	6	10	18	22	11	12	16	18

Parameter	Units	Case 5 - Air-Fired CFB with Carbonate Regeneration and CO2 Capture													
Power Generation															
Net Output	kW	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184	161,184
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2	30.2
Net Plant Heat Rate, HHV	Btu / kWh	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307	11,307
Net Generation	MWh / year	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577	1,129,577
Costs															
EPC Price	$ / kW	1,677	1,677	1,677	1,677	1,677	1,677	1,677	1,677	1,677	1,677	1,677	1,677	1,677	1,677
EPC Price	$1000s	270,232	270,232	270,232	270,232	270,232	270,232	270,232	270,232	270,232	270,232	270,232	270,232	270,232	270,232
Fixed O&M Costs	$1000 / year	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799	5,799
Fixed O&M Costs	$ / kW	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98	35.98
Variable O&M Costs	$1000 / year	8,264	8,264	8,264	8,264	8,264	8,264	8,264	8,264	8,264	8,264	8,264	8,264	8,264	8,264
Variable O&M Costs	cents / kWh	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73
Total O&M Costs	cents / kWh	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997	359,997
CO2 Produced	lbm / kWh	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23	2.23
CO2 Emitted	lbm / h	967	967	967	967	967	967	967	967	967	967	967	967	967	967
CO2 Emitted	lbm / kWh	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		3.0	3.2	3.4	3.5	3.2	3.2	3.3	3.4	2.7	3.0	3.6	3.9	3.3	3.3
Fixed O&M		0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Variable O&M		0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Fuel		1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.1	-2.2
Total		5.7	5.8	6.1	6.2	5.8	5.9	6.0	6.1	5.4	5.7	6.2	6.6	4.8	3.7
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	12	13	15	16	13	14	15	15	9	11	17	20	3	-8

9.4.6. Case 6 – Oxygen-Fired CMB with OTM and CO_2 Capture

Results for the Case 6 COE sensitivity study are shown in Figure 9.4.6 and summarized in Table 9.4.6. The levelized COE for the base parameter values is 7.1 cents per kWh. Levelized COE ranges from a low of 4.8 to a high of 8.7 cents per kWh. CO_2 mitigation costs ranged from $3 to 45 per ton of CO_2 avoided (reference plant is Case 1) with the baseline value at $27 per ton of CO_2 avoided.



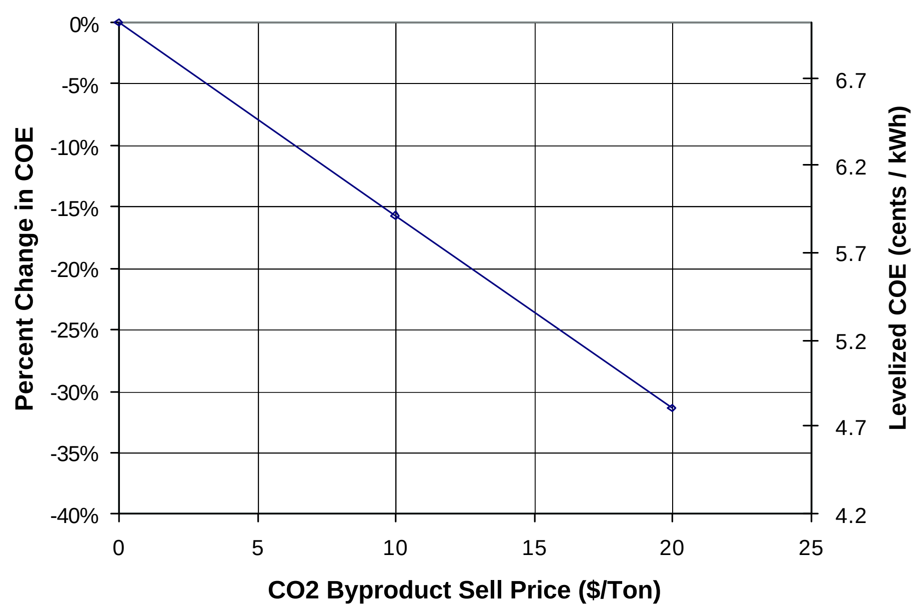


Figure 9.4.6: Case 6 - Oxygen-Fired CMB with OTM and CO_2 Capture Economic Sensitivity Results

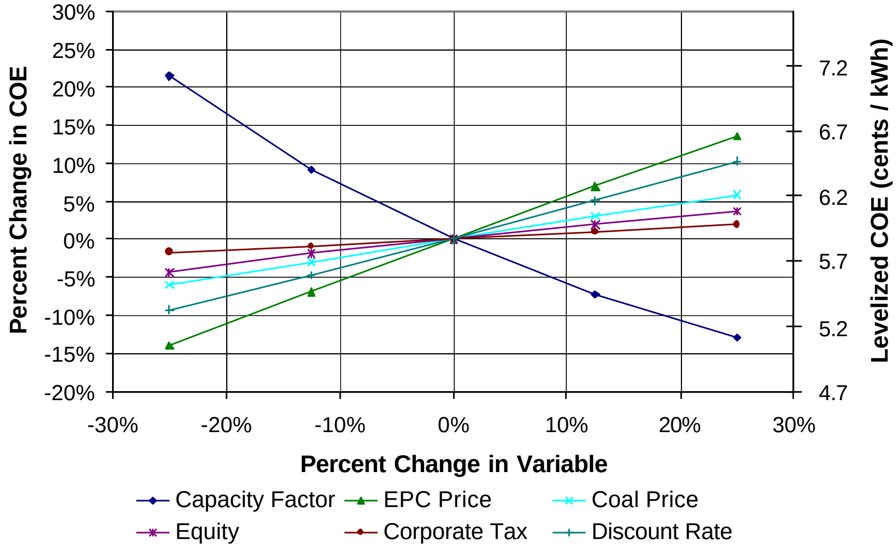
Table 9.4. 6: Case 6 – Oxygen-Fired CMB with OTM and CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 6 - Oxygen-Fired CMB with OTM and CO2 Capture												
Power Generation														
Net Output	kW	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0
Net Plant Heat Rate, HHV	Btu / kWh	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380
Net Generation	MWh / year	1,383,624	1,037,718	1,210,671	1,556,578	1,729,531	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624
Costs														
EPC Price	$ / kW	2,375	2,375	2,375	2,375	2,375	1,781	2,078	2,672	2,969	2,375	2,375	2,375	2,375
EPC Price	$1000s	**468,919**	468,919	468,919	468,919	468,919	351,689	410,304	527,534	586,149	468,919	468,919	468,919	468,919
Fixed O&M Costs	$1000 / year	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538
Fixed O&M Costs	$ / kW	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11
Variable O&M Costs	$1000 / year	10,134	7,600	8,867	11,400	12,667	10,134	10,134	10,134	10,134	10,134	10,134	10,134	10,134
Variable O&M Costs	cents / kWh	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73
Total O&M Costs	cents / kWh	1.20	1.36	1.27	1.15	1.11	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20
Credits														
CO2 Price	$ / ton	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301
CO2 Produced	lbm / kWh	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36
CO2 Emitted	lbm / h	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217
CO2 Emitted	lbm / kWh	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		4.4	5.9	5.1	3.9	3.5	3.4	3.9	5.0	5.5	4.4	4.4	4.4	4.4
Fixed O&M		0.5	0.6	0.5	0.4	0.4	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Variable O&M		0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Fuel		1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.1	1.2	1.6	1.8
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		7.1	8.7	7.8	6.5	6.1	6.0	6.5	7.6	8.1	6.7	6.9	7.2	7.4
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	27	45	35	21	17	16	22	33	39	23	25	29	31

Parameter	Units	Case 6 - Oxygen-Fired CMB with OTM and CO2 Capture													
Power Generation															
Net Output	kW	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435	197,435
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0	30.0
Net Plant Heat Rate, HHV	Btu / kWh	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380	11,380
Net Generation	MWh / year	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624	1,383,624
Costs															
EPC Price	$ / kW	2,375	2,375	2,375	2,375	2,375	2,375	2,375	2,375	2,375	2,375	2,375	2,375	2,375	2,375
EPC Price	$1000s	468,919	468,919	468,919	468,919	468,919	468,919	468,919	468,919	468,919	468,919	468,919	468,919	468,919	468,919
Fixed O&M Costs	$1000 / year	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538	6,538
Fixed O&M Costs	$ / kW	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11	33.11
Variable O&M Costs	$1000 / year	10,134	10,134	10,134	10,134	10,134	10,134	10,134	10,134	10,134	10,134	10,134	10,134	10,134	10,134
Variable O&M Costs	cents / kWh	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73	0.73
Total O&M Costs	cents / kWh	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301	466,301
CO2 Produced	lbm / kWh	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36	2.36
CO2 Emitted	lbm / h	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217	29,217
CO2 Emitted	lbm / kWh	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		4.0	4.2	4.6	4.8	4.3	4.4	4.5	4.6	3.7	4.1	4.8	5.2	4.4	4.4
Fixed O&M		0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Variable O&M		0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Fuel		1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.1	-2.2
Total		6.7	6.9	7.3	7.4	6.9	7.0	7.1	7.2	6.3	6.7	7.4	7.8	5.9	4.8
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	23	25	29	32	26	27	28	29	19	23	32	36	15	3

9.4.7. Case 7 – Chemical Looping Combustion with CO_2 Capture

Results for the Case 7 COE sensitivity study are shown in Figure 9.4.7 and summarized in Table 9.4.7. The levelized COE for the base parameter values is 5.8 cents per kWh. Levelized COE ranges from a low of 3.5 to a high of 7.1 cents per kWh. CO_2 mitigation costs ranged from $-10 to 26 per ton of CO_2 avoided (reference plant is Case 1) with the baseline value at $13 per ton of CO_2 avoided.



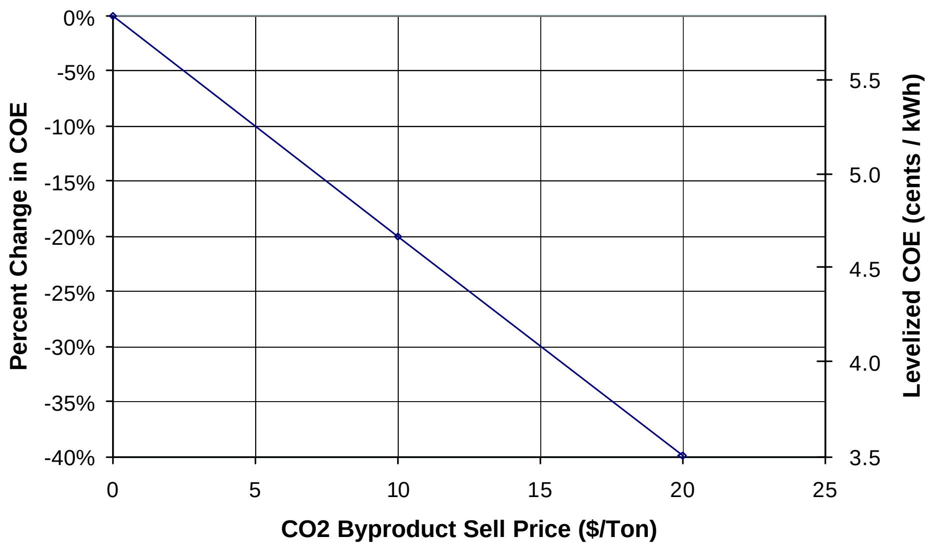


Figure 9.4.7: Case 7 – Chemical Looping Combustion with CO_2 Capture Economic Sensitivity Results

Table 9.4. 7: Case 7 – Chemical Looping Combustion with CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 7 - Chemical Looping Combustion with CO2 Capture												
Power Generation														
Net Output	kW	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9
Net Plant Heat Rate, HHV	Btu / kWh	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051
Net Generation	MWh / year	1,152,704	864,528	1,008,616	1,296,792	1,440,880	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704
Costs														
EPC Price	$ / kW	1,663	1,663	1,663	1,663	1,663	1,247	1,455	1,871	2,079	1,663	1,663	1,663	1,663
EPC Price	$1000s	**273,568**	273,568	273,568	273,568	273,568	205,176	239,372	307,764	341,960	273,568	273,568	273,568	273,568
Fixed O&M Costs	$1000 / year	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797
Fixed O&M Costs	$ / kW	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25
Variable O&M Costs	$1000 / year	8,015	6,011	7,013	9,017	10,018	8,015	8,015	8,015	8,015	8,015	8,015	8,015	8,015
Variable O&M Costs	cents / kWh	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70
Total O&M Costs	cents / kWh	1.20	1.37	1.27	1.14	1.10	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20
Credits														
CO2 Price	$ / ton	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453
CO2 Produced	lbm / kWh	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34
CO2 Emitted	lbm / h	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033
CO2 Emitted	lbm / kWh	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		3.3	4.4	3.7	2.9	2.6	2.4	2.9	3.7	4.1	3.3	3.3	3.3	3.3
Fixed O&M		0.5	0.7	0.6	0.4	0.4	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Variable O&M		0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Fuel		1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.0	1.2	1.6	1.7
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		5.8	7.1	6.4	5.4	5.1	5.0	5.4	6.3	6.6	5.5	5.7	6.0	6.2
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	13	26	19	9	6	5	9	17	21	10	11	15	17

Parameter	Units	Case 7 - Chemical Looping Combustion with CO2 Capture													
Power Generation															
Net Output	kW	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484	164,484
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9	30.9
Net Plant Heat Rate, HHV	Btu / kWh	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051	11,051
Net Generation	MWh / year	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704	1,152,704
Costs															
EPC Price	$ / kW	1,663	1,663	1,663	1,663	1,663	1,663	1,663	1,663	1,663	1,663	1,663	1,663	1,663	1,663
EPC Price	$1000s	273,568	273,568	273,568	273,568	273,568	273,568	273,568	273,568	273,568	273,568	273,568	273,568	273,568	273,568
Fixed O&M Costs	$1000 / year	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797	5,797
Fixed O&M Costs	$ / kW	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25	35.25
Variable O&M Costs	$1000 / year	8,015	8,015	8,015	8,015	8,015	8,015	8,015	8,015	8,015	8,015	8,015	8,015	8,015	8,015
Variable O&M Costs	cents / kWh	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70	0.70
Total O&M Costs	cents / kWh	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453	384,453
CO2 Produced	lbm / kWh	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34	2.34
CO2 Emitted	lbm / h	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033	1,033
CO2 Emitted	lbm / kWh	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		3.0	3.2	3.4	3.5	3.2	3.2	3.3	3.4	2.7	3.0	3.6	3.9	3.3	3.3
Fixed O&M		0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Variable O&M		0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Fuel		1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.2	-2.3
Total		5.6	5.7	6.0	6.1	5.7	5.8	5.9	6.0	5.3	5.6	6.1	6.4	4.7	3.5
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	11	12	14	15	12	13	14	14	8	10	16	19	2	-10

9.4.8. Case 8 – Texaco Built and Operating IGCC without CO_2 Capture

Results for the Case 8 COE sensitivity study are shown in Figure 9.4.8 and summarized in Table 9.4.8. The levelized COE for the base parameter values is 5.3 cents per kWh. Levelized COE ranges from a low of 4.6 a high of 6.6 cents per kWh.

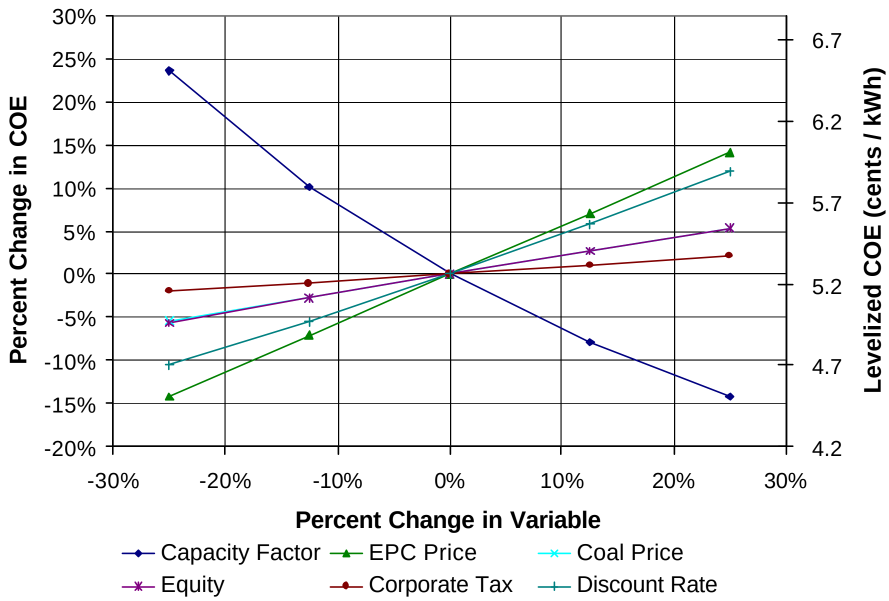


Figure 9.4.8: Case 8 – Texaco Built and Operating IGCC without CO_2 Capture Economic Sensitivity Results

Table 9.4. 8: Case 8 – Texaco Built and Operating IGCC without CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 8 - Texaco Built and Operating IGCC without CO2 Capture												
Power Generation														
Net Output	kW	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	5,256	6,132	7,884	8,760	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6
Net Plant Heat Rate, HHV	Btu / kWh	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069
Net Generation	MWh / year	1,843,714	1,382,785	1,613,249	2,074,178	2,304,642	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714
Costs														
EPC Price	$ / kW	1,565	1,565	1,565	1,565	1,565	1,174	1,369	1,761	1,956	1,565	1,565	1,565	1,565
EPC Price	$1000s	**411,731**	411,731	411,731	411,731	411,731	308,798	360,265	463,197	514,664	411,731	411,731	411,731	411,731
Fixed O&M Costs	$1000 / year	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180
Fixed O&M Costs	$ / kW	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70
Variable O&M Costs	$1000 / year	7,746	5,809	6,778	8,714	9,682	7,746	7,746	7,746	7,746	7,746	7,746	7,746	7,746
Variable O&M Costs	cents / kWh	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42
Total O&M Costs	cents / kWh	0.97	1.16	1.05	0.91	0.86	0.97	0.97	0.97	0.97	0.97	0.97	0.97	0.97
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093
CO2 Produced	lbm / kWh	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81
CO2 Emitted	lbm / h	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093
CO2 Emitted	lbm / kWh	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		3.2	4.3	3.7	2.8	2.6	2.4	2.8	3.6	3.9	3.2	3.2	3.2	3.2
Fixed O&M		0.6	0.7	0.6	0.5	0.4	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	0.9	1.0	1.3	1.4
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		5.3	6.6	5.8	4.9	4.6	4.6	4.9	5.7	6.1	5.0	5.2	5.4	5.6

Parameter	Units	Case 8 - Texaco Built and Operating IGCC without CO2 Capture											
Power Generation													
Net Output	kW	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087	263,087
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6	37.6
Net Plant Heat Rate, HHV	Btu / kWh	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069	9,069
Net Generation	MWh / year	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714	1,843,714
Costs													
EPC Price	$ / kW	1,565	1,565	1,565	1,565	1,565	1,565	1,565	1,565	1,565	1,565	1,565	1,565
EPC Price	$1000s	411,731	411,731	411,731	411,731	411,731	411,731	411,731	411,731	411,731	411,731	411,731	411,731
Fixed O&M Costs	$1000 / year	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180	10,180
Fixed O&M Costs	$ / kW	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70	38.70
Variable O&M Costs	$1000 / year	7,746	7,746	7,746	7,746	7,746	7,746	7,746	7,746	7,746	7,746	7,746	7,746
Variable O&M Costs	cents / kWh	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42
Total O&M Costs	cents / kWh	0.97	0.97	0.97	0.97	0.97	0.97	0.97	0.97	0.97	0.97	0.97	0.97
Fuel Cost Calculation													
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions													
CO2 Produced	lbm / h	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093
CO2 Produced	lbm / kWh	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81
CO2 Emitted	lbm / h	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093	477,093
CO2 Emitted	lbm / kWh	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81	1.81
Financing Assumptions													
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13
Levelized Cost of Electricity (cents / kWh)													
Financial Component		2.9	3.0	3.3	3.5	3.1	3.1	3.3	3.3	2.6	2.9	3.5	3.8
Fixed O&M		0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		5.0	5.2	5.4	5.6	5.2	5.2	5.4	5.4	4.7	5.0	5.6	5.9

9.4.9. Case 9 – Texaco Built and Operating IGCC with CO_2 Capture

Results for the Case 9 COE sensitivity study are shown in Figure 9.4.9 and summarized in Table 9.4.9. The levelized COE for the base parameter values is 7.2 cents per kWh. Levelized COE ranges from a low of 5.1 to a high of 8.9 cents per kWh. CO_2 mitigation costs ranged from $-3 to 45 per ton of CO_2 avoided (reference plant is Case 8) with the baseline value at $23 per ton of CO_2 avoided.

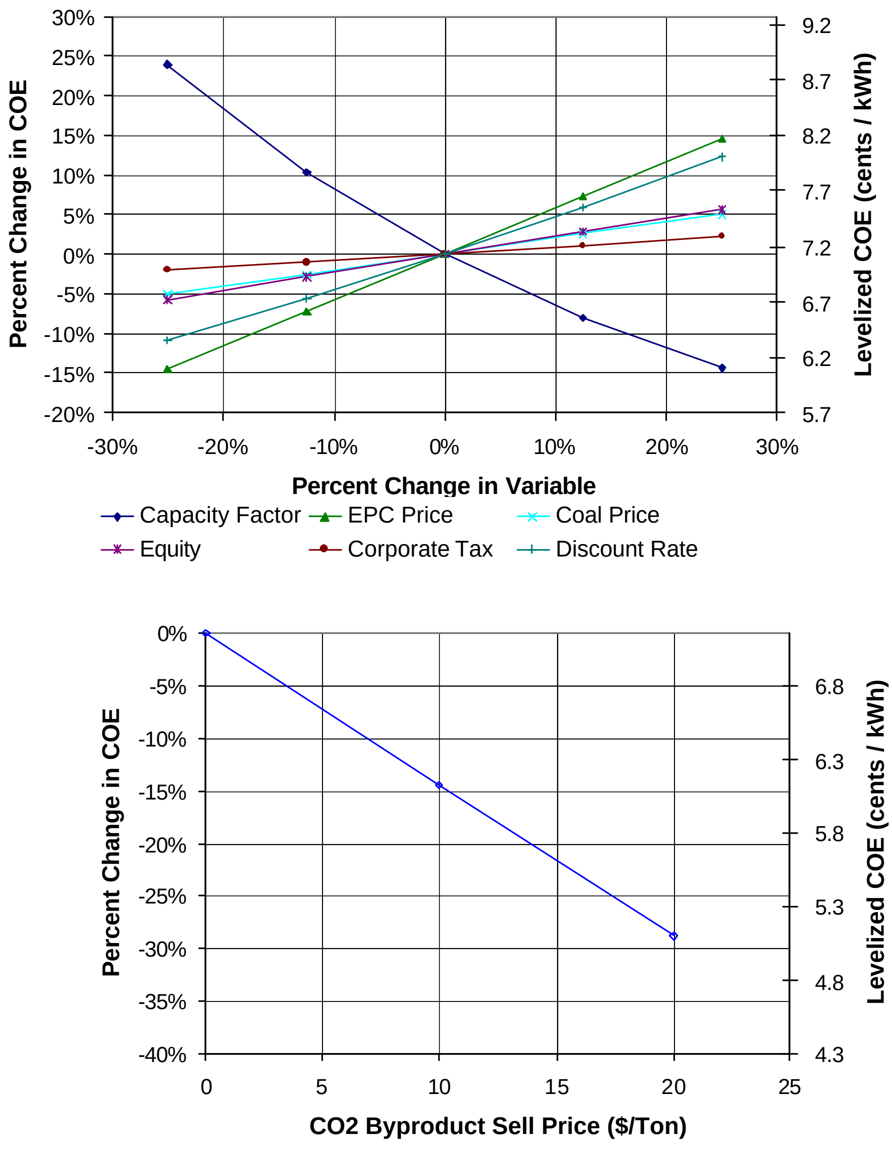


Figure 9.4.9: Case 9 – Texaco Built and Operating IGCC with CO_2 Capture Economic Sensitivity Results

Table 9.4. 9: Case 9 – Texaco Built and Operating IGCC with CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 9 - Texaco Built and Operating IGCC with CO2 Capture												
Power Generation														
Net Output	kW	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	5,256	6,132	7,884	8,760	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8
Net Plant Heat Rate, HHV	Btu / kWh	11,467	11,467	11,467	11,467	8,600	11,467	11,467	11,467	11,467	11,467	11,467	11,467	11,467
Net Generation	MWh / year	1,615,449	1,211,587	1,413,518	1,817,380	2,019,311	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449
Costs														
EPC Price	$ / kW	2,179	2,179	2,179	2,179	2,179	1,634	1,907	2,452	2,724	2,179	2,179	2,179	2,179
EPC Price	$1000s	**502,330**	502,330	502,330	502,330	502,330	376,748	439,539	565,121	627,913	502,330	502,330	502,330	502,330
Fixed O&M Costs	$1000 / year	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139
Fixed O&M Costs	$ / kW	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66
Variable O&M Costs	$1000 / year	9,202	6,901	8,052	10,352	11,502	9,202	9,202	9,202	9,202	9,202	9,202	9,202	9,202
Variable O&M Costs	cents / kWh	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57
Total O&M Costs	cents / kWh	1.32	1.57	1.43	1.24	1.17	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32
Credits														
CO2 Price	$ / ton	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791
CO2 Produced	lbm / kWh	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29
CO2 Emitted	lbm / h	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749
CO2 Emitted	lbm / kWh	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		4.4	5.9	5.0	3.9	3.5	3.4	3.9	4.9	5.4	4.4	4.4	4.4	4.4
Fixed O&M		0.8	1.0	0.9	0.7	0.6	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Fuel		1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.1	1.3	1.6	1.8
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		7.2	8.9	7.9	6.6	6.1	6.1	6.6	7.7	8.2	6.8	7.0	7.3	7.5
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	23	45	33	16	10	10	17	30	37	19	21	26	28

Parameter	Units	Case 9 - Texaco Built and Operating IGCC with CO2 Capture													
Power Generation															
Net Output	kW	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515	230,515
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8	29.8
Net Plant Heat Rate, HHV	Btu / kWh	11,467	11,467	11,467	11,467	11,467	11,467	11,467	11,467	11,467	11,467	11,467	13,760	13,760	13,760
Net Generation	MWh / year	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449	1,615,449
Costs															
EPC Price	$ / kW	2,179	2,179	2,179	2,179	2,179	2,179	2,179	2,179	2,179	2,179	2,179	2,179	2,179	2,179
EPC Price	$1000s	502,330	502,330	502,330	502,330	502,330	502,330	502,330	502,330	502,330	502,330	502,330	502,330	502,330	502,330
Fixed O&M Costs	$1000 / year	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139	12,139
Fixed O&M Costs	$ / kW	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66	52.66
Variable O&M Costs	$1000 / year	9,202	9,202	9,202	9,202	9,202	9,202	9,202	9,202	9,202	9,202	9,202	9,202	9,202	9,202
Variable O&M Costs	cents / kWh	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57
Total O&M Costs	cents / kWh	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791	528,791
CO2 Produced	lbm / kWh	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29	2.29
CO2 Emitted	lbm / h	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749	52,749
CO2 Emitted	lbm / kWh	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		4.0	4.2	4.6	4.8	4.3	4.3	4.5	4.6	3.6	4.0	4.8	5.3	4.4	4.4
Fixed O&M		0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Fuel		1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.0	-2.1
Total		6.7	7.0	7.4	7.6	7.0	7.1	7.2	7.3	6.4	6.8	7.6	8.0	6.1	5.1
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	18	21	26	28	22	22	24	25	14	18	29	34	10	-3

9.4.10. Case 10 – Texaco Commercially Offered IGCC with CO_2 Capture

Results for the Case 10 COE sensitivity study are shown in Figure 9.4.10 and summarized in Table 9.4.10. The levelized COE for the base parameter values is 5.2 cents per kWh. Levelized COE ranges from a low of 4.5 to a high of 6.4 cents per kWh.

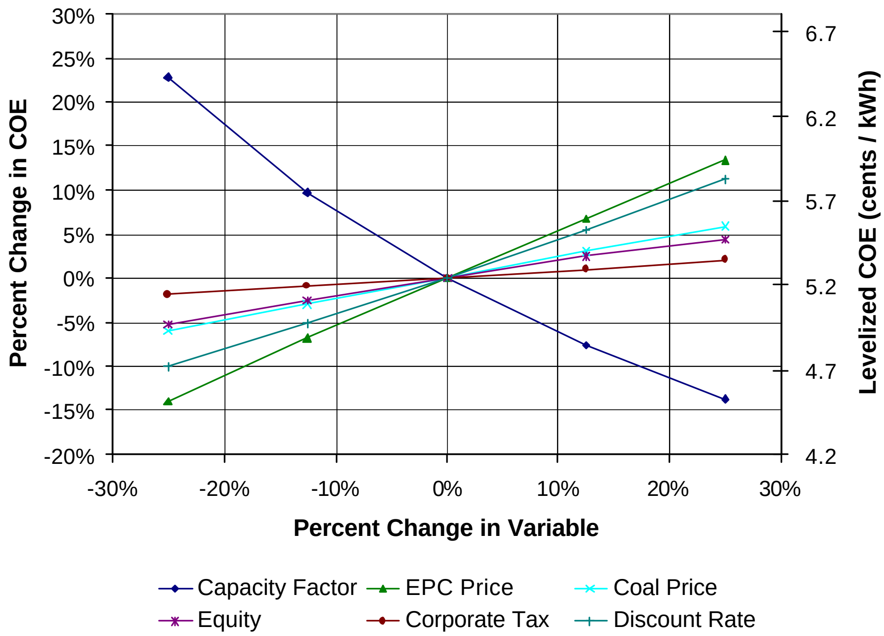


Figure 9.4.10: Case 10 - Texaco Commercially Offered IGCC without CO_2 Capture Economic Sensitivity Results

Table 9.4.10: Case 10 – Texaco Commercially Offered IGCC without CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 10 - Texaco Commercially Offered IGCC without CO2 Capture												
Power Generation														
Net Output	kW	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	5,256	6,132	7,884	8,760	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5
Net Plant Heat Rate, HHV	Btu / kWh	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884
Net Generation	MWh / year	1,648,940	1,236,705	1,442,823	1,855,058	2,061,175	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940
Costs														
EPC Price	$ / kW	1,451	1,451	1,451	1,451	1,451	1,088	1,270	1,633	1,814	1,451	1,451	1,451	1,451
EPC Price	$1000s	**341,468**	341,468	341,468	341,468	341,468	256,101	298,785	384,152	426,835	341,468	341,468	341,468	341,468
Fixed O&M Costs	$1000 / year	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344
Fixed O&M Costs	$ / kW	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71
Variable O&M Costs	$1000 / year	6,900	5,175	6,037	7,762	8,625	6,900	6,900	6,900	6,900	6,900	6,900	6,900	6,900
Variable O&M Costs	cents / kWh	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42
Total O&M Costs	cents / kWh	0.99	1.17	1.07	0.92	0.87	0.99	0.99	0.99	0.99	0.99	0.99	0.99	0.99
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940
CO2 Produced	lbm / kWh	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98
CO2 Emitted	lbm / h	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940
CO2 Emitted	lbm / kWh	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		3.0	4.0	3.4	2.7	2.4	2.3	2.6	3.3	3.7	3.0	3.0	3.0	3.0
Fixed O&M		0.6	0.8	0.6	0.5	0.5	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	0.9	1.1	1.4	1.5
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		5.2	6.4	5.7	4.8	4.5	4.5	4.9	5.6	5.9	4.9	5.1	5.4	5.5

Parameter	Units	Case 10 - Texaco Commercially Offered IGCC without CO2 Capture											
Power Generation													
Net Output	kW	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294	235,294
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5
Net Plant Heat Rate, HHV	Btu / kWh	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884	9,884
Net Generation	MWh / year	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940	1,648,940
Costs													
EPC Price	$ / kW	1,451	1,451	1,451	1,451	1,451	1,451	1,451	1,451	1,451	1,451	1,451	1,451
EPC Price	$1000s	341,468	341,468	341,468	341,468	341,468	341,468	341,468	341,468	341,468	341,468	341,468	341,468
Fixed O&M Costs	$1000 / year	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344	9,344
Fixed O&M Costs	$ / kW	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71	39.71
Variable O&M Costs	$1000 / year	6,900	6,900	6,900	6,900	6,900	6,900	6,900	6,900	6,900	6,900	6,900	6,900
Variable O&M Costs	cents / kWh	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42	0.42
Total O&M Costs	cents / kWh	0.99	0.99	0.99	0.99	0.99	0.99	0.99	0.99	0.99	0.99	0.99	0.99
Fuel Cost Calculation													
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions													
CO2 Produced	lbm / h	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940
CO2 Produced	lbm / kWh	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98
CO2 Emitted	lbm / h	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940
CO2 Emitted	lbm / kWh	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98
Financing Assumptions													
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13
Levelized Cost of Electricity (cents / kWh)													
Financial Component		2.7	2.9	3.1	3.2	2.9	3.0	3.1	3.1	2.5	2.7	3.3	3.6
Fixed O&M		0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		4.9	5.1	5.3	5.4	5.1	5.2	5.3	5.3	4.7	5.0	5.5	5.8

9.4.11. Case 11 – Texaco Commercially Offered IGCC with CO_2 Capture

Results for the Case 11 COE sensitivity study are shown in Figure 9.4.11 and summarized in Table 9.4.11. The levelized COE for the base parameter values is 7.2 cents per kWh. Levelized COE ranges from a low of 4.9 to a high of 8.8 cents per kWh. CO_2 mitigation costs ranged from $-3 to 42 per ton of CO_2 avoided (reference plant is Case 10) with the baseline value at $23 per ton of CO_2 avoided.

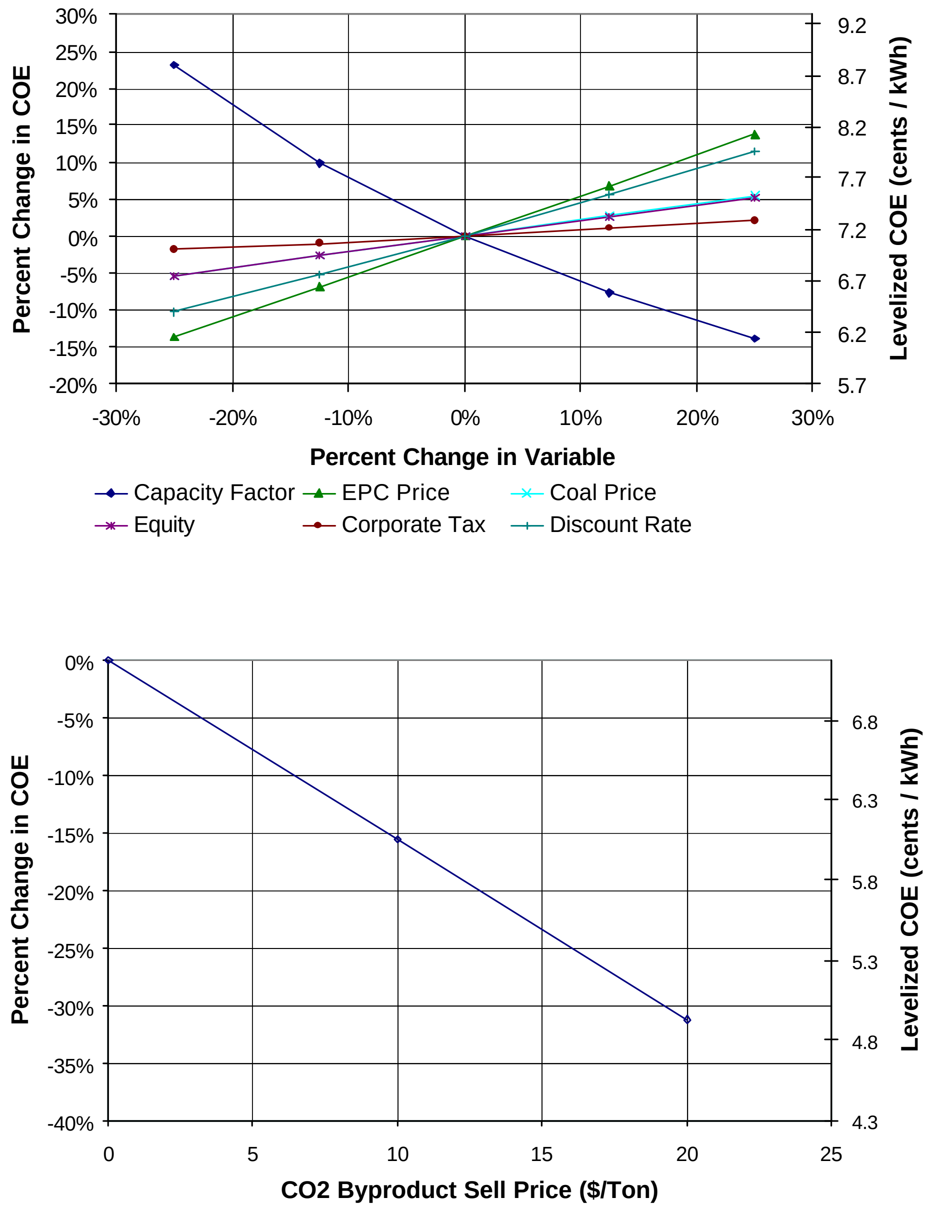


Figure 9.4.11: Case 11 - Texaco Commercially Offered IGCC with CO_2 Capture Economic Sensitivity Results

Table 9.4.11: Case 11 – Texaco Commercially Offered IGCC with CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 11 - Texaco Commercially Offered IGCC with CO2 Capture												
Power Generation														
Net Output	kW	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	5,256	6,132	7,884	8,760	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4
Net Plant Heat Rate, HHV	Btu / kWh	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441
Net Generation	MWh / year	1,408,636	1,056,477	1,232,557	1,584,716	1,760,795	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636
Costs														
EPC Price	$ / kW	2,052	2,052	2,052	2,052	2,052	1,539	1,795	2,308	2,564	2,052	2,052	2,052	2,052
EPC Price	$1000s	**412,377**	412,377	412,377	412,377	412,377	309,283	360,830	463,924	515,471	412,377	412,377	412,377	412,377
Fixed O&M Costs	$1000 / year	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068
Fixed O&M Costs	$ / kW	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06
Variable O&M Costs	$1000 / year	9,111	6,833	7,972	10,250	11,388	9,111	9,111	9,111	9,111	9,111	9,111	9,111	9,111
Variable O&M Costs	cents / kWh	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65
Total O&M Costs	cents / kWh	1.43	1.69	1.54	1.35	1.28	1.43	1.43	1.43	1.43	1.43	1.43	1.43	1.43
Credits														
CO2 Price	$ / ton	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275
CO2 Produced	lbm / kWh	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49
CO2 Emitted	lbm / h	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896
CO2 Emitted	lbm / kWh	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		4.2	5.6	4.8	3.7	3.4	3.2	3.7	4.7	5.2	4.2	4.2	4.2	4.2
Fixed O&M		0.8	1.0	0.9	0.7	0.6	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Fuel		1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.2	1.4	1.7	1.9
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		7.2	8.8	7.9	6.6	6.2	6.2	6.7	7.7	8.2	6.8	7.0	7.4	7.6
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	23	42	31	16	11	11	17	28	34	18	20	25	27

Parameter	Units	Case 11 - Texaco Commercially Offered IGCC with CO2 Capture													
Power Generation															
Net Output	kW	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004	201,004
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4	27.4
Net Plant Heat Rate, HHV	Btu / kWh	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441	12,441
Net Generation	MWh / year	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636	1,408,636
Costs															
EPC Price	$ / kW	2,052	2,052	2,052	2,052	2,052	2,052	2,052	2,052	2,052	2,052	2,052	2,052	2,052	2,052
EPC Price	$1000s	412,377	412,377	412,377	412,377	412,377	412,377	412,377	412,377	412,377	412,377	412,377	412,377	412,377	412,377
Fixed O&M Costs	$1000 / year	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068	11,068
Fixed O&M Costs	$ / kW	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06	55.06
Variable O&M Costs	$1000 / year	9,111	9,111	9,111	9,111	9,111	9,111	9,111	9,111	9,111	9,111	9,111	9,111	9,111	9,111
Variable O&M Costs	cents / kWh	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65	0.65
Total O&M Costs	cents / kWh	1.43	1.43	1.43	1.43	1.43	1.43	1.43	1.43	1.43	1.43	1.43	1.43	1.43	1.43
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275	500,275
CO2 Produced	lbm / kWh	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49	2.49
CO2 Emitted	lbm / h	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896	49,896
CO2 Emitted	lbm / kWh	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		3.8	4.0	4.4	4.6	4.1	4.1	4.3	4.3	3.5	3.8	4.6	5.0	4.2	4.2
Fixed O&M		0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Fuel		1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.1	-2.2
Total		6.8	7.0	7.4	7.5	7.0	7.1	7.3	7.3	6.4	6.8	7.6	8.0	6.1	4.9
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	18	20	25	27	21	22	23	24	14	18	27	32	10	-3

9.4.12. Case 12 – ALSTOM Chemical Looping Gasification without CO_2 Capture

Results for the Case 12 COE sensitivity study are shown in Figure 9.4.12 and summarized in Table 9.4.12. The levelized COE for the base parameter values is 4.3 cents per kWh. Levelized COE ranges from a low of 3.7 to a high of 5.2 cents per kWh.

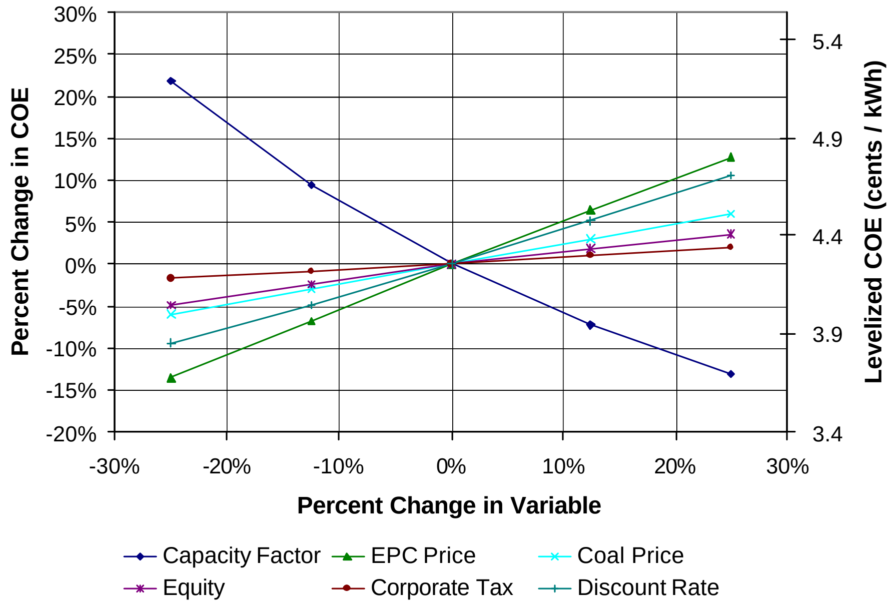


Figure 9.4.12: Case 12 – ALSTOM Chemical Looping Gasification without CO_2 Capture Economic Sensitivity Results

Table 9.4.12: Case 12 – ALSTOM Chemical Looping Gasification without CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 12 - ALSTOM Chemical Looping Gasification without CO2 Capture												
Power Generation														
Net Output	kW	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146
Availability Factor	%	**80**	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	5,256	6,132	7,884	8,760	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4
Net Plant Heat Rate, HHV	Btu / kWh	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248
Net Generation	MWh / year	1,858,143	1,393,607	1,625,875	2,090,411	2,322,679	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143
Costs														
EPC Price	$ / kW	1,120	1,120	1,120	1,120	840	980	1,260	1,400	1,120	1,120	1,120	1,120	1,120
EPC Price	$1000s	**296,991**	296,991	296,991	296,991	222,743	259,867	334,115	371,239	296,991	296,991	296,991	296,991	296,991
Fixed O&M Costs	$1000 / year	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814
Fixed O&M Costs	$ / kW	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24
Variable O&M Costs	$1000 / year	6,167	7,195	9,251	10,279	8,223	8,223	8,223	8,223	8,223	8,223	8,223	8,223	8,223
Variable O&M Costs	cents / kWh	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44
Total O&M Costs	cents / kWh	1.08	0.98	0.86	0.82	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92
Fuel Cost Calculation														
Coal Price	$ / MMBtu	**1.25**	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940
CO2 Produced	lbm / kWh	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75
CO2 Emitted	lbm / h	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940
CO2 Emitted	lbm / kWh	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98
Financing Assumptions														
Equity	%	**50**	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	**20**	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	**10**	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		2.3	3.1	2.7	2.1	1.9	1.8	2.0	2.6	2.9	2.3	2.3	2.3	2.3
Fixed O&M		0.5	0.6	0.5	0.4	0.4	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.8	0.9	1.2	1.3
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		4.3	5.2	4.7	4.0	3.7	3.7	4.0	4.6	4.8	4.0	4.2	4.4	4.5

Parameter	Units	Case 12 - ALSTOM Chemical Looping Gasification without CO2 Capture											
Power Generation													
Net Output	kW	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146	265,146
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4	41.4
Net Plant Heat Rate, HHV	Btu / kWh	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248	8,248
Net Generation	MWh / year	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143	1,858,143
Costs													
EPC Price	$ / kW	1,120	1,120	1,120	1,120	1,120	1,120	1,120	1,120	1,120	1,120	1,120	1,120
EPC Price	$1000s	296,991	296,991	296,991	296,991	296,991	296,991	296,991	296,991	296,991	296,991	296,991	296,991
Fixed O&M Costs	$1000 / year	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814	8,814
Fixed O&M Costs	$ / kW	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24	33.24
Variable O&M Costs	$1000 / year	8,223	8,223	8,223	8,223	8,223	8,223	8,223	8,223	8,223	8,223	8,223	8,223
Variable O&M Costs	cents / kWh	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44
Total O&M Costs	cents / kWh	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92
Fuel Cost Calculation													
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions													
CO2 Produced	lbm / h	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940
CO2 Produced	lbm / kWh	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75	1.75
CO2 Emitted	lbm / h	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940	464,940
CO2 Emitted	lbm / kWh	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98	1.98
Financing Assumptions													
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13
Levelized Cost of Electricity (cents / kWh)													
Financial Component		2.1	2.2	2.4	2.5	2.3	2.3	2.4	2.4	1.9	2.1	2.6	2.8
Fixed O&M		0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		4.1	4.2	4.4	4.4	4.2	4.2	4.3	4.4	3.9	4.1	4.5	4.7

9.4.13. Case 13 – ALSTOM Chemical Looping Gasification with CO_2 Capture

Results for the Case 13 COE sensitivity study are shown in Figure 9.4.13 and summarized in Table 9.4.13. The levelized COE for the base parameter values is 5.2 cents per kWh. Levelized COE ranges from a low of 3.3 to a high of 6.3 cents per kWh. CO_2 mitigation costs ranged from $-11 to 24 per ton of CO_2 avoided (reference plant is Case 12) with the baseline value at $11 per ton of CO_2 avoided.

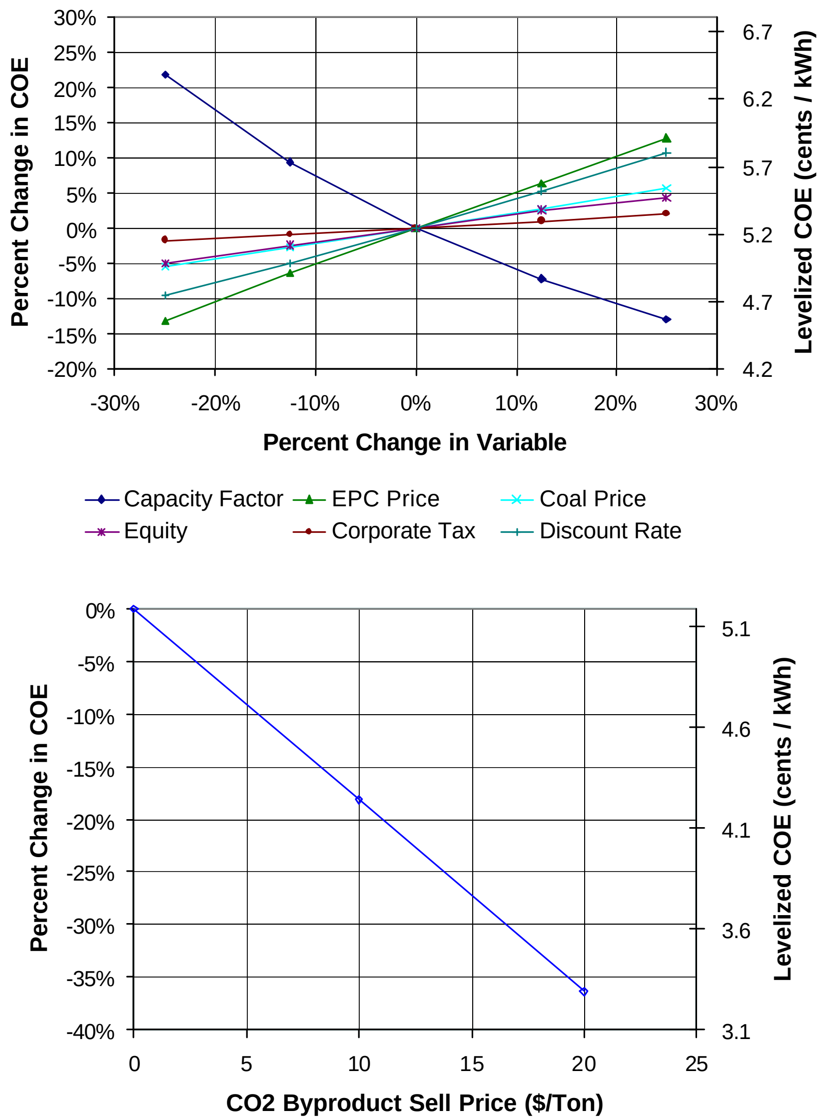


Figure 9.4.13: Case 13 – ALSTOM Chemical Looping Gasification with CO_2 Capture Economic Sensitivity Results

Table 9.4.13: Case 13 – ALSTOM Chemical Looping Gasification with CO_2 Capture Sensitivity Analysis Results

Parameter	Units	Case 13 - ALSTOM Chemical Looping Gasification with CO2 Capture												
Power Generation														
Net Output	kW	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830
Availability Factor	%	80	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	5,256	6,132	7,884	8,760	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9
Net Plant Heat Rate, HHV	Btu / kWh	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249
Net Generation	MWh / year	1,799,865	1,349,741	1,574,698	2,024,611	2,249,568	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654
Costs														
EPC Price	$ / kW	1,383	1,383	1,383	1,383	1,383	1,037	1,210	1,556	1,729	1,383	1,383	1,383	1,383
EPC Price	$1000s	355,132	355,132	355,132	355,132	355,132	266,349	310,741	399,524	443,915	355,132	355,132	355,132	355,132
Fixed O&M Costs	$1000 / year	7,916	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919
Fixed O&M Costs	$ / kW	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63
Variable O&M Costs	$1000 / year	9,888	8,858	10,335	13,287	14,764	11,811	11,811	11,811	11,811	11,811	11,811	11,811	11,811
Variable O&M Costs	cents / kWh	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66
Total O&M Costs	cents / kWh	0.99	1.39	1.29	1.15	1.10	1.21	1.21	1.21	1.21	1.21	1.21	1.21	1.21
Credits														
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation														
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600
CO2 Produced	lbm / kWh	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92
CO2 Emitted	lbm / h	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028
CO2 Emitted	lbm / kWh	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
Financing Assumptions														
Equity	%	50	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		2.9	3.8	3.3	2.5	2.9	2.2	2.5	3.2	3.5	2.9	2.9	2.9	2.9
Fixed O&M		0.6	0.7	0.6	0.5	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Variable O&M		0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Fuel		1.2	1.2	1.2	1.2	1.4	1.2	1.2	1.2	1.2	0.9	1.0	1.3	1.4
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		5.2	6.3	5.7	4.8	5.5	4.5	4.9	5.5	5.9	4.9	5.1	5.4	5.5
CO2 Mitigation														
CO2 Mitigation Cost	$ / ton	11	24	17	7	14	3	7	15	19	8	9	13	14

Parameter	Units	Case 13 - ALSTOM Chemical Looping Gasification with CO2 Capture													
Power Generation															
Net Output	kW	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830	256,830
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9	36.9
Net Plant Heat Rate, HHV	Btu / kWh	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249	9,249
Net Generation	MWh / year	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654	1,799,654
Costs															
EPC Price	$ / kW	1,383	1,383	1,383	1,383	1,383	1,383	1,383	1,383	1,383	1,383	1,383	1,383	1,383	1,383
EPC Price	$1000s	355,132	355,132	355,132	355,132	355,132	355,132	355,132	355,132	355,132	355,132	355,132	355,132	355,132	355,132
Fixed O&M Costs	$1000 / year	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919	9,919
Fixed O&M Costs	$ / kW	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63	38.63
Variable O&M Costs	$1000 / year	11,811	11,811	11,811	11,811	11,811	11,811	11,811	11,811	11,811	11,811	11,811	11,811	11,811	11,811
Variable O&M Costs	cents / kWh	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66	0.66
Total O&M Costs	cents / kWh	1.21	1.21	1.21	1.21	1.21	1.21	1.21	1.21	1.21	1.21	1.21	1.21	1.21	1.21
Credits															
CO2 Price	$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600	492,600
CO2 Produced	lbm / kWh	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92	1.92
CO2 Emitted	lbm / h	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028	6,028
CO2 Emitted	lbm / kWh	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		2.6	2.7	3.0	3.1	2.8	2.8	2.9	3.0	2.4	2.6	3.1	3.4	2.9	2.9
Fixed O&M		0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Variable O&M		0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Fuel		1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-0.9	-1.9
Total		5.0	5.1	5.3	5.4	5.1	5.2	5.3	5.3	4.7	5.0	5.5	5.8	4.3	3.3
CO2 Mitigation															
CO2 Mitigation Cost	$ / ton	8	9	13	14	10	10	12	12	5	8	14	18	0	-11