ENGINEERING FEASIBILITY AND ECONOMICS OF CO_2 CAPTURE ON AN EXISTING COAL-FIRED POWER PLANT

FINAL REPORT

VOLUME I
AMERICAN ELECTRIC POWER'S CONESVILLE POWER PLANT UNIT NO. 5 CO_2 CAPTURE RETROFIT STUDY

VOLUME II
BENCH-SCALE TESTING AND COMPUTATIONAL FLUID DYNAMICS (CFD) EVALUATION

SUBMITTED BY

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(CONTRACT NO. CDO/D-98-8)

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(CONTRACT NO. DE-FC26-99FT40576)

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PPL REPORT NO. PPL-01-CT-09

VOLUME I

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PUBLIC ABSTRACT

ALSTOM Power Inc.'s Power Plant Laboratories (ALSTOM) has teamed with American Electric Power (AEP), ABB Lummus Global Inc. (ABB), the US Department of Energy National Energy Technology Laboratory (DOE NETL), and the Ohio Coal Development Office (OCDO) to conduct a comprehensive study evaluating the technical feasibility and economics of alternate CO_2 capture and sequestration technologies applied to an existing US coal-fired electric generation power plant.

The motivation for this study was to provide input to potential US electric utility actions concerning GHG emissions reduction. If the US decides to reduce CO_2 emissions, action would need to be taken to address existing power plants. Although fuel switching from coal to natural gas may be one scenario, it will not necessarily be a sufficient measure and some form of CO_2 capture for use or disposal may also be required. The output of this CO_2 capture study will enhance the public's understanding of control options and influence decisions and actions by government, regulators, and power plant owners in considering the costs of reducing greenhouse gas CO_2 emissions.

The total work breakdown structure is encompassed within three major reports, namely: (1) Literature Survey, (2) AEP's Conesville Unit #5 Retrofit Study, and (3) Bench-Scale Testing and CFD Evaluation. The report on the literature survey results was issued earlier by Bozzuto, et al. (2000). Reports entitled "AEP's Conesville Unit #5 Retrofit Study" and "Bench-Scale Testing and CFD Evaluation" are provided as companion volumes, denoted Volumes I and II, respectively, of the final report. The work performed, results obtained, and conclusions and recommendations derived therefrom are summarized below.

Volume I

The Volume I report discusses three retrofit technology concepts, which were evaluated in conjunction with AEP's Conesville Unit #5, namely:

- **Concept A:** Coal combustion in air, followed by CO_2 separation with Kerr-McGee/ABB Lummus Global's commercial MEA-based absorption/stripping process
- **Concept B:** Coal combustion with O_2 firing and flue gas recycle (oxy-fuel firing)
- **Concept C:** Coal Combustion in air with oxygen removal and CO_2 separation by a mixture of primary and tertiary amines, i.e., monoethanolamine (MEA)/methyldiethanolamine (MDEA).

Each of these technologies was evaluated against a baseline case, the existing design without CO_2 capture, from the standpoints of performance, impacts on power generating cost, and CO_2 emissions. AEP's 450 MW Conesville Unit No. 5, located in Conesville, Ohio, was used for the power plant case study. All technical performance and cost results associated with the available options were evaluated in a comparative manner.

Major conclusions:

- No major technical barriers exist for retrofitting AEP's Conesville Unit #5 to capture CO_2 for any of the three concepts considered under this study.

- Concept B (oxygen firing with flue gas recycle) appears clearly to be the best alternative of the three concepts studied from both an efficiency and incremental COE viewpoint for systems designed for very high CO_2 capture (i.e. > 90%). If lower CO_2 capture fractions are considered, it appears that Concept A would likely be the best alternative for capture fractions below some as yet undetermined value. Concept C would also improve considerably with lower capture fractions.

- This study has confirmed two important issues related to firing coal in a CO_2-rich flue gas / O_2 mixture:

    - Modifications to the existing steam generator unit pressure arts are not required, and as such will also allow the unit to continue to operate in the conventional air- fired mode.

    - CO_2-rich flue gas can be cleaned and compressed with a relatively simple system to provide high purity CO_2 for usage or sequestration.

- Energy requirements and power consumption are high for all three concepts, resulting in significant decreases in overall power plant efficiencies (HHV basis), ranging from efficiencies of 20.5 to 22.5%, as compared to 35.0% for the Base case (air firing without CO_2 capture), which is equivalent to an energy penalty ranging from 36% to 41%. That is, the net power plant output was reduced to 77% - 59%.

- The MEA/MDEA mixture of Concept C requires about 28% less energy per pound of CO_2 captured to regenerate the solvent as compared to the MEA used in Concept A.

- Specific investment costs are high, ranging from about 800 to1800 $/kW and from 1000 to 2200 $/kW, corresponding to scenarios with and without replacement power, respectively.

- All cases studied indicate significant increases to the COE as a result of CO_2 capture. The incremental COE as compared to the Base case (air firing without CO_2 capture) ranges from 3.4 to 8.4 ¢/kWh. Similarly CO_2 mitigation costs range from about 42-98 $/ton of CO_2 avoided for the range of cases studied.

- Specific carbon dioxide emissions were reduced from about 2 lbm/kWh for the Base case to 0.13 – 0.27 lbm/kWh for the study cases. Recovery or capture of CO_2 ranged from 91 to 96%.

- Nominally, 5-8 acres of new equipment space is needed for Unit #5 alone on the existing 200-acre power plant site, which accommodates a total of 6 units with a total power generating capacity of 2,080 MW.

Major recommendations:

- A sensitivity study, for Concepts A and C, showing the impact of reducing CO_2 capture percentage on plant efficiency, investment cost, emissions, and cost of electricity. The current work utilized 90% CO_2 capture (nominal). The sensitivity study would investigate nominal CO_2 capture percentages of 70% and 50%.

- Detailed analysis of the existing steam turbine for Concept A: In Concept A about 79% of the steam leaving the intermediate-pressure (IP) turbine is extracted from the IP/LP crossover pipe for solvent regeneration. The capability of the existing low-pressure (LP) turbine to operate under these conditions of very low steam flow over the load range should be investigated in detail, preferably by the original equipment manufacturer (OEM).

- Optimization of the amine system reboiler steam pressure for Concepts A and C.

- Concept B detailed boiler system analysis: A startup/shutdown procedure and system design, particularly the transition from air to oxygen firing (including transient conditions) should be developed. Detailed metal temperature analysis for all heat exchanger sections, including operation at part loads should be analyzed. This should also include furnace wall metal temperatures and analysis of the circulation system. The existing fans should be checked (preferably by the OEM's) for operation under the new conditions. The feasibility of operating the boiler under a slight positive pressure to eliminate air infiltration should also be investigated.

- Investigation of Improved oxygen production systems for Concept B, in line with the membrane-based air separation research being conducted by various research groups (e.g., Praxair, Air Products, Norsk Hydro). Also optimization of an integrated boiler and oxygen production system.

- Measurement of furnace heat transfer in CO_2/O_2 environments in a proof of concept boiler simulation facility.

- Improved solvents, which require lower regeneration energy requirements and/or can be regenerated at a lower temperature level, similar to MHI's KS1 system, but for coal-firing application.

- Hybrid process using oxygen-enriched combustion and amine based CO_2 absorption, to accrue, simultaneously, both CO_2 capture and drastic NOx emissions reduction.

- Investigation of a new novel high risk CO_2 capture process that would reduce efficiency penalty and retrofit investment cost. This would likely not utilize the existing boiler.

Volume II

The bench-scale testing consisted primarily of drop-tube (DTFS-1) and thermo-gravimetric analyses (TGA). Three cases were investigated: (1) an air-fired case (Base case); (2) a case in which the N_2 was replaced by an equal mass of CO_2 (Constant Mass case); and (3) a case in which the N_2 volume was replaced by an equal volume of CO_2 (Constant Volume case). The Constant Mass case represents the appropriate retrofit scenario in a commercial pulverized coal firing application, albeit without flue gas recycling.

Major conclusions:

(a) Based on TGA results

- Both the Conesville Unit #5 and Pittsburgh Seam #8 coals have, for air (Base case) and O_2/CO_2 environments (Constant Mass/Volume cases), similar burning characteristics, as they both go to completion within the same temperature range (550-575 oC). Burning either of the coals in air or any of the O_2/CO_2 mixture ratios used here would produce virtually identical results; and (2) both coals have similar reactivity characteristics.

(b) Based on DTFS-1 results

- Residence time, as expected, has a significant impact on the overall coal combustion efficiency, and hence, on the unburned carbon emission in the fly ash for both coals under consideration. That is, while the combustion efficiency for both coals is only about 80% (dry-ash-free coal basis) at 0.2 sec., it is more than 98% at greater than 0.5 sec. Correspondingly, unburned carbon emissions range from about 65% to about 2% (dry fly ash basis).

- The impact of reaction medium is also significant. While the overall combustion efficiencies do not seem to exhibit significant differences between the baseline and Constant Mass firing, the more sensitive parameter, unburned carbon (UBC) in the fly ash, on the other hand, clearly shows better performance for the Constant Mass case. That is, the UBC at the furnace outlet was about a factor of 4.5 lower for the Constant Mass case (~2% vs. 9%). This is believed to be due to higher reactant gas temperature, and yet longer residence (0.77 sec. vs. 0.57 sec.). It appears, from the shape of the UBC profiles, that, if given enough residence time, these differences would narrow down considerably, implying that the ultimate performance in both mediums would be similar.

- The coal sample from Conesville Unit #5 appears to perform better than the sample from Pittsburgh seam #8 coal (e.g., 9% vs. 12.5% UBC for the air firing case). However, based on the shape of the UBC profiles, it can be seen that if given additional residence time, they would both be expected to perform similarly.

- NOx emission from oxy-fuel firing is about a factor of 3 (0.73 vs. 2.25 lb/MMBtu) lower than that of the Base Case. Correspondingly, sulfur dioxide is lower by about 19% (3.6 vs. 4.3 lb/MMBtu), and CO is significantly lower (0.09 lb/MMBtu vs. a negligible amount).

(c) Based on CFD results

- The baseline case exhibits higher carbon-in-ash (by 1.6 percentage points), higher outlet NOx emissions (by a factor of 2), and higher outlet CO (by a factor of 2) than the Concept-B case. These same computational trends are also qualitatively exhibited by the bench-scale testing.

- The baseline case exhibits a slightly higher peak gas temperature (maximum difference of about 200 °F), and a correspondingly higher average (cross-sectional) gas temperature (difference of 90 °F at the HFOT), than that of the Concept-B case.

- The net wall absorption in the furnace region for the baseline case is larger (by less than 1%) than that of the Concept-B case. However, A significant variation in both calculated and experimental irradiation to the wall between baseline and oxy-firing scenarios was also found in the literature (e.g., +6 to –18%). Differences may be partially attributed to the sensitivity of the irradiation and local emissivities to the aerodynamic and flame patterns, which are, in turn, a function of the furnace and firing system. It is suggested that radiatively absorbing gas species can either enhance or inhibit the irradiation in the vicinity of the wall, depending upon their local temperature and their relative spatial placement.

Major recommendations:

- Carry out a systematic coal fundamentals study concerning this field of endeavor. This study should be geared toward the creation of a database of information depicting the impact of coal nature (using coals of various rank coming from both domestic and international sources) when fired in an oxy-fuel environment on:
    - ➢ Coal devolatilization and char oxidation kinetics, and unburned carbon emissions in the fly ash
    - ➢ Acid rain-related gaseous (NOx, SO_2, and CO) emissions
    - ➢ Air toxics, particularly, mercury (Hg), volatile organic compounds (VOC's), and particulate emissions.
- Derive furnace heat transfer data from natural gas and coal firing in an oxy-fired pilot-scale furnace, which simulates the temperature/time history of a boiler. Use first the data from natural gas firing to validate the CFD code, without the complication of burning particles transiting the boiler. Apply the experience learned from modeling a natural gas fired boiler to model the coal-fired boiler.

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Background

International discussions continue to debate the need for reductions in emission levels of greenhouse gases (GHG). Because fossil fuel fired power plants are among the largest and most concentrated producers of CO_2 emissions, it stands to reason that recovery of CO_2 from the flue gas of such plants has been identified as one of the primary means for reducing CO_2 emissions.

ALSTOM Power Inc.'s Power Plant Laboratories (referred to herein as ALSTOM) teamed with American Electric Power (AEP), ABB Lummus Global Inc. (referred to herein as ABB), the US Department of Energy National Energy Technology Laboratory (DOE NETL), and the Ohio Coal Development Office (OCDO) to conduct a comprehensive study evaluating the technical and economic feasibility of three alternate CO_2 capture technologies applied to an existing US coal-fired electric generation power plant. The steam generator analyzed in this study is the Conesville No. 5 unit, operated by American Electric Power (AEP) of Columbus, Ohio. This steam generator is a nominal 450 MW, coal-fired, subcritical pressure, controlled circulation unit. It has a single cell furnace that employs corner-fired, tilting, tangential burners and which fires bituminous coal from the state of Ohio. The flue gas leaving the boiler system is cleaned of particulate matter in an electrostatic precipitator (ESP) and of SO_2 in a lime-based flue gas desulfurization (FGD) system before being discharged to the atmosphere.

ALSTOM is managing and performing the subject study from its Power Plant Laboratories office in Windsor, CT. ABB, from its offices in Houston, Texas, is participating as a sub-contractor. AEP is participating by offering their Conesville Generating Station as the case study and cost sharing consultation, along with relevant technical and cost data. AEP is one of the largest US utilities, and as the largest consumer of Ohio coal, is bringing considerable value to the project. Similarly, ALSTOM and ABB are well established as global leaders in the design and manufacturing of steam generating equipment, petrochemical, and CO_2 separation technology. The DOE National Energy Technology Laboratory and the Ohio Coal Development Office provided consultation and funding. All participants contributed to the cost share of this project.

Objectives

The motivation for this study was to provide input to potential US electric utility actions concerning GHG emissions reduction. If the US decides to reduce CO_2 emissions, action would need to be taken to address existing power plants. Although fuel switching from coal to natural gas may be one scenario, it will not necessarily be a sufficient measure and some form of CO_2 capture for use or disposal may also be required. The output of this CO_2 capture study will enhance the public's understanding of control options and influence decisions and actions by government, regulators, and power plant owners in considering the costs of reducing greenhouse gas CO_2 emissions.

The key goals of the study were to evaluate the impacts on the plant output, efficiency, and CO_2 emissions, resulting from the addition of the CO_2-capture systems. Cost estimates were developed for the systems required to produce, extract, clean and compress 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 cost of electricity (COE) and on the mitigation cost for CO_2 (\$/ton of CO_2 avoided) was also evaluated.

Work Scope

The total work breakdown structure is encompassed within three major reports, namely: (1) Literature Survey, (2) AEP's Conesville Unit #5 Retrofit Study, and (3) Bench-Scale Testing and CFD Evaluation. The report on the literature survey results was issued earlier by Bozzuto, et al. (2000). The report entitled "Bench-Scale Testing and CFD Evaluation" constitutes Volume II of the final report.

Work on AEP's Conesville Unit #5 Retrofit Study is presented here as Volume I report. The work performed, results obtained, and conclusions and recommendations derived therefrom are summarized below.

System Descriptions

The three retrofit CO_2 capture technology concepts evaluated in this study are listed below:

- **Concept A**: Coal combustion in air, followed by CO_2 separation with Kerr-McGee/ABB Lummus Global's commercial MEA-based absorption/stripping process
- **Concept B**: Coal combustion with O_2 firing and flue gas recycle (oxy-fuel firing)
- **Concept C**: Coal combustion in air with oxygen removal and CO_2 separation by a mixture of primary and tertiary amines

The results from each of these three evaluations were compared to the Base Case, which represents the "business as usual" operating scenario for the existing plant without CO_2 capture.

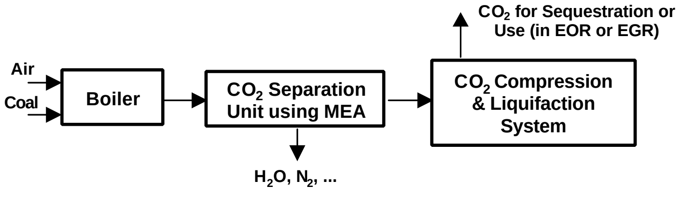
The CO_2 capture systems were designed for a minimum of 90% CO_2 capture. The Dakota Gasification Company's CO_2 specification (DGC WebPages, 2001) for EOR, given in the following table, was used as the basis for the CO_2 capture system design. A very low concentration of oxygen, in particular, is specified for meeting current pipeline operating practices, due to the corrosive nature of the oxygen. Hence, for Concept B, whereby the final CO_2 liquid product was found to contain about 9300 vppm of O_2, the design of the transport pipe to an EOR site for example would have to take this characteristic under consideration.

Dakota Gasification Project's CO_2 Specification for EOR

Component	Unit	Concentration
CO_2	Vol.%	96.0
H_2S	Vol.%	0.9
CH4	Vol.%	0.7
C2+HCs	Vol.%	2.3
CO	Vol.%	0.1
N_2	vppm	< 300
H_2O	vppm	< 20
O_2	vppm	< 50

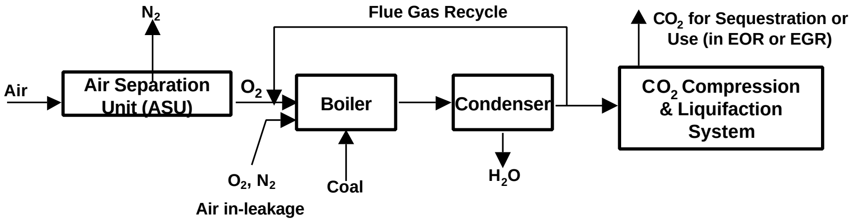
Concepts A and C are both low temperature CO_2 absorption systems located downstream of the FGD system and as such do not impact the operation of the boiler.

In **Concept A** Coal is burned conventionally in air as schematically depicted below. The flue gases leaving the modified FGD system (a secondary absorber is added to reduce the SO_2 concentration for MEA system) are cooled with a direct contact cooler and ducted to the MEA system where more than 96% of the CO_2 is removed, compressed, and liquefied for usage or sequestration. The remaining flue gases leaving the new MEA system, consisting of primarily oxygen, nitrogen, water vapor and a relatively small amount of sulfur dioxide and carbon dioxide, are discharged to the atmosphere.



Although boiler performance is identical to the Base Case in Concept A, there is a major impact to the steam cycle system where low-pressure steam is extracted to provide the energy for solvent regeneration. About 79% of the intermediate pressure (IP) turbine exhaust is extracted from the IP/LP crossover pipe. This steam is expanded from 200 psia to 65 psia through a new steam turbine. The exhaust from the new turbine provides the heat requirement for solvent regeneration in the reboiler's of the MEA CO_2 recovery system. Solvent regeneration requires about $4.7x10^6$ Btu/Ton CO_2. The condensate from the reboilers is pumped to the existing deaerator. The remaining 21% of the IP turbine exhaust is expanded in the existing low-pressure turbine before being exhausted to the existing condenser. The total output from both the existing and new generators is 331,422 kW. This represents a gross output reduction of 132,056 kW (about 28%) as compared to the Base Case.

The basic concept of the overall system for **Concept B** is to replace air with oxygen for combustion in the furnace, thus directly producing a high CO_2 content flue gas. In designing the Concept B system, emphasis was placed on utilizing as much as possible of the existing equipment, minimizing boiler modifications required, and providing operational flexibility to permit switching to the conventional mode of operation if desired. This process is depicted schematically below.

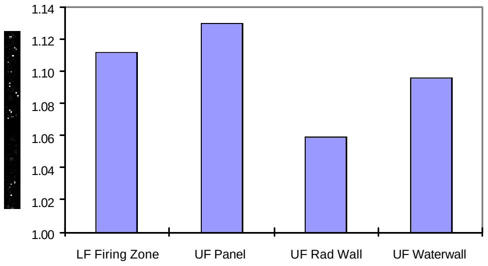


The proposed combustion system uses a mixture of nearly pure oxygen with recycled flue gas. A conventional cryogenic Air Separation Unit (ASU) supplies the oxygen. The recycle flow rate is established by the need to maintain a thermal balance between the radiant and convective heat transfer surfaces of the existing steam generator. The quantity of the recirculated gas is approximately equal to the quantity of nitrogen contained in the combustion air that would have been supplied to the steam generator as part of the oxidant. Inherent in the flue gas recirculation process is the build-up of carbon dioxide and water vapor content in the flue gas. This increased concentration produces significant changes in the thermal and mechanical properties of the flue gas. These gas property differences cause significant differences in the heat transfer processes, which occur within the steam generator unit. Analyses were made to determine the impact of the heat transfer differences on boiler behavior.

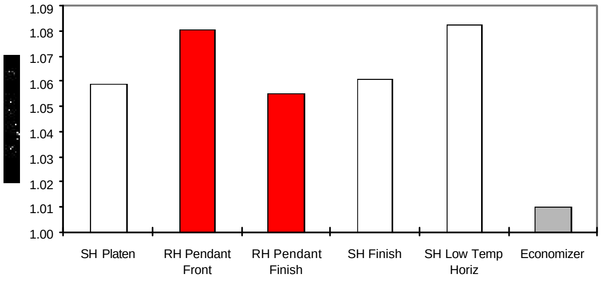
Heat transfer in the lower and upper furnace regions is compared in the figure to the left below. The figure on the left compares heat fluxes (Btu/hr-ft^2) in the lower and upper furnace region between air firing and oxygen firing. Lower furnace results show firing zone heat flux to be about 11% higher with oxygen firing. Upper furnace region results show the reheat radiant wall is about 6% higher and the superheat panels are about 13% higher with oxygen firing. Similarly, the upper furnace waterwall region is about 10% higher.

Convective heat transfer in utility steam generator units is dependent upon many of the transport properties of the flue gas (viscosity, thermal conductivity, density, specific heat and others). Additionally, convection depends strongly on gas velocity. With this system, there are significant changes in the flue gas analysis as compared with air firing. These gas analysis changes cause both transport property and gas velocity changes throughout the unit. Significant differences in non-luminous radiant heat transfer are also expected. Of the gases produced by the complete combustion of a fuel, only carbon dioxide, water vapor and sulfur dioxide emit radiation over a sufficiently wide band of wavelengths to warrant consideration. With this system the primary change in the flue gas as compared to air firing is the large increase in the CO_2 content and decrease in N_2 content. The total heat transfer rates (convective + non-luminous radiation) for the convection pass are shown in the figure on the right, below. Increases are calculated to be in the range of 1 to 8% for oxygen firing over the values with air firing.

Steam Generator Heat Transfer Comparison
(Oxygen and Air Firing)



Furnace Region Heat Flux Comparison



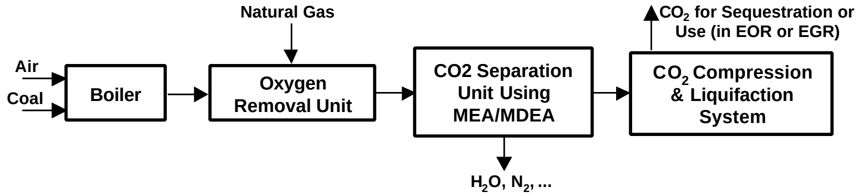
Convection Pass Total Heat Transfer Rate Comparison

Despite these heat transfer changes, boiler performance is achieved with the existing steam temperature control system that includes burner tilts and desuperheating sprays. Although more detailed analysis are needed, no boiler pressure parts modifications appear to be required.

The products of combustion that exit the furnace are cooled in a series of heat exchangers. Two of these heat exchangers, located downstream of the existing air heater, are new. The first heat exchanger is provided to preheat the oxygen required for combustion. The second heat exchanger recovers additional heat from the flue gas in a low temperature economizer installed in parallel with the low-pressure extraction feedwater heaters. This heat exchanger was necessary as a result of reduced air heater performance with oxygen firing.

The steam cycle system for Concept B is very similar to the Base Case. The only modification was the addition of a low-pressure feedwater heater arrangement in parallel with two low-pressure extraction feedwater heaters. The modified steam cycle system produces almost exactly the same output as the Base Case.

In **Concept C**, coal is burned conventionally in air. A process comprising an optimized mixture of monoethanolamine (MEA) and methydiethanolamine (MDEA), installed downstream of the flue gas desulfurization unit, is integrated into the power plant to strip CO_2 from the effluent gas stream. The mixture of MEA and MDEA cannot be made to be oxygen-resistant. Therefore, while this process potentially offers an improved system from the standpoint of solvent regeneration energy requirement, it is necessary that the excess oxygen in the flue gas be converted to CO_2 by combustion with natural gas over a De-Oxy catalyst upstream of the solvent contactor. Solvent regeneration requires about $3.4x10^6$ Btu/Ton CO_2 (about 72% of that required for Concept A). High temperature heat recovery is provided in the De-Oxy system by the generation of high pressure superheated steam for power generation and solvent regeneration. Concept C is depicted schematically below.



Boiler performance for Concept C is again identical to the Base Case. There is a major impact to the steam cycle system, similar to Concept A, where low-pressure steam is extracted from the existing steam turbine to provide energy for solvent regeneration. In this case, about 45% of the IP turbine exhaust is extracted from the IP/LP crossover pipe. This steam is expanded from 200 psia to about 65 psia through a new low-pressure steam turbine. The exhaust from the new turbine provides most of the heat requirements for solvent regeneration in the reboiler's of the MEA/MDEA CO_2 capture system. The high-pressure steam generated in the De-Oxy heat recovery system is expanded through a new high-pressure steam turbine for additional power generation. The exhaust from this turbine provides about 20% of the energy requirement for the reboilers. Low temperature heat recovery is provided in the De-Oxy system with a low-pressure feedwater heater which is located in a feedwater stream which is in parallel with the three existing low pressure extraction feedwater heaters. The total output from the modified steam cycle is 431,290 kW. This represents a gross output reduction of 32,188 kW, which represents about 7% of the Base Case output.

Performance Analysis Results

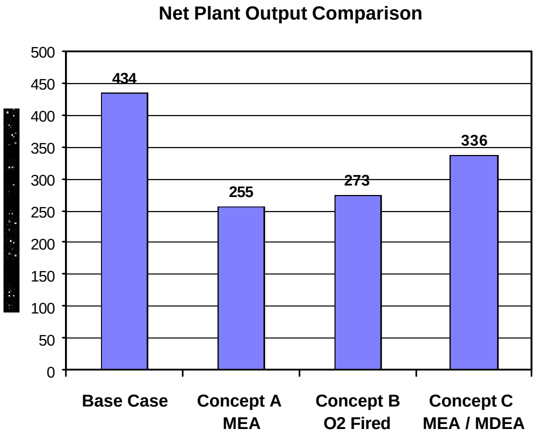
The table shown below summarizes the performance differences between the cases. The primary design constraint among all cases was the supply of an equivalent main steam flow to the existing steam turbine.

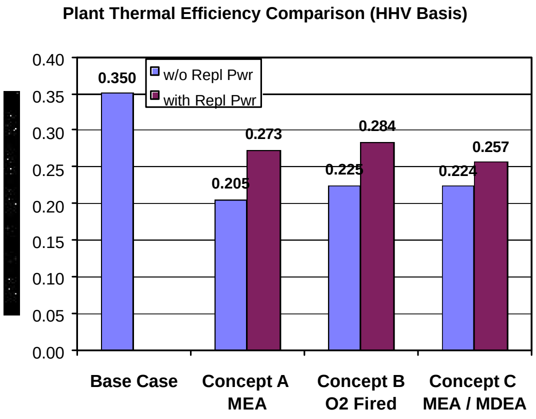
	(units)	Original Plant (Base)	Concept A MEA	Concept B O2 Fired	Concept C MEA-MDEA
~~***Fuel Paramaters***~~					
Coal Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4228.7	4140.0	4228.7
Natural Gas Heat Input (HHV)	(10^6 Btu/hr)	---	17.7	11.4	885.9
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4246.4	4151.5	5114.6
Steam Cycle Paramaters					
Existing Steam Turbine Generator Output	(kW)	463478	269341	463056	357196
CO2 Removal System Turbine Generator Output	(kW)	0	62081	0	36343
De-Oxy System Turbine Generator Output (Concept C)	(kW)	0	0	0	37751
Total Turbine Generator Output	(kW)	463478	331422	463056	431290
Total Auxiliary Power	(kW)	29700	76007	189709	95317
Net Plant Output	(kW)	433778	255414	273347	335973
Overall Plant Performance Paramaters					
Net Plant Efficiency (HHV)	(fraction)	0.3501	0.2053	0.2247	0.2242
Net Plant Efficiency (LHV)	(fraction)	0.3666	0.2150	0.2354	0.2371
Normalized Efficiency (HHV; Relative to Base Case)	(fraction)	1.0000	0.5864	0.6419	0.6404
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9749	16626	15188	15223
Net Plant Heat Rate (LHV)	(Btu/kwhr)	9309	15872	14500	14395

As described above, significant reductions in Net Plant Output are incurred as a result of the CO_2 capture systems (refer to the figure to the left below). Therefore, each case was also analyzed with replacement power to make up this difference. For cases with replacement power, it was assumed to be generated with a state-of-the-art natural gas fired combined cycle (NGCC) plant. The NGCC plant was analyzed without capturing the CO_2 from its flue gas. The NGCC plant was assumed to generate power with an efficiency of 57.1 percent (LHV basis). The additional CO_2 emitted from the NGCC was 0.762 lbm/kWh.

Net plant efficiency is reduced from about 35% (HHV basis) for the Base Case to about 20.5% for Concept A and about 22.5% for Concepts B and C (without replacement power) as shown in the figure on the right below. These efficiencies represent energy penalties of 41% and 36%, respectively. The efficiency reductions for Concepts A and C are due to large reductions in the existing steam turbine output and significant auxiliary power requirement increases. Concept C additionally utilizes a large quantity of natural gas in the de-oxy system increasing the total fuel heat input by about 21%. The steam turbine output reductions for Concepts A and C result from solvent regeneration energy requirements, which are provided by low-pressure steam extraction. The auxiliary power increases are due to the CO_2 compression and liquefaction system. Concept B, with a gross output essentially the same as the Base Case, suffers from high auxiliary power primarily from the ASU and the CO_2 compression and liquefaction system.

The efficiencies for these cases including replacement power are also shown on this figure and range from about 26% to 28% (HHV basis) with Concept B being the most efficient.

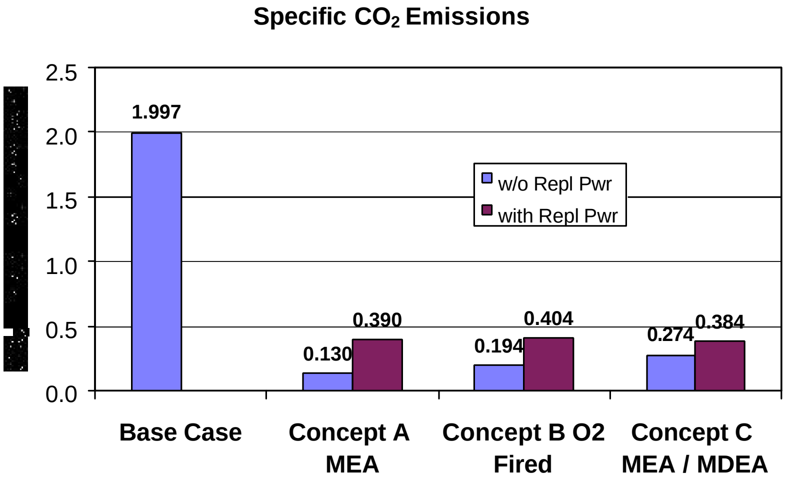




Plant CO_2 Emissions

Specific carbon dioxide emissions were reduced from about 2 lbm/kWh for the Base Case to between 0.13 – 0.27 lbm/kWh for the study cases as shown in the following table and figure (without replacement power). Recovery of CO_2 ranged from 91 to 96%. The CO_2 emissions with replacement power are also indicated on the figure below.

	(units)	Original Plant (Base)	Concept A MEA	Concept B O2 Fired	Concept C MEA-MDEA
Plant CO_2 Emissions					
Carbon Dioxide Produced	(lbm/hr)	866102	868137	849255	967806
Carbon Dioxide Recovered	(lbm/hr)	0	835053	796238	875653
Carbon Dioxide Emissions	(lbm/hr)	866102	33084	53016	92153
Fraction of Carbon Dioxide Recovered	(fraction)	0	0.962	0.938	0.905
Specific Carbon Dioxide Emissions	(lbm/kwhr)	1.997	0.130	0.194	0.274
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.065	0.097	0.137
Avoided Carbon Dioxide Emissions (as compared to Base)	(lbm/kwhr)	---	1.867	1.803	1.722



Project Costs

The project capital cost estimates, including construction, are shown in the following table and figure. These costs include all required equipment such as the amine based CO_2 scrubbing systems of Concepts A and C, the modified FGD system of Concept A, the de-Oxy system of Concept C, the CO_2 compression and liquefaction systems for all three concepts, steam cycle modifications for all concepts, the air separation unit and the boiler island modifications of Concept B. Boiler island modifications for Concept B are relatively small and include costs for such items as air and gas ducts and dampers, booster fans, parallel feedwater heaters and piping, O_2 heater, and controls and instrumentation.

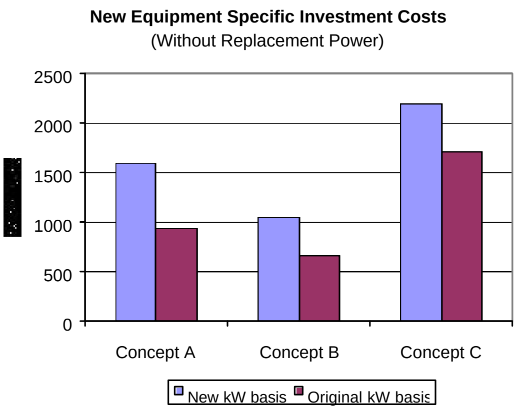
All these CO_2 capture options produce less net plant output than the original plant (Base Case). Therefore, each case was also analyzed with replacement power to make up this difference. For cases with replacement power, it was assumed to be generated with a state-of-the-art natural gas fired combined cycle (NGCC) plant. The NGCC plant was analyzed without capturing the CO_2 from its flue gas. The NGCC plant was assumed to cost $450/kW, installed and its efficiency (LHV basis) was 57.1 percent. The additional CO_2 emitted from the NGCC was 0.762 lbm/kWh.

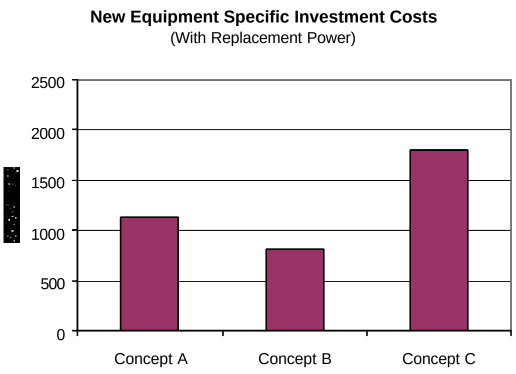
Two sets of costs are shown for each Concept in the table, one without replacement power and one including replacement power. The figure on the left shows new equipment specific investment costs ($/kW net) for the three concepts, without replacement power, based on both the original and modified net output. The figure on the right shows new equipment specific investment costs ($/kW net) for the three concepts, with replacement power, and therefore based on the original net output.

Total Retrofit Investment Capital Costs

CO_2 Capture Concept	Units	w/o Repl Pwr	with Repl Pwr
Concept A	10^6 US$	409	489
	$/kW	1602	1128
Concept B	10^6 US$	285	357
	$/kW	1042	823
Concept C	10^6 US$	738	782
	$/kW	2197	1803

Note: The specific costs ($/kW) shown above for cases without replacement power are based on the new net kW output.


New Equipment Specific Investment Costs
(Without Replacement Power)


New Equipment Specific Investment Costs
(With Replacement Power)

Operating and maintenance (O&M) costs were calculated for all systems. The O&M costs for the Base Case were provided by AEP. The fixed O&M costs (FOM) for the existing plant were about $16/kW (0.27 Cents/kWh using the base parameter assumptions) and the variable O&M costs (VOM) were about 0.45 Cents/kWh. The VOM costs are comprised exclusively of the lime cost required for the FGD system and the costs for fly ash, bottom ash and FGD system waste disposal.

For the retrofit CO_2 capture system evaluations, additional O&M costs were calculated for the new equipment. The VOM costs for the new equipment included such categories as chemicals and desiccants, waste handling, maintenance material and labor, and contracted services. The FOM costs for the new equipment includes operating labor only.

For the variable ranges considered in the economic sensitivity study, total FOM costs for the retrofit systems, including existing and new equipment, ranged from about 0.4 to 0.6 Cents/kWh and VOM costs ranged from about 0.9 to 2.3 Cents/kWh.

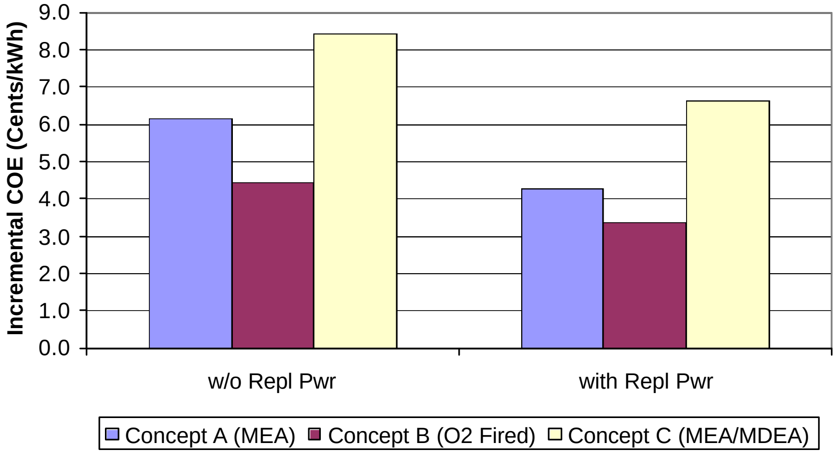
Economic Evaluation

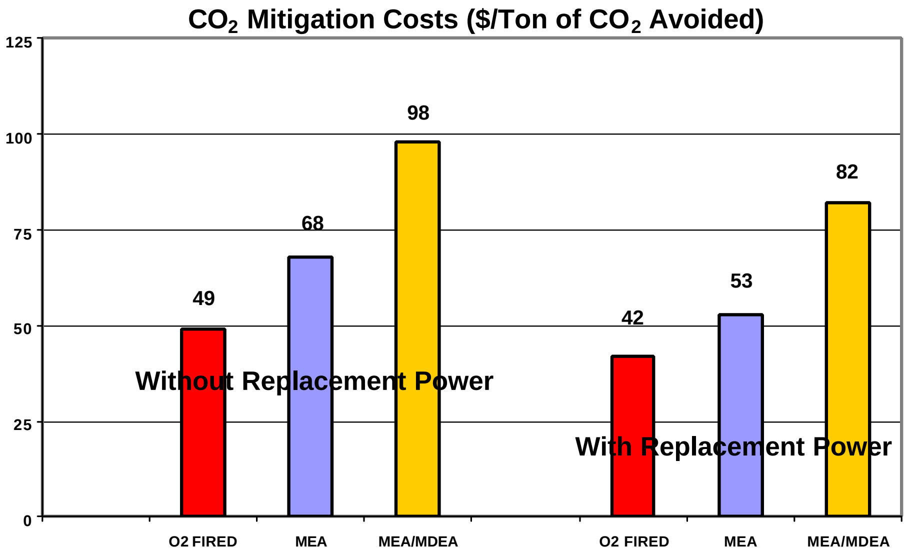
A total of 66 CO_2 capture cases were compared in the reported sensitivity studies. All cases studied indicate significant increases to the COE as a result of CO_2 capture. The incremental COE as compared to the Base Case (air firing without CO_2 capture) ranges from 1.5 to 9.8 Cents/kWh. Similarly CO_2 mitigation costs range from about 17-113 \$/Ton of CO_2 avoided for the range of cases studied. The following table and figures summarize the economic analysis results for the six primary cases using base parameter values (i.e.; Coal Cost = 1.32 $\$/10^6$ Btu, Natural Gas Cost = 4.0 $\$/10^6$ Btu, Capacity Factor = 67%, CO_2 Byproduct Sell Price = 0.0 \$/Ton, Investment Cost = As Estimated).

Summary of Economic Analysis Results
(Using base parameter values)

Parameter	Units	Concept A		Concept B		Concept C	
		w/o Repl Pwr	with Repl Pwr	w/o Repl Pwr	with Repl Pwr	w/o Repl Pwr	with Repl Pwr
Incremental COE	Cents/kWh	6.2	4.3	4.4	3.4	8.4	6.6
CO2 Mitigation Cost	$/Ton	68	53	49	42	98	82

Incremental Cost of Electricity
(Using base parameter values)





Concept B (oxygen firing with flue gas recycle) appears to clearly be the best alternative of the three concepts studied based on incremental COE or mitigation cost evaluation criteria. Using the base parameters of the sensitivity study for Concept B yields incremental COE values of 3.4 and 4.6 Cents/kWh for cases with and without replacement power respectively. The actual COE values for Concept B are 19% and 12% lower than the corresponding values for Concept A and 47% and 51% lower than the corresponding values for Concept C for cases with and without replacement power respectively. Similar results are also obtained when comparing mitigation costs.

If CO_2 byproduct were sold for $20/ton, incremental COE values would be reduced to 1.59 and 1.64 Cents/kWh respectively. Additional reductions would be possible with capacity factor increases, investment cost decreases or system efficiency increases as quantified in the sensitivity studies. Significant investment cost decreases and efficiency increases may be possible as oxygen transport membrane technology develops. Previous studies, Liljedahl, et al. (1999), have shown membrane technology to provide specific investment cost ($/kW) reductions of about 30% and net plant heat rate improvements greater than 20%.

Conclusions and Recommendations

No major technical barriers exist for retrofitting AEP's Conesville Unit #5 to capture CO_2 with any of the three concepts considered under this study. Nominally, 5-8 acres of new equipment plan area is needed. For this site, new equipment is located approximately 1500 feet north of the Unit #5 stack on the existing ~200 acre power plant site which accommodates a total of 6 units with a total power generating capability of 2,080 MW.

Energy requirements and power consumption are high, resulting in significant decreases in overall power plant efficiencies (HHV basis), ranging from about 21 to 23% as compared to 35% for the Base Case. The MEA/MDEA mixture of Concept C requires about 28% less energy per pound of CO_2 captured to regenerate the solvent as compared to the MEA used for Concept A.

Specific investment costs are also high ranging from about 1000 to 2200 $/kW without replacement power, and from about 800 to 1800 $/kW with replacement power. All cases studied indicate significant increases to the COE as a result of CO_2 capture. The incremental COE as compared to the Base case (air firing without CO_2 capture) ranges from 3.4 to 8.4 ¢/kWh. Similarly CO_2 mitigation costs range from about 42-98 $/ton of CO_2 avoided for the range of cases studied.

Specific carbon dioxide emissions were reduced from about 2.0 lbm/kWh for the Base Case to 0.13, 0.19, and 0.27 lbm/kWh for study cases A, B, and C, respectively. Recovery of CO_2 ranged from 91 to 96%.

Concept B (oxygen firing with flue gas recycle) appears to clearly be the best alternative of the three concepts studied from both an efficiency and incremental COE viewpoint for systems designed for very high CO_2 capture (i.e. > 90%). If lower CO_2 capture fractions are considered, it appears that Concept A would likely be the best alternative for capture fractions below some as yet undetermined value. Concept C would also improve considerably with lower capture fractions. This study has confirmed two important issues related to firing coal in a CO_2 rich flue gas /O_2 mixture:

- Modifications to the existing steam generator unit pressure parts may not be required, and as such will also allow the unit to continue to operate in the conventional air fired mode. This is an important consideration as it is unlikely that much new coal-fired capacity can be built in the short term.

- CO_2 rich flue gas can be cleaned and compressed with a relatively simple system to provide high purity CO_2 for usage or sequestration.

While overall plant performance penalty and costs are high, the integration of power generation, air separation, and CO_2 separation, compression and liquifaction systems may significantly improve the overall system efficiency and economics. The major cost item in Concept B is in the air separation plant, both from a capital and an operating cost point of view. While in recent years advances have been made in air separation technology which have steadily improved the costs, new membrane based technologies are being researched which promise to provide greatly reduced auxiliary power requirements and investment costs which will significantly improve the overall system efficiency and economics. Previous studies, Liljedahl, et al. (1999), have shown membrane technology to provide specific investment cost ($/kW) reductions of about 30% and net plant heat rate improvements greater than 20%.

Barring governmental mandates, it is clear that none of these three retrofit concepts will be acceptable to the electric utility industry from the standpoint of cost competitiveness.

Recommendations for Future Work are listed below:

- A sensitivity study, for Concepts A and C, showing the impact of reducing CO_2 capture percentage on plant efficiency, investment cost, emissions, and cost of electricity. The current work utilized 90% CO_2 capture (nominal). The sensitivity study would investigate nominal CO_2 capture percentages of 70% and 50%.

- Detailed analysis of the existing steam turbine for Concept A: In Concept A about 79% of the steam leaving the intermediate-pressure (IP) turbine is extracted from the IP/LP crossover pipe for solvent regeneration. The capability of the existing low-pressure (LP) turbine to operate under these conditions of very low steam flow should be investigated in detail, preferably by the original equipment manufacturer (OEM).

- Optimization of the amine system reboiler steam pressure for Concepts A and C.

- Concept B detailed boiler system analysis: A startup/shutdown procedure and system design, particularly the transition from air to oxygen firing (including transient conditions) should be developed. Detailed metal temperature analysis for all heat exchanger sections, including operation at part loads should be analyzed. This should also include furnace wall metal temperatures and analysis of the circulation system. The existing fans should be checked (preferably by the OEM's) for operation under the new conditions. The feasibility of operating the boiler under a slight positive pressure to eliminate air infiltration should also be investigated.

- Investigation of Improved oxygen production systems for Concept B, in line with the membrane-based air separation work being conducted by various research groups (e.g., Praxair, Air Products, Norsk Hydro). Also optimization of an integrated boiler and oxygen production system.

- Measurement of furnace heat transfer in CO_2/O_2 environments in a proof of concept boiler simulation facility.

- Improved solvents, which require lower regeneration energy requirements and/or can be regenerated at a lower temperature level, similar to MHI’s KS-1 and KS-2 solvents (Iijima, et al., Feb. 1998), but for coal-firing application.

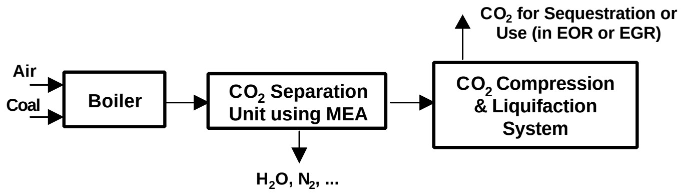
- Hybrid process using oxygen-enriched combustion and amine based CO_2 absorption, to accrue, simultaneously, both CO_2 capture and drastic NOx emissions reduction.

- Investigation of a new novel high risk CO_2 capture process that would reduce efficiency penalty and retrofit investment cost. This would likely not utilize the existing boiler.

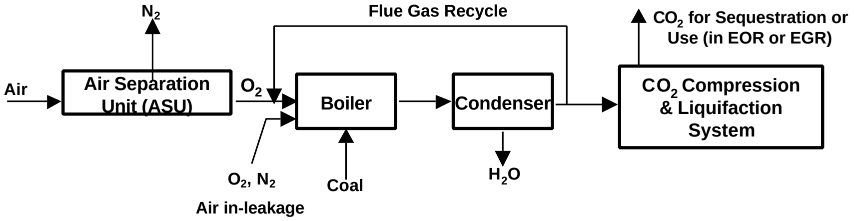
1.0 INTRODUCTION

ALSTOM Power Inc. teamed with American Electric Power, ABB Lummus Global Inc., the US DOE NETL, and the Ohio Coal Development Office to conduct a comprehensive study evaluating the technical feasibility and economics of alternate CO_2 capture and sequestration technologies applied to an existing US coal-fired electric generation power plant. The three retrofit technology concepts being evaluated are shown below.

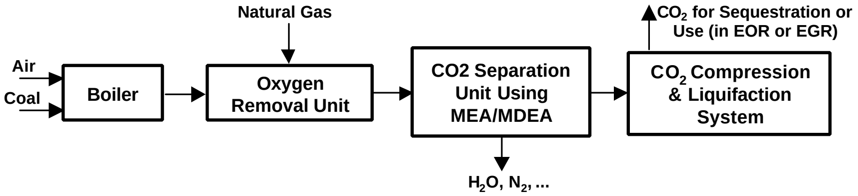
- Concept A: Coal combustion in air, followed by CO_2 separation from flue gas with Kerr-McGee/ABB Lummus Global's commercial MEA-based absorption/stripping process.



- Concept B: Coal combustion with O_2 firing and flue gas recycle.



- Concept C: Coal combustion in air with oxygen removal and CO_2 separation from flue gas by a mixture of primary and tertiary Amines



Each of these technologies was evaluated against a Base Case from the standpoints of performance and impacts on power generating cost. The Base Case represents the "business as usual" operation scenario for the plant without CO_2 capture. A typical existing US domestic pulverized coal-fired power plant is being used in this evaluation. Specifically, American Electric Power's (AEP's) 450 MW Conesville Unit No. 5, located in Conesville, Ohio is the power plant case study. All technical performance and cost results associated with these options are being evaluated in comparative manner.

ALSTOM Power Inc. is managing and performing the subject study from its US Power Plant Laboratories office in Windsor, CT. ABB Lummus Global, from its offices in Houston, Texas, is participating as a sub-contractor. American Electric Power is participating by offering their Conesville Generating Station as the case study and cost sharing consultation, and relevant technical and cost data. AEP is one of the largest US utilities and is the largest consumer of Ohio coal, and as such, is bringing considerable value to the project. Similarly, ALSTOM Power and ABB Lummus Global are well established as global leaders in the design and manufacture of power generation equipment, petrochemical and CO_2 separation technology. ALSTOM Power Performance Projects and Environmental Business Units are world leaders in providing equipment and services for boilers and power plant environmental control, respectively, and are providing their expertise to this project. The US Department of Energy (US DOE)

National Energy Technology Laboratory (NETL) and the Ohio Coal Development Office (OCDO) provided consultation and funding. All participants provided cost share to this project.

The motivation for this study was to provide input to potential US electric utility actions to meet Kyoto protocol targets. If the US decides to reduce CO_2 emissions consistent with the Kyoto protocol, action would need to be taken to address existing power plants. Although fuel switching from coal to gas is a likely scenario, it will not be a sufficient measure and some form of CO_2 capture for use or disposal may also be required. The output of this CO_2 capture study will enhance the public's understanding of control options and influence decisions and actions by government, regulators, and power plant owners to reduce their greenhouse gas CO_2 emissions.

The key objectives of the study were to evaluate the impacts on the plant output, efficiency, CO_2 emissions, investment costs, and the cost of generating electricity resulting from the addition of the CO_2 capture systems. All technical performance and cost results associated with these options are being evaluated in comparative manner. Technical and economic issues being evaluated include:

- Boiler performance and plant efficiency
- Purity of O_2 produced and flue gas recycled
- Heat transfer into the radiant and convective sections of the boiler
- NO_X, SO_2, CO and unburned carbon emissions
- Heat transfer surface materials
- Steam temperature control
- Boiler and Steam Cycle modifications
- Electrostatic Precipitator system performance
- Flue Gas Desulfurization system performance
- Plant systems integration and control
- Retrofit investment cost and cost of electricity (COE)
- CO_2 Mitigation Costs

Each of the CO_2 capture systems are designed for a minimum of 90% CO_2 capture. Cost estimates were developed for all the systems required to produce, extract, clean, compress and liquefy the CO_2, to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO_2 specification (2) 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

Component	Unit	Concentration
CO_2	Vol.%	96.0
H_2S	Vol.%	0.9
CH4	Vol.%	0.7
C2+HCs	Vol.%	2.3
CO	Vol.%	0.1
N_2	vppm	< 300
H_2O	vppm	< 20
O_2	vppm	< 50

The CO_2 product could then be available for usage in enhanced oil or gas recovery or sequestration. Additionally, an economic evaluation, showing the impact of CO_2 capture on the cost of electricity (COE), was developed for each concept. Included in the economic evaluation was a sensitivity study showing the effects of coal cost, natural gas cost, plant capacity factor, CO_2 byproduct sell price, investment cost, and replacement power, on the incremental cost of electricity and on the mitigation cost for the CO_2 ($/ton of CO_2 avoided).

2.0 STUDY UNIT SELECTION, DESCRIPTION AND PERFORMANCE

This section includes the criteria used for selection of the study unit, and provides a brief description of the selected unit. Additionally, the Base Case is defined as the unmodified existing unit firing coal at full load, utilizing air as the oxidant, without capture of CO_2 from the flue gas. This represents the "business as usual" operating scenario and is used as the basis of comparison for the three CO_2 removal options investigated in this study. The overall performance of the Base Case is presented in Section 2.3.

2.1 Study Unit Selection

A decision was required at the end of Task 2 (Site Selection/Plant Evaluation) regarding the selection of either unit #5 or #6 of AEP's Conesville, Ohio, Power Generating Station, as the study unit for our retrofit evaluations.

To provide the necessary background to facilitate this selection, a trip was made to the Conesville site on April 20, 2000. This visit provided the opportunity to gather operating data for unit #5, overall site information for both units #5 and 6, and additionally, to make arrangements for collecting a raw coal sample for Bench-Scale Testing (Task 4a).

It was learned from the operators and plant maintenance personnel that both twin units #5 and #6 operate nearly identically, and that no renovations had been made to either unit aside from replacement in kind. AEP, OCDO and Alstom personnel conducted a walk down of both units. Site drawings were provided and photographs were taken to document the footprints of both units. This information was used to evaluate space availability for the various new equipment that would be required for the three CO_2 capture systems being evaluated in this study. Figure 2.1.1 shows several potential areas for new equipment placement in the neighborhood of units #5 and #6. A large amount of additional land area (about 10 acres) is available in a location about 1,500 feet northeast of the common stack used for units #5 and #6 (upper right corner of Figure 2.1.1).

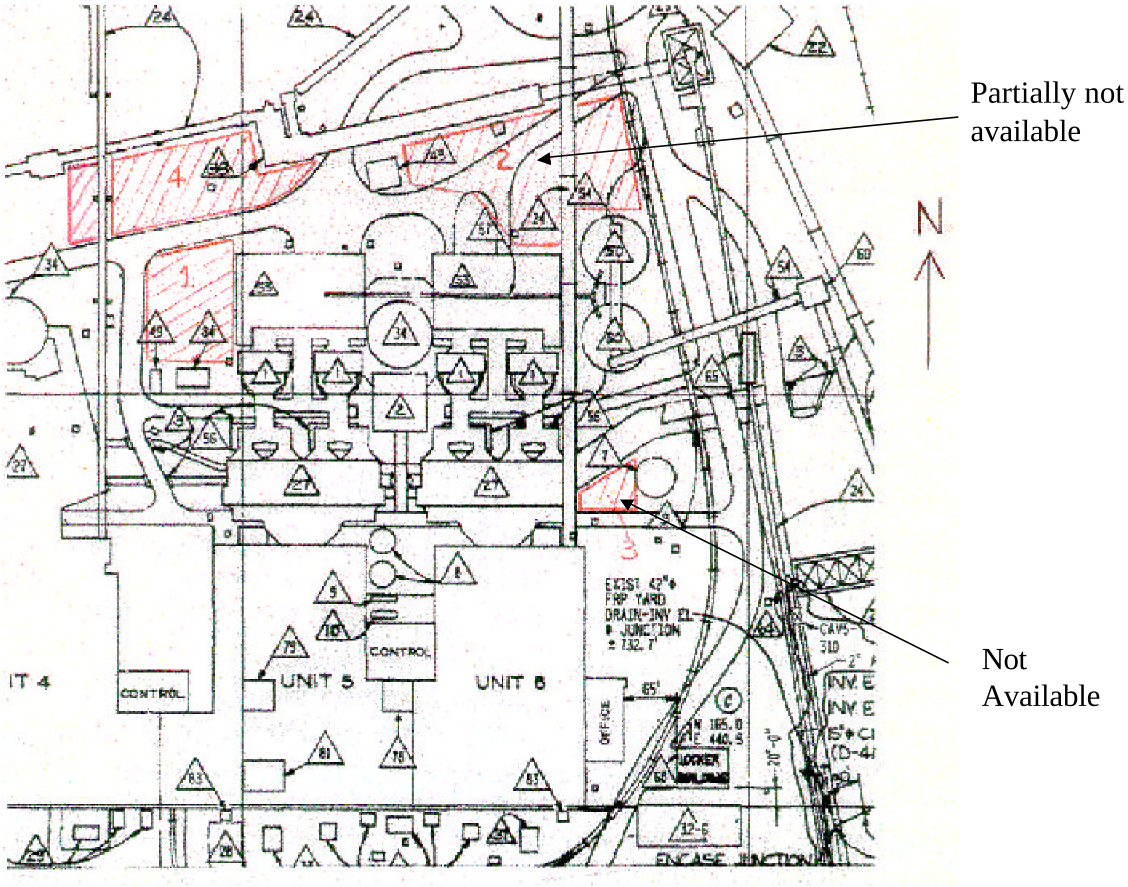
The approximate size of the four areas in the immediate vicinity of units #5 and #6 for locating some of the new equipment on the ground are listed below:

Area #	Approximate Size (square feet)
1	13,000 (Available)
2	29,000 (Partially available)
3	2,600 (Not available)
4	25,000 (Available)

Additionally, consideration was given for hanging some of the equipment, where appropriate, in the space between the far end of the ESP and the boiler: fifth floor and up.

The major new equipment envisioned for the three retrofit concepts for this study include:

- An Air Separation Unit of about 9,000 tons of O_2 per day (Concept B),
- An MEA system for removal of about 10,000 tons of CO_2 per day from the flue gas (Concept A),
- An MEA/MDEA system for removal of about than 10,500 tons of CO_2 per day from the flue gas (Concept C),
- A flue gas oxygen removal system upstream of the MEA/MDEA system (Concept C),
- A CO_2 Compression and Liquefaction system (required for all 3 concepts),
- Various other ancillary equipment



Partially not available

Not Available

Figure 2.1. 1: Plan View of AEP's Conesville, Ohio, Power Generating Station Site Showing Area Occupied by Units #5 and #6

The four areas identified previously and shown in Figure 2.1.1 indicate the following:

- These areas, even combined, are not large enough to locate all the equipment required for any one of these CO_2 removal concepts. They may however be useful for location of various ancillary equipment.

- Areas 1 and 4, which are in the vicinity of unit #5 total to ~38,000 square feet

- Areas 2 and 3, which are in the vicinity of unit #6, total ~31,000 square feet, of which less than 28,000 square feet are available.

- Furthermore, conveniently we already had performance data from unit #5.

- Although we could have obtained data from unit #6, this step could be eliminated with the selection of unit #5. Hence, it made more sense to use unit #5 in this initial evaluation. If there is in the future a desire to evaluate unit #6, then a further re-examination of areas #2 and #3 would be warranted and performance data from unit #6 could be obtained.

In summary, the preceding information indicated that:

- Unit #5 offers more "on-the-ground" area near the unit for erecting new equipment than does unit #6.

- Both units #5 and #6 are equally accessible for erecting new equipment in the space between the boiler and the far end of the ESP, fifth floor and up.

- Both units #5 and #6 are equally accessible to the large area available about 1,500 feet northeast of the unit #5/#6 stack.

- Both units #5 and #6 perform similarly with respect to power generation.

Considering the totality of the information given above, we concluded that unit #5 offered a relatively more cost-effective scenario for our retrofit evaluation. Cognizant personnel of the Conesville Power Generating Station were in complete agreement with these findings and Conesville Unit #5 was selected as the study unit.

2.2 Study Unit Description

The steam generator unit analyzed in this study was American Electric Power's Conesville Unit #5. A sectional side elevation drawing of the study unit is shown in Figure 2.2.1. This unit can be described as a nominal 450 MW-gross, tangentially coal fired, subcritical pressure, controlled circulation, radiant reheat wall unit. The furnace is a single cell design utilizing five elevations of tilting tangential coal burners. The unit fires mid-western bituminous coal. The coal is supplied to the five elevations with five RP-903 coal pulverizers. The unit is configured in a "Conventional Arch" type design and is representative in many ways of a large number of coal fired units in use today. The unit is designed to generate about 3.1×10^6 lbm/hr of steam at 2400 psi and 1005 F with reheat also to 1005 F. These represent the most common steam cycle operating conditions for existing utility scale power generation systems. Outlet steam temperature control is provided with de-superheating spray and burner tilt.

The furnace is tangentially fired and is of the single cell design. It is about 51.67 ft wide, 44.33 ft deep and 171.67 ft high. The superheater is divided into four major sections. Saturated steam leaving the steam drum first cools the roof and rear pass walls before supplying the low temperature superheater section. The low temperature superheater section is located in the rear pass of the unit and is a horizontal section with the outlet tubes in a vertical orientation adjacent to the finishing superheater section. Steam leaving the low temperature superheater section first flows through the de-superheater spray stations and then to the radiant superheat division panel section. The division panels are located in the upper furnace directly above the combustion zone of the lower furnace. Steam leaving the panels flows to the superheater platen section, which is a more closely spaced vertical section located between the panels and the finishing pendant reheater. Steam leaving the platens flows into the finishing superheater section which is also a pendant section located downstream of the pendant reheater, just before the gas turns downward to enter the low temperature superheater section in the rear pass of the unit. 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 de-superheating spray station.

The reheater is divided into two sections, a low temperature radiant wall section followed by a spaced finishing pendent section. Steam is supplied to the reheater radiant wall from the de-superheating spray station, which is fed from the high-pressure turbine exhaust. The reheater radiant wall section is located in the upper furnace and covers the entire front wall and most of the two sidewalls of the upper furnace. The pendant finishing reheat section is located above the arch between the superheat platen and superheat finishing sections. 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.

The gases leaving the low temperature superheater section are then further cooled in an economizer section. The economizer is comprised of four banks of spiral finned tubes (2 fins/inch) which heats high-pressure boiler feedwater before it is supplied to the steam drum. The feedwater supplying the economizer comes from the final extraction feedwater heater.

Finally, a Lungstrom trisector regenerative air heater is used to heat both the primary and secondary air streams prior to combustion in the lower furnace. Particulate matter is removed from the cooled flue gas leaving the air heater in an electrostatic precipitator and sulfur dioxide is removed in a lime based flue gas de-sulfurization system. The flue gas is then exhausted to the atmosphere through the stack. The induced draft and forced draft fans are controlled to operate the unit in a balanced draft mode with the furnace maintained at a slightly negative pressure (typically –0.5 inwg).

The high pressure superheated steam leaving the finishing superheater is expanded through the high pressure steam turbine, reheated in the two stage reheater and returned to the intermediate pressure turbine. The steam continues its expansion through the low-pressure turbine sections where it expands to condenser pressure. The generator produces about 463 MW of electric power at Maximum Continuous Rating (MCR). The steam cycle utilizes seven feedwater heaters (three low-pressure heaters, a deaerator, and three high-pressure heaters) where the feedwater is preheated to about 493 F before entering the economizer of the steam generator unit. The boiler feed pump is steam turbine driven with steam provided from the intermediate pressure turbine exhaust and expanded to condenser pressure.

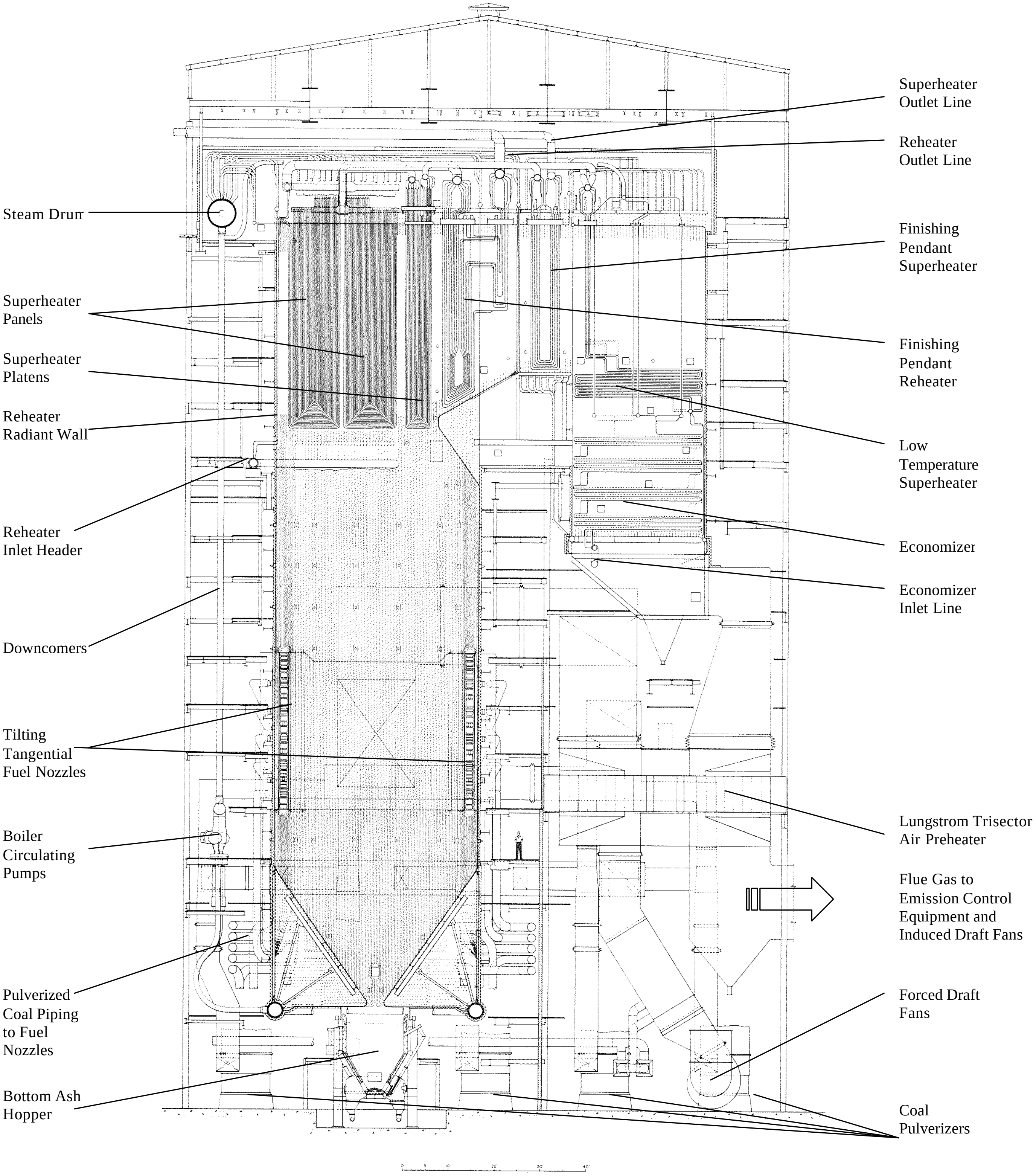


Figure 2.1. 2: Study Unit (Existing Conesville Unit #5 Steam Genertor)

2.3 Base Case Performance Analysis

The Base Case can be described as the unmodified existing unit firing coal at full load and utilizing air as the oxidant without capture of CO_2 from the flue gas. This represents the "business as usual" operating scenario and is used as the basis of comparison for the three the 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. Using test data from the existing unit, the computer model was then calibrated. The calibrated boiler model was then used for analysis of the Base Case and the three CO_2 removal concepts.

2.3.1 Calibration of Boiler Computer Model

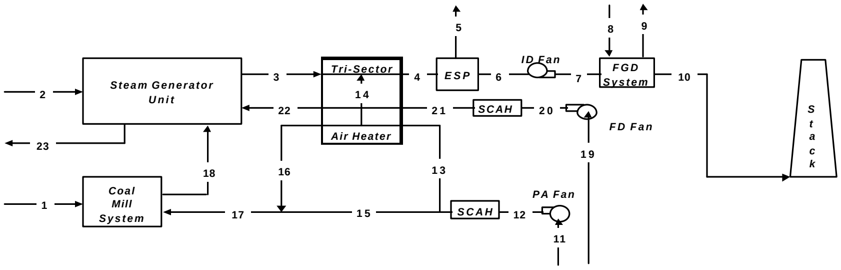
The first step in the calculation of a Base Case was to set up a steady state performance computer model of the Conesville #5 steam generator unit. This involves calculating or obtaining all the geometric information for the unit as required by the proprietary Reheat Boiler Program (RHBP). The RHBP provides an integrated, steady state performance model of the Boiler Island including, in addition to the steam generator unit, pulverizers, air heater, and steam temperature control logic. The RHBP is used to size components and/or predict performance of existing components. In this study, since the boiler island component sizes are known, the RHBP was used exclusively for calculating unit performance.

The next step in the heat transfer analysis of the Base Case was to calibrate the RHBP model of the unit. This involves obtaining test data (with air firing) for the existing unit and "adjusting" the performance model to match the test data. The required test data includes steam temperatures entering and leaving each major heat exchanger section in the unit, steam pressures, coal analysis, flue gas oxygen content, etc. The "adjustments or calibration factors" for the model are in the form of "surface effectiveness factors" and "fouling factors" for the various heat exchanger sections throughout the unit. Unfortunately, the test data used for calibration of this model was not totally complete and several assumptions were required in the calibration process. Although all the required data was not available, primarily due to existing instrumentation limitations, a satisfactory calibrated model was obtained.

Using the calibrated boiler model and providing it with new steam side inputs (mass flows, temperatures, and pressures) from the agreed upon MCR steam turbine material and energy balance, the model was run and performance was calculated for the Base Case. The performance for the overall power plant system is described in Section 2.3.2 with the boiler performance shown in Section 2.3.3 and the steam turbine performance in Section 2.3.4.

2.3.2 Overall System Description and Material and Energy Balance

The simplified gas side process flow diagram for the Base Case is shown in Figure 2.3.1and the associated material and energy balance for this case is shown in Table 2.3.1. Overall plant performance is summarized in Table 2.3.2. This system is described previously in Section 2.2. Boiler efficiency is calculated to be 88.13 percent. The net plant heat rate is calculated to be 9,749 Btu/kwhr for this case as shown in Table 2.3.2. Auxiliary power is 29,700 kw and the net plant output is 433,778 kw. Carbon dioxide emissions are 866,156 lbm/hr or about 2.00 lbm/kwhr.



Material Flow Stream Identification

1	Raw Coal to Pulverizers
2	Air Infiltration Stream
3	Fluegas from Economizer to Air Heater
4	Fluegas Leaving Air Heater to ESP
5	Flyash Leaving ESP
6	Fluegas Leaving ESP to Induced Draft Fan
7	Fluegas to Fluegas De-Sulfurization System
8	Lime feed to FGD System
9	FGD System Solids to Disposal
10	Fluegas to Stack
11	Air to Primary Air Fan
12	Primary Air to Steam Coil Air Heater
13	Primary Air to Air Heater
14	Air Heater Leakage Air Stream
15	Tempering Air to Pulverizers
16	Hot Primary Air to Pulverizers
17	Mixed Primary Air to Pulverizers
18	Pulverized Coal and Air to Furnace
19	Secondary Air to Forced Draft Fan
20	Secondary Air to Steam Coil Air Heater
21	Secondary Air to Air Heater
22	Heated Secondary Air to Furnace
23	Bottom Ash from Furnace

Figure 2.3. 1: Simplified Gas Side Process Flow Diagram (Base Case)

Table 2.3. 1: Base Case Gas Side Material and Energy Balance

Constituent	**(Units)**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**
O2	(Lbm/hr)	26586	42147	101097	144817		144817	144817	5355		144578	203237	203237	112918
N2	"	4868	139626	2797385	2942220		2942220	2942220			2942220	673283	673283	374075
H2O	"	37820	2357	228849	231294		231294	231294	250709	45979	436024	11365	11365	6314
CO2	"			867210	867210		867210	867210			866156			
SO2	"			20202	20202		20202	20202			1063			
H2	"	16102												
Carbon	"	236655												
Sulfur	"	10110												
Ca	"								12452					
Mg	"								584					
MgO	"									484				
MgSO3	"									1293				
MgSO4	"									94				
CaSO3	"									35179				
CaSO4	"									2468				
CaCO3	"									2398				
Ash / Inerts	"	42313		33851	33851	33851			968	968				
		Raw Coal	Leakage Air	Fluegas to AH	Fluegas to ESP	Flyash	Fluegas to ID Fan	Fluegas to FGD	Lime Slurry	FGD Disposal	Fgas to CO2 Sep	Pri Air to PA Fan	PA from PA Fan	Pri Air to AH
Total Gas	(Lbm/hr)		184130	4014743	4205743		4205743	4205743			4390042	887885	887885	493308
Total Solids	"	**374455**		33851	33851	33851			14003	42884				
Total Flow	"	374455	184130	4048594	4239594	33851	4205743	4205743	270067	88863	4390042	887885	887885	493308
Temperature	(Deg F)	**80**	80	**706**	**311**	311	311	325	80	136	136	**80**	92	92
Pressure	(Psia)	14.7	14.7	14.6	14.3	14.7	14.2	15.0	14.7	14.7	14.7	14.7	15.6	15.6
hsensible	(Btu/lbm)	0.000	0.000	161.831	57.924	57.750	57.924	61.384	0.000	14.116	14.543	0.000	2.899	2.899
Chemical	106 Btu/hr	4228.715												
Sensible	106 Btu/hr	0.000	0.000	655.007	245.567	1.955	243.612	258.166	0.000	3.314	63.916	0.000	2.574	1.430
Latent	106 Btu/hr	**0.000**	2.475	240.291	242.858	0.000	242.858	242.858	**0.000**	0.000	464.020	11.933	11.933	6.630
Total Energy(1)	106 Btu/hr	4228.715	2.475	895.298	488.425	1.955	486.470	501.024	0.000	3.314	527.936	11.933	14.507	8.060

Constituent	**(Units)**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**
O2	(Lbm/hr)	43720	90319	66680	156999	183585	641283	641283	641283	643801	
N2	"	144835	299208	220899	520107	524975	2124443	2124443	2124443	2132785	
H2O	"	2445	5051	3729	8779	46599	35860	35860	35860	36001	
CO2	"										
SO2	"										
H2	"					16102					
Carbon	"					236655					
Sulfur	"					10110					
Ca	"										
Mg	"										
MgO	"										
MgSO3	"										
MgSO4	"										
CaSO3	"										
CaSO4	"										
CaCO3	"										
Ash / Inerts	"					42313					8463
		Air Htr Lkg Air	Tempering Air	Hot Pri Air	Mixed Pri Air	Coal-Pri Air Mix	Sec Air to FD	Sec Air to SCAH	Sec Air to AH	Hot Sec Air	Bottom Ash
Total Gas	(Lbm/hr)	191000	**394577**	291308	685885		2801587	2801587	2801587	2812587	
Total Solids	"										8463
Total Flow	"	191000	394577	291308	685885	1060340	2801587	2801587	2801587	2812587	8463
Temperature	(Deg F)	92	92	**666**	**339**		**80**	86.4	**86.4**	**616**	**2000**
Pressure	(Psia)	15.6	15.6	15.6	15.6	15.0	14.7	15.2	15.1	14.9	14.7
hsensible	(Btu/lbm)	2.899	2.899	145.249	63.358		0.000	1.549	1.549	132.582	480.000
Chemical	106 Btu/hr)					4228.715					
Sensible	106 Btu/hr	0.554	1.144	42.312	43.456		0.000	4.341	4.341	372.898	4.062
Latent	106 Btu/hr	2.567	5.303	3.915	9.218		37.653	37.653	37.653	37.801	0.000
Total Energy(1)	106 Btu/hr	3.121	6.447	46.227	52.674	4281.389	37.653	41.994	41.994	410.699	4.062

Table 2.3. 2: Overall Plant Performance Summary (Base Case)

	(units)	Original Plant (Base)
Fuel Paramaters		
Coal Heat Input (HHV)	(10^6 Btu/hr)	4228.7
Natural Gas Heat Input (HHV)	(10^6 Btu/hr)	---
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	4228.7
Steam Cycle Paramaters		
Existing Steam Turbine Generator Output	(kW)	463478
CO2 Removal System Turbine Generator Output	(kW)	0
Total Turbine Generator Output	(kW)	463478
Total Auxiliary Power	(kW)	29700
Net Plant Output	(kW)	433778
Overall Plant Performance Paramaters		
Net Plant Efficiency (HHV)	(fraction)	0.3501
Net Plant Efficiency (LHV)	(fraction)	0.3666
Normalized Efficiency (HHV; Relative to Base Case)	(fraction)	1.0000
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9749
Net Plant Heat Rate (LHV)	(Btu/kwhr)	9309
Overall Plant CO_2 Emissions		
Carbon Dioxide Emissions	(lbm/hr)	866102
Specific Carbon Dioxide Emissions	(lbm/kwhr)	1.997
Normalized Specific CO2 Emissions (Relative to Base Case)	(fraction)	1.000
Avoided Carbon Dioxide Emissions (as compared to Base)	(lbm/kwhr)	---
Specific Carbon Dioxide Emissions	(kg/kwhr)	0.906
Avoided Carbon Dioxide Emissions (as compared to Base)	(kg/kwhr)	---

2.3.3 Boiler Analysis Results

The main steam flow for this case and all other cases in this study is 3,131,619 lbm/hr. The cold reheat flow leaving the high-pressure turbine for this case and all other cases in this study is 2,766,780 lbm/hr. The hot reheat flow (including de-superheating spray) returning to the intermediate pressure turbine for this case is 2,853,590 lbm/hr. The overall steam conditions produced by the existing Conesville #5 steam generator unit are shown in Table 2.3.3 below. To produce these conditions, the superheat circuit requires about 3.6 percent spray and the reheat circuit requires about 3.1 percent spray to maintain required steam outlet temperatures. The burner tilts are –10 degrees (the minimum value the customer uses). The boiler was fired with 15 percent excess air and the resulting boiler efficiency calculated for this case was 88.13 percent with an air heater exit gas temperature of 311 F.

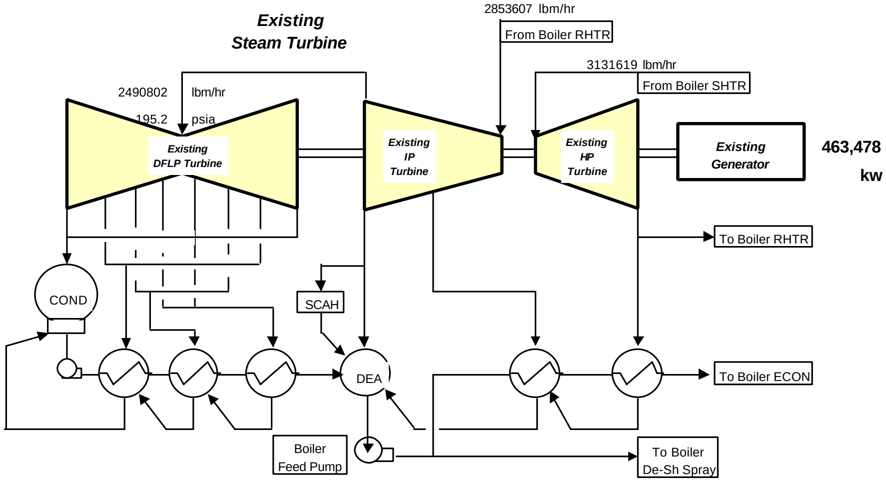
Table 2.3. 3: Base Case Boiler/Turbine Steam Flows and Conditions

		SHO	FWI	ECO	RHO	RHI
Mass Flow	(lbm/hr)	**3131619**	3131619	**3017507**	**2853590**	2853590
Pressure	(psia)	**2535**	3165	3070	**590.8**	**656.5**
Temperature	(Deg F)	**1005**	**492.6**	630	**1005**	**607.7**
Enthalpy	(Btu/lbm)	1459.7	479.2	**652.8**	1520.4	1290.4

Notes:
SHO = Superheater Outlet; FWI = Feedwater Inlet; ECO = Economizer Outlet; RHO = Reheater Outlet; RHI = Reheater Inlet

2.3.4 Steam Cycle Performance

The steam cycle for the Base Case is shown schematically in Figure 2.3.2. Figure 2.3.3 shows the associated Mollier diagram which illustrates the process on enthalpy - entropy coordinates. The high pressure turbine expands about 3.1 x 10^6 lbm/hr of steam at 2535 psia and 1005 F. Reheat steam is returned to the intermediate pressure turbine at 591 psia and 1005 F. These conditions (temperatures, pressures) represent the most common steam cycle operating conditions for existing utility scale power generation systems in use today. The condenser pressure used for the Base Case and all other cases in this study was 2.5 in Hga. The steam turbine performance analysis results show the generator produces 463,478 kw output and the steam turbine heat rate is about 7999 Btu/kwhr.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)	Turbine Heat Rate
Steam Turbine Power Output	1606	Boiler Heat Input	3707	7999
Steam Coil Air Heater Output	0	Condensate Pump	2	(Btu/kwhr)
Condenser Loss	2103	Total Energy Input	3709	
Total Energy Output	3709	In - Out	0	

Figure 2.3. 2: Base Case Steam Cycle Diagram and Performance

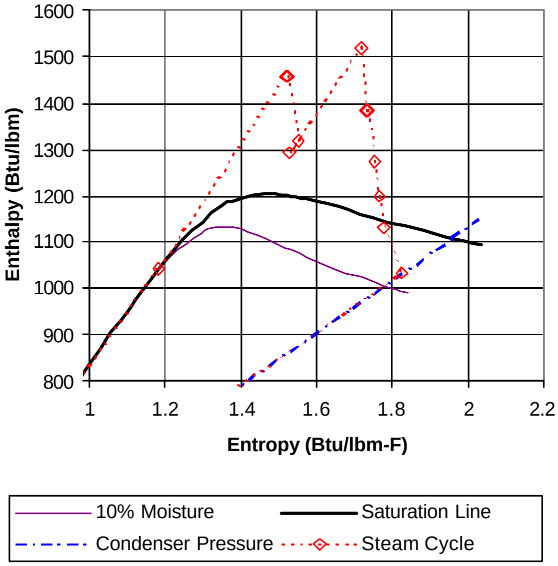


Figure 2.3. 3: Base Case Steam Cycle Mollier Diagram

2.3.5 Flue Gas Desulfurization System Analysis

Figure 2.3.4 shows the process flow diagram for the existing Flue Gas Desulfurization System. The stream numbers in Figure 2.3.4 also correspond to stream numbers shown in Figure 2.3.1 and Table 2.3.1. The flue gas leaving the ID fan (Stream 7) is delivered to the Absorber, which consists of a tray followed by a two-stage spray system. The incoming gas is saturated as it passes through the scrubbing slurry contained on the tray and through the two spray levels. The active component of the scrubbing slurry is calcium oxide (Stream 8a), which reacts with sulfur dioxide to form calcium bisulfite (Stream 9). The scrubbing slurry is circulated from the reagent feed tank that forms the base of the scrubber to the spray levels. The solids loading in the scrubbing slurry controls the blowdown from the reaction tank to byproduct disposal. The flue gas passes through chevron type mist eliminators that remove entrained liquid before exiting the scrubber (Stream 10). The water utilized in spray washing the mist eliminators also serves as make-up (Stream 8b).

Table 2.3.4 identifies the assumptions that were made in predicting the FGD performance. Table 2.3.5 shows the gas constituents at the existing Absorber inlet and outlet locations. Results show a CO_2/SO_2 mole ratio of 63 and an SO_2 removal efficiency of 94.8%, corresponding to a value of 104 vppm at the outlet of the absorber.

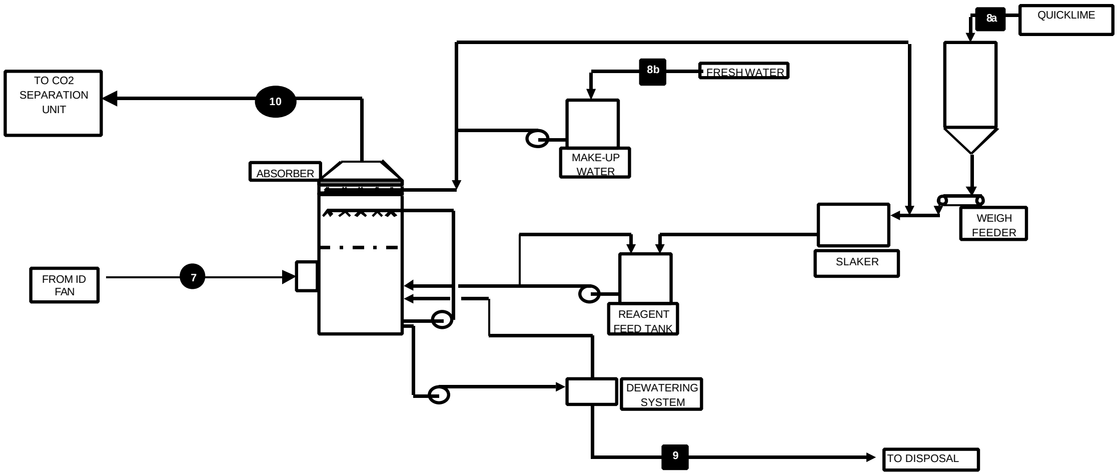


Figure 2.3. 4: Base Case Flue Gas Desulfurization System Process Flow Diagram

Table 2.3. 4: FGD Assumptions

Base Case		
Quantity	Unit	Existing Absorber
Ca/S)	Mol Ratio	1.04
Solids	Wt.%	20
CaO	Wt.%	90
MgO	Wt.%	5
Inerts	Wt.%	5
Bypass Leakage	Wt.%	2.5
Liquid/Gas (L/G) Ratio	gpm/1000 acfm	55
SO_2 Removal Efficiency		
APC	%	94.8
Absorber	%	97.2

Table 2.3.5: FGD Performance

Species	Base Case					
	Existing Absorber Inlet			Existing Absorber Outlet		
	Mol/hr	Vol.%	Unit	Mol/hr	Vol.%	Unit
O_2	4,469	3.14	Vol.%	4,461	2.91	Vol.%
N_2	105,018	73.74	Vol.%	105,018	68.44	Vol.%
H_2O	12,863	9.03	Vol.%	24,228	15.79	Vol.%
CO_2	19,743	13.86	Vol.%	19,720	12.85	Vol.%
SO_2	315	2,212	vppm	16	104	vppm
SO_2 Removal Efficiency, %					94.9	
CO_2/SO_2 Mole Ratio		63				

3.0 CO_2 REMOVAL SYSTEMS

The following three basic process options were studied:

- **Concept A**: Coal combustion in air, followed by CO_2 separation with Kerr-McGee/ABB Lummus Global's commercial MEA-based absorption/stripping process
- **Concept B**: Coal combustion with O_2 firing and flue gas recycle (oxy-fuel firing)
- **Concept C**: Coal combustion in air with oxygen removal and CO_2 separation

The Capture of CO_2 from the flue gas for Concept A is accomplished using the Kerr-McGeee / ABB Lummus Global oxygen inhibited MEA technology. Only Kerr-McGee/ ABB Lummus oxygen inhibited amine was studied. It is the most commercially proven process evaluated in this study.

Concept B utilizes oxygen firing with the recycle of flue gas to bring the firing and heat transfer characteristics of the boiler back close to original design. The oxygen firing is used to produce a flue gas stream with high enough CO_2 content such that simple compression, refrigeration, and rectification can produce a suitable CO_2 liquid product. Optimizations in various areas such as oxygen purity, air leakage into the boiler and associated equipment operating below atmospheric pressure, and flue gas recycle ratio have been made. On the liquefaction side, an optimization of the flow scheme to remove the inerts while maintaining CO_2 recovery at the lowest energy consumption was made.

For Concept C, CO_2 recovery from the flue gas is accomplished by using generic amines (i.e. MEA, MDEA). However, if more efficient and cost effective, proprietary amines may be used. The difference between the amines used in this case and in Concept A is that the amines are not oxygen resistant. Hence, the oxygen is converted to CO_2 by combustion with methane over a de-oxy catalyst upstream of the amine contactor.

Although triethanolamine (TEA) was first proposed for use for Concept C, due to its relatively low energy requirement for solvent regeneration, the literature survey conducted earlier in the present study, Bozzuto, et al. (2000), revealed that the TEA had a very low recovery of CO_2 when operating at near atmospheric pressure. Operating at higher pressures to improve recoveries is not economically feasible as the carbon dioxide concentration in the flue gas is less than 13% and compressing the entire flue gas stream to 200 psig would require approximately 250,000 hp. Therefore, only generic amines and proprietary amines that absorb efficiently at atmospheric pressure were evaluated for this case. Specifically, the amines evaluated were monoethanolamine/ methyldiethanolamime (MEA /MDEA) blend, BASF activated MDEA, and Mitsubishi Heavy Industries amines (Iijima, et al., 1998; Iijima, 1998). Since these amines are required to be oxygen resistant, de-oxygenation via catalytic combustion is required upstream of the amine unit. Neither the de-oxy catalyst nor any of these amines have been tested in coal fired power plant flue gas streams.

3.1 CO_2 Removal Systems Design Basis

3.1.1 Site Data

Listed below is the summary of the bases used for this design:

- Plant is located in Connesville, Ohio, elevation 744 feet.
- Atmospheric pressure is 29.92 inches of Hg.
- Ambient temperature for air cooler design is 80 °F.
- Wet bulb temperature for cooling tower design is 75 °F.
- Electric power may be available from the existing facilities. Auxiliary power, 7-8% of the gross, is provided through auxiliary transformers at 4160-volt bus and is reduced down to 480 volts.

- Flue gas is high in halides (chlorides and fluorides) as well as SO_3 so that stainless steel 316L can be used as the material of construction for the flue gas cooling system.

- Pressure of liquid product CO_2 is 2000 psig.

3.1.2 Battery Limit Definition

The CO_2 recovery and liquefaction sections are located approximately 1200 feet north of the Unit #5 stack (Figure 3.1.1). The new cooling tower is at a different location, which is about 1900 feet southwest of the CO_2 recovery and liquefaction equipment. The CO_2 recovery and liquefaction equipment, for Concepts A and C, actually receives cooling water from both the existing plants cooling system and the new cooling tower. This is due to the fact that the existing plant cooling water capacity will be partially available as a result of the amine system reboiler’s consumption of steam for solvent regeneration. This steam was previously exhausted from the existing low-pressure turbine and condensed in the existing power plant surface condensers.

The new cooling tower section receives frequent deliveries of sodium hypochlorite. It is also periodically attended by cooling tower chemical service such as Betz or Nalco. They bring with them the required dispersant, pH control, and corrosion inhibitors. However, their visits do not totally relieve the power plant personnel from taking samples themselves at least once per shift. The blowdown from the cooling tower must be treated with sodium bisulfite to dechlorinate it. Sodium bisulfite will be unloaded from drums into the injection package. Concept A, which makes use of an existing cooling water make-up stream must have its blowdown filtered to reduce the suspended solids to within acceptable limits. Backwash from the sand filters will go back to the existing cooling tower blowdown disposal system. Concepts B and C require significant amounts of cooling tower make up water. Thus, river water taken from existing pumps will be sent to a new clarifier. It is expected that this clarifier will eliminate suspended solids to the extent that there will be no need for sand filters on the cooling tower blowdown. Blowdown from the clarifier will be sent to an existing clarifier blow down system.

The new CO_2 recovery and liquefaction section comprises another section of the power plant. It has its own control room and motor control center (MCC). In addition to the flue gas, which serves as the feed to the unit, it must also receive the required utilities and chemicals. Caustic, if available from existing facilities, can be used to maintain levels in this facilities day tanks. Otherwise it can be offloaded from trucks into the day tanks. Diatomaceous earth will be off loaded on skids. The spent diatomaceous earth leaves the plant in drums. Reclaimer effluent will be collected in a tank truck parked at one end of the unit. Potable water for eyewashes and cooling tower make-up water for hose down will be routed along side the feed gas duct and 180 psig steam. Corrosion inhibitor to provide oxygen resistance to the amine in Concept A will be directly from drums into an injection package.

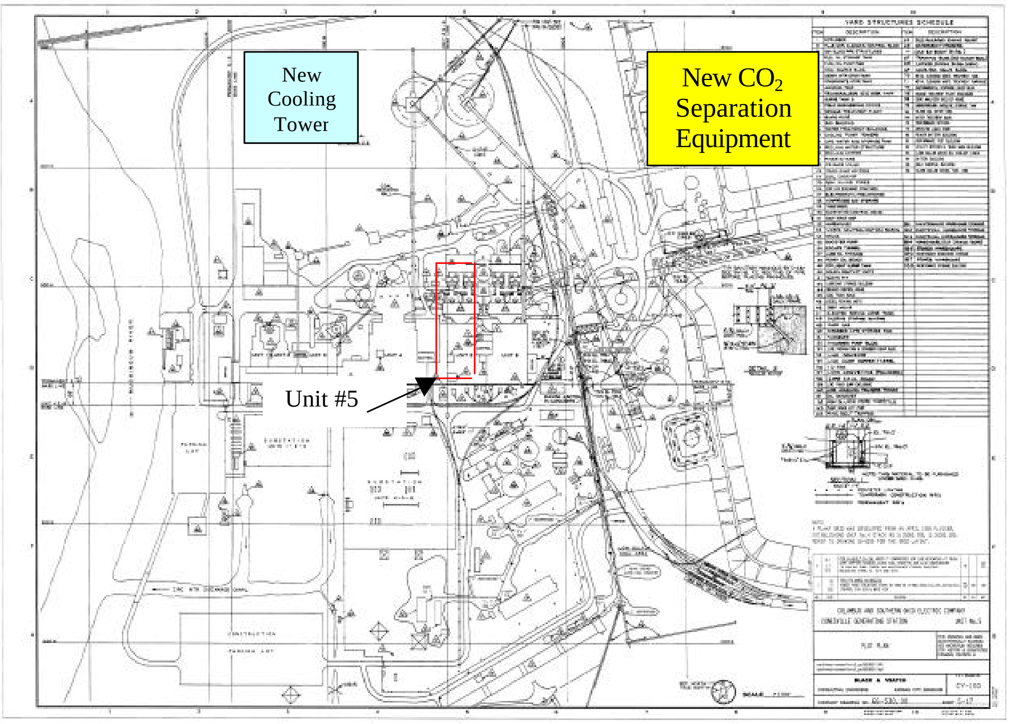


Figure 3.1. 1: AEP Conesville, Ohio, Electric Power Generating Station Site

The design of the CO_2 recovery and liquefaction sections has been based on the flue gas data leaving the Flue Gas Desulfurization System, shown in Table 3.1.1.

Table 3.1. 1: Flue Gas Analysis

Component	Concept A	Concept B	Concept C
	Mole %	Mole %	Mole %
O2	2.94	2.91	2.94
N2	68.45	5.03	68.45
H2O	15.77	19.01	15.77
CO_2	12.83	73.04	12.83
SO2	<10 ppmv	0.01	106 ppmv
MW	28.61	37.92	28.61
T (°F)	136	144	136
P (psig)	1	0	1

3.1.3 CO_2 Product Specification

The product specification for sequeststration (Orr, 2001) is shown in Table 3.1.2. This composition is what can be obtained from the liquefaction scheme for Concept B. Its hydrate locus was verified to be sufficiently close to that of pure CO_2 which is the criterion used by one of the world's foremost CO_2 sequestration experts, Lynn Orr of Stanford University.

Table 3.1. 2: Specification for Sequestration

Component	Product
	Mole %
O_2	1.32
N_2	1.23
H_2O	0
CO_2	97.29
SO_2	.16
MW	43.68
T (°F)	82
P (psia)	2015

Table 3.1.3 represents the specification required for Enhanced Oil Recovery (EOR). It was compiled by comparing the Dakota Gas specification supplied by AEP and contracts signed through Kinder Morgan CO_2 company and taking the strictest specification of the two.

Table 3.1. 3: Specification for EOR

Component	Specification
O_2	10 weight ppm maximum of the CO_2
N_2	300 ppmv maximum
H_2O	20 ppmv maximum
CO_2	96% minimum
H_2S	20 weight ppm of the CO_2
CO	0.1% maximum
C_2+ Hydrocarbons	2.3 mole % max
Methane	0.7 % maximum
T (°F)	120 F max

A product pressure of 2000 psig was used in the designs that follow.

3.1.4 Chemicals

This section provides data for the chemicals available on site and used by the CO_2 Recovery systems. Conditions for liquid chemicals are specified at grade level.

Table 3.1. 4: Caustic (NaOH)

	Pressure at B.L. psig	Temperature °F
Minimum		
Normal	15	ambient
Maximum		
Mechanical Design	50	125

- Available for reclaiming MEA (Concept A)
- The import and dilution facilities will be used to keep a day tank in the process area at desirable levels.

3.1.5 Utilities

The data below applies to both Concepts A and C except where noted. For cooling water, new utility systems will have to be provided and are included in the cost estimate. For steam, obviously the export of 195 psia steam to an external process represents a change from the current operation. It was decided to use extraction steam from the turbine rather than build additional facilities. The reason for this decision comes from plant efficiency, investment cost, and plot considerations. Conditions for all liquid utilities are given at grade level.

Steam:

Low Pressure Steam (LP) (Concepts A and C)
Used mainly for process heating

	Pressure at B.L. psig	Temperature °F
Minimum (for process design)		700
Normal	180	716
Maximum	200	
Mechanical Design	250 /FV	775

Water:

Cooling Water (Concepts A, B, and C)
Source: New Cooling Tower

CW Supply:	Pressure at B.L. psig	Temperature °F
Minimum		
Normal	50	85
Maximum		
Mechanical Design	100	150

CW Return:	Pressure at B.L. psig	Temperature °F
Minimum		
Normal	30	103
Maximum		
Mechanical Design	100	175

Surface Condensate (Concept C process cooling, Concept A and C amine make-up**)**

	Pressure at B.L. psig	Temperature °F
Minimum		
Normal	300	109
Maximum		
Mechanical Design	350	200

Raw Water (Fresh Water)
Distributed for general use at hose stations (Concepts A and C). The source of this is a clarifier for cooling tower make-up. The capacity of this clarifier is sufficient for make up for Concept A, but not B or C. Its quality is as follows:

Components	Unit	Specifications
Si	ppm.	22
Iron (as Fe)	ppm.	0.18
Copper (as Cu)	ppm	0.05

Suspended Solids	ppm	15
Chlorine	ppm	100-180
Alkalinity	ppm	100
Na	ppm	100

For Concepts B and C water from the Muskingum River must be sent to a new clarifier. The specifications for this water are as follows:

	Unit	Specification
Suspended Solids	ppm	10 to many hundreds
Turbidity	ppm	5 to 300

Potable Water (Concepts A and C)

	Pressure at B.L. Psig	Temperature °F
Minimum (for process design)		
Normal	100	ambient
Maximum		
Mechanical Design	150	150

- Comes from public network
- For safety showers (Concepts A and C)

Air:

Plant Air (Concepts A, B, C)

	Pressure at B.L. psig	Temperature °F
Minimum (for process design)		
Normal	115	100
Maximum		
Mechanical Design	175	150

Dew point (at normal supply pressure - 40°C)

Instrument Air (Concepts A, B, C)

	Pressure at B.L. psig	Temperature °F
Minimum (for process design)		
Normal	115	100
Maximum		
Mechanical Design	175	150

Dew point (at normal supply pressure - 40°C)
Dust, oil and grease free

Fuel Gas:

LP Fuel Gas (Concepts A, B, C)

	Pressure at OSBL psig	Temperature °F
Minimum		
Normal	50	Ambient
Maximum		
Mechanical Design	100	150

Power Supply:

All of the required power (100%) for the CO_2 Recovery Unit will be provided by AEP either from the local supply or from the Ohio Grid.

(Concepts A, B, C)

Service	Voltage	Phase
Auxiliary plant power system	4160	3-phase
Large Motors	4160	3-phase
Small Motors	480	3-phase
Instruments, Lighting etc	480 / 230	3/1-phase

3.1.6 Equipment Design

Equipment Numbering:

Equipment numbering will be based on Lummus standards.

Sparing Philosophy:

A sparing philosophy was employed to reduce investment. Therefore, spares were spread out upon multiple trains or provide 50% or 33% sized spares when possible. Small pumps (e.g. metering pumps) and equipment related to protection of major equipment (e.g. lube oil pumps) may be spared.

Mechanical Design Conditions:

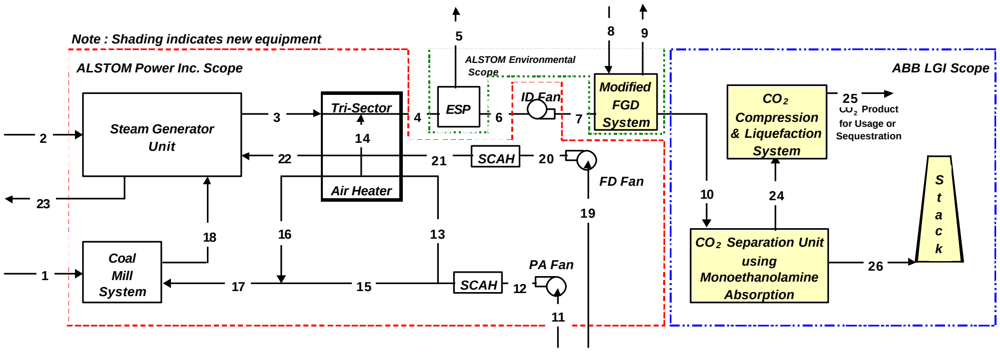
ABBLGI practices will be followed (page 24, T3.1.TP07.402 in “Relief and Flare Systems” manual).

3.2 Concept A: CO_2 Separation with Monoethanolamine Absorption

Concept A is defined as a system designed for the removal and recovery of CO_2 from the boiler flue gas using an amine scrubbing system. The amine system selected was the Kerr-McGeee / ABB Lummus Global oxygen inhibited MEA technology.

3.2.1 Overall System Description and Material and Energy Balance

A simplified process flow diagram for the modified unit is shown in Figure 3.2.1. The operation and performance of the existing Boiler and ESP systems are identical to the Base Case and are not affected by the addition of the MEA based CO_2 removal system. The FGD system is modified with the addition of a secondary absorber to reduce the SO_2 content to about 10 ppmv (See Section 3.2.4). The overall material and energy balance for the system shown in Figure 3.2.1 is shown in Table 3.2.1. The flue gases leaving the modified FGD system are ducted to the new MEA system where ~94 percent of the CO_2 is removed, compressed, and liquefied for usage or sequestration. The remaining flue gases leaving the new MEA system, consisting of primarily oxygen, nitrogen, water vapor and a relatively small amount of sulfur dioxide and carbon dioxide, is discharged to the atmosphere.



Material Flow Stream Identification

1	Raw Coal to Pulverizers
2	Air Infiltration Stream
3	Fluegas from Economizer to Air Heater
4	Fluegas Leaving Air Heater to ESP
5	Flyash Leaving ESP
6	Fluegas Leaving ESP to Induced Draft Fan
7	Fluegas to Fluegas De-Sulfurization System
8	Lime feed to FGD System
9	FGD System Solids to Disposal
10	Fluegas to MEA System
11	Air to Primary Air Fan
12	Primary Air to Steam Coil Air Heater
13	Primary Air to Air Heater
14	Air Heater Leakage Air Stream
15	Tempering Air to Pulverizers
16	Hot Primary Air to Pulverizers
17	Mixed Primary Air to Pulverizers
18	Pulverized Coal and Air to Furnace
19	Secondary Air to Forced Draft Fan
20	Secondary Air to Steam Coil Air Heater
21	Secondary Air to Air Heater
22	Heated Secondary Air to Furnace
23	Bottom Ash from Furnace
24	Separated CO_2 to Compressor
25	Compressed CO_2 Product
26	CO_2 Depleted Fluegas to Stack

Figure 3.2. 1: Simplified Gas Side Process Flow Diagram for CO_2 Separation by Monoethanolamine Absorption (Concept A)

Table 3.2. 1: Gas Side Material and Material Energy Balance (Concept A)

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11	12	13
O_2	(Lbm/hr)	26586	42147	101097	144817		144817	144817	5628		144566	203237	203237	112918
N_2	"	4868	139626	2797385	2942220		2942220	2942220			2942220	673283	673283	374075
H_2O	"	37820	2357	228849	231294		231294	231294	258954	48324	441924	11365	11365	6314
CO_2	"			867210	867210		867210	867210			866102			
SO_2	"			20202	20202		20202	20202			87			
H_2	"	16102												
CH_4	"													
Carbon	"	236655												
Sulfur	"	10110												
Ca	"								13087					
Mg	"								613					
MgO	"									509				
$MgSO_3$	"									1251				
$MgSO_4$	"									76				
$CaSO_3$	"									34395				
$CaSO_4$	"									2051				
$CaCO_3$	"									2520				
Ash / Inerts	"	42313		33851	33851	33851			1017	1017				
		Raw Coal	Leakage Air	Fluegas to AH	Fluegas to ESP	Flyash	Fluegas to ID Fan	Fluegas to FGD	Lime Slurry	FGD Disposal	Fgas to CO2 Sep	Pri Air to PA Fan	PA from PA Fan	Pri Air to AH
Total Gas	(Lbm/hr)		184130	4014743	4205743		4205743	4205743			4394900	887885	887885	493308
Total Solids	"	**374455**		33851	33851	33851			14717	41819				
Total Flow	"	374455	184130	4048594	4239594	33851	4205743	4205743	279300	90143	4394900	887885	887885	493308
Temperature	(Deg F)	**80**	80	**706**	**311**	311	311	325	80	136	136	**80**	92	92
Pressure	(Psia)	14.7	14.7	14.6	14.3	14.7	14.2	15.0	14.7	14.7	14.7	14.7	15.6	15.6
$h_{sensible}$	(Btu/lbm)	0.000	0.000	161.831	57.924	57.750	57.924	61.384	0.000	14.116	14.543	0.000	2.899	2.899
Chemical	(10^6 Btu/hr)	4228.715												
Sensible	(10^6 Btu/hr)	0.000	0.000	655.007	245.567	1.955	243.612	258.166	0.000	3.314	63.916	0.000	2.574	1.430
Latent	(10^6 Btu/hr)	**0.000**	2.475	240.291	242.858	0.000	242.858	242.858	**0.000**	0.000	464.020	11.933	11.933	6.630
Total Energy$^{(1)}$	(10^6 Btu/hr)	4228.715	2.475	895.298	488.425	1.955	486.470	501.024	0.000	3.314	527.936	11.933	14.507	8.060

Constituent	(Units)	14	15	16	17	18	19	20	21	22	23	24	25	26
O_2	(Lbm/hr)	43720	90319	66680	156999	183585	641283	641283	641283	643801				
N_2	"	144835	299208	220899	520107	524975	2124443	2124443	2124443	2132785				
H_2O	"	2445	5051	3729	8779	46599	35860	35860	35860	36001				
CO_2	"													
SO_2	"													
H_2	"					16102								
CH_4	"													
Carbon	"					236655								
Sulfur	"					10110								
Ca	"													
Mg	"													
MgO	"													
$MgSO_3$	"													
$MgSO_4$	"													
$CaSO_3$	"													
$CaSO_4$	"													
$CaCO_3$	"													
Ash / Inerts	"					42313					8463			
		Air Htr Lkg Air	Tempering Air	Hot Pri Air	Mixed Pri Air	Coal-Pri Air Mix	Sec Air to FD	Sec Air to SCAH	Sec Air to AH	Hot Sec Air	Bottom Ash	CO2 to Comp	CO2 Product	Vent Stream
Total Gas	(Lbm/hr)	191000	**394577**	291308	685885		2801587	2801587	2801587	2812587				
Total Solids	"										8463			
Total Flow	"	191000	394577	291308	685885	1060340	2801587	2801587	2801587	2812587	8463			
Temperature	(Deg F)	92	92	**666**	**339**		**80**	86.4	**86.4**	**616**	**2000**			
Pressure	(Psia)	15.6	15.6	15.6	15.6	15.0	14.7	15.2	15.1	14.9	14.7			
$h_{sensible}$	(Btu/lbm)	2.899	2.899	145.249	63.358		0.000	1.549	1.549	132.582	480.000			
Chemical	(10^6 Btu/hr)					4228.715								
Sensible	(10^6 Btu/hr)	0.554	1.144	42.312	43.456		0.000	4.341	4.341	372.898	4.062			
Latent	(10^6 Btu/hr)	2.567	5.303	3.915	9.218		37.653	37.653	37.653	37.801	0.000			
Total Energy$^{(1)}$	(10^6 Btu/hr)	3.121	6.447	46.227	52.674	4281.389	37.653	41.994	41.994	410.699	4.062			

Notes:

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

Boiler efficiency is calculated to be 88.13 percent (HHV Basis), the same as for the Base Case. The net plant heat rate is increased significantly to 16,626 Btu/kwhr (HHV Basis) for this option as shown in Table 3.2.2 which also includes the Base Case for comparison. The plant thermal efficiency for Concept A (20.53%, HHV basis) is about 59 percent of the Base Case value (35.01%), indicating an energy penalty of ~41%. Auxiliary power is increased to 76,007 kw and the net plant output is reduced to 254,414 kw. Carbon dioxide emissions are 33,084 lbm/hr or about 0.13 lbm/kwhr.

Table 3.2. 2: Overall Plant Performance Summary (Concept A)

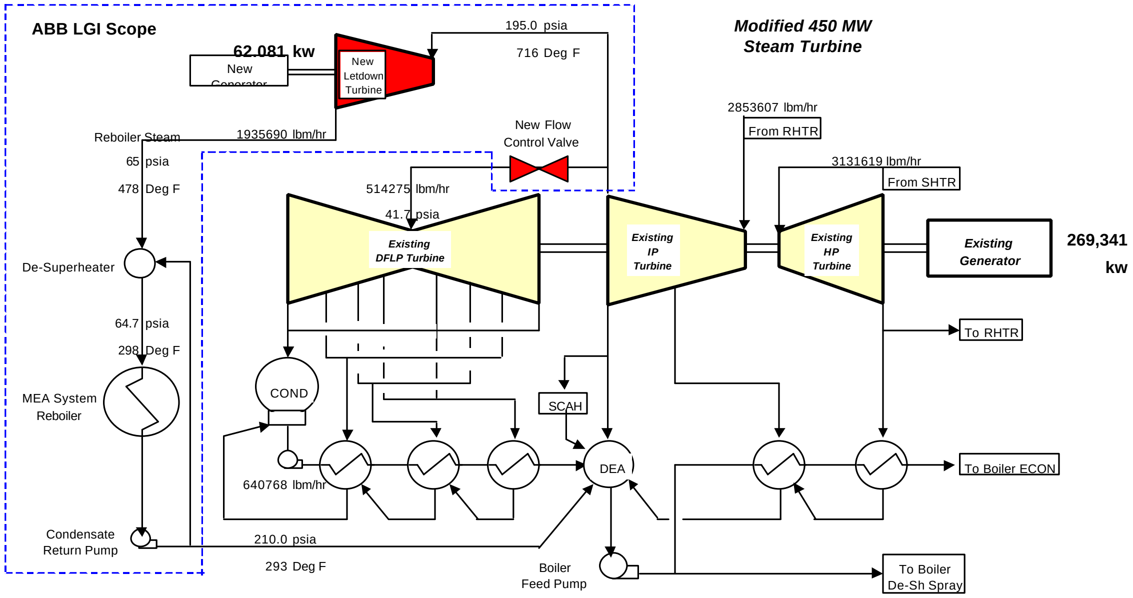
	(units)	Original Plant (Base)	Concept A MEA
Fuel Paramaters			
Coal Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4228.7
Natural Gas Heat Input (HHV)	(10^6 Btu/hr)	---	17.7
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4246.4
Steam Cycle Paramaters			
Existing Steam Turbine Generator Output	(kW)	463478	269341
CO2 Removal System Turbine Generator Output	(kW)	0	62081
Total Turbine Generator Output	(kW)	463478	331422
Total Auxiliary Power	(kW)	29700	76007
Net Plant Output	(kW)	433778	255414
Overall Plant Performance Paramaters			
Net Plant Efficiency (HHV)	(fraction)	0.3501	0.2053
Net Plant Efficiency (LHV)	(fraction)	0.3666	0.2150
Normalized Efficiency (HHV; Relative to Base Case)	(fraction)	1.0000	0.5864
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9749	16626
Net Plant Heat Rate (LHV)	(Btu/kwhr)	9309	15872
Overall Plant CO_2 Emissions			
Carbon Dioxide Emissions	(lbm/hr)	866102	33084
Specific Carbon Dioxide Emissions	(lbm/kwhr)	1.997	0.130
Normalized Specific CO2 Emissions (Relative to Base Case)	(fraction)	1.000	0.065
Avoided Carbon Dioxide Emissions (as compared to Base)	(lbm/kwhr)	---	1.867

3.2.2 Boiler Analysis

The boiler performance in Concept A is identical to that of the Base Case described previously in Section 2.3 and will not be repeated here.

3.2.3 Steam Cycle Modifications and Performance

The steam cycle system for Concept A is modified as shown in Figure 3.2.2 while Figure 3.2.3 shows the associated Mollier diagram. About 79 percent of the IP turbine exhaust is extracted from the IP/LP crossover pipe. This steam is expanded to about 65 psia through a new steam turbine generating 62,081 kw. The exhaust from the new turbine, at about 478 F, is de-superheated and then provides the energy requirement for the solvent regeneration done in the reboiler's of the MEA CO_2 removal system. The condensate from the reboiler's is pumped to the Deaerator. The modified existing steam cycle system produces 269,341 kw. The total output from both generators is 331,422 kw. This represents a gross output reduction of 132,056 kw (about 28.5%) as compared to the Base Case. A yet unsolved uncertainty is whether the existing low-pressure turbine can be operated over the required load range with the such a small fraction of the original design steam mass flow rate. The answer to this question was beyond the scope of the current project.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Power Output (Existing and New Turbines)	1151	Boiler Heat Input	3707
Steam Coil Air Heater Output	0	Condensate Pump	0
MEA System Reboiler Duty Output	1953	Total Energy Input	3708
Condenser Loss	603		
Total Energy Output	3708	In - Out	0

Figure 3.2. 2: Modified Steam Cycle Diagram and Performance (Concept A)

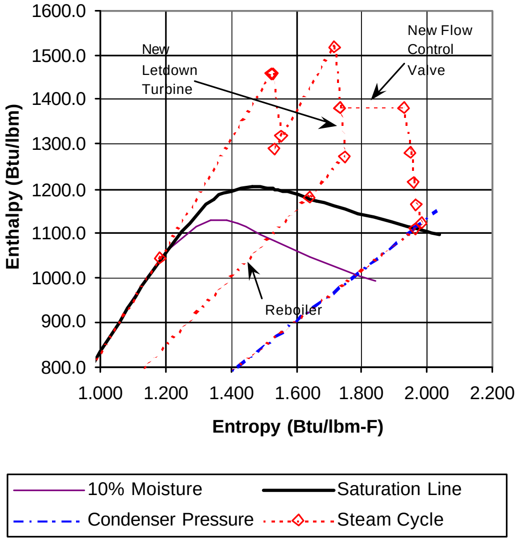


Figure 3.2. 3: Modified Steam Cycle Mollier Diagram (Concept A)

3.2.4 Flue Gas Desulfurization System Analysis

The FGD system for Concept A is modified with the addition of a secondary absorber to reduce the SO_2 content to 10 vppm or less as required by the amine system downstream. The principle of operation is as briefly described previously in Section 2.3.5. In this case, however, the flue gas leaving the existing FGD system's absorber is supplied to the new secondary absorber and the flue gas leaving the secondary absorber provides the feed stream for the amine CO_2 absorption system, as discussed later in Section 3.2.5. Additional piping and ductwork is required as is shown in Figure 3.2.4 process flow diagram.

Table 3.2.3 identifies the assumptions that were made in predicting the FGD performance.

Table 3.2.4 shows the gas constituents at the existing Absorber inlet and secondary Absorber outlet. Results show a CO_2/SO_2 mole ratio of 63 and an SO_2 removal efficiency of 99.7%, corresponding to a value of 6.5 vppm at the outlet of the secondary absorber.

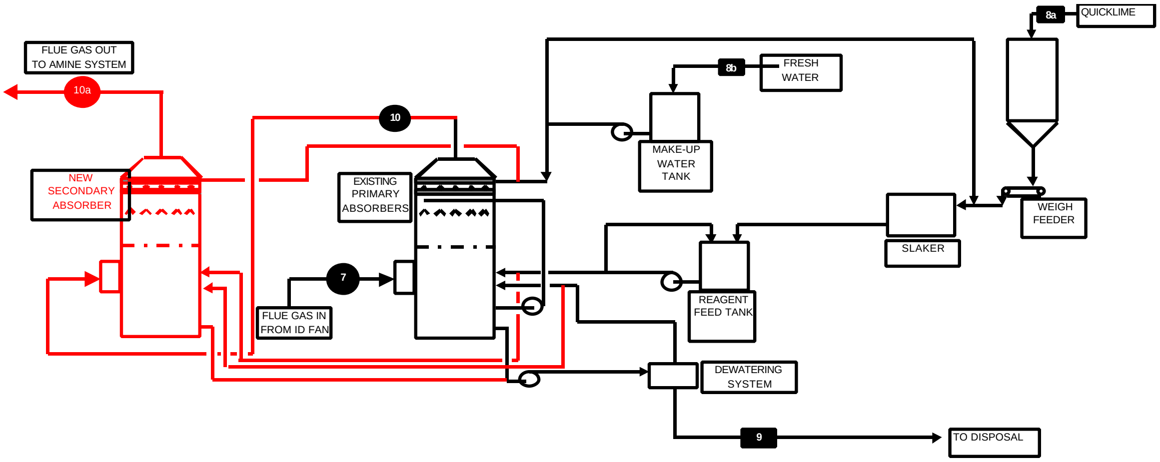


Figure 3.2. 4: Modified FGD System Process Flow Diagram (Concept A)

Table 3.2. 3: FGD Assumptions

Concept A (MEA)			
Quantity	Unit	Existing Absorber	Secondary Absorber
Ca/S)	Mol Ratio	1.04	1.04
Solids	Wt.%	20	20
CaO	Wt.%	90	90
MgO	Wt.%	5	5
Inerts	Wt.%	5	5
By-pass Leakage	Wt.%	2.5	0
Liquid/Gas (L/G) Ratio	gpm/1000 acfm	75	45
SO_2 Removal Efficiency			
APC	%	94.8	93.0
Absorber	%	97.2	93.0

Table 3.2.4: FGD Performance

	Concept A (MEA)					
	Existing Absorber Inlet			Secondary Absorber Outlet		
Species	Mol/hr	Vol.%	Unit	Mol/hr	Vol.%	Unit
O_2	4,469	3.14	Vol.%	4,461	2.90	Vol.%
N_2	105,018	73.74	Vol.%	105,018	68.30	Vol.%
H_2O	12,863	9.03	Vol.%	24,555	15.97	Vol.%
CO_2	19,743	13.86	Vol.%	19,718	12.82	Vol.%
SO_2	315	2,212	vppm	1	6.50	vppm
SO_2 Removal Efficiency, %					99.7	
CO_2/SO_2 Mole Ratio		63				

3.2.5 Carbon Dioxide Separation and Compression System

The Kerr-McGee/ ABB Lummus amine technology is used for the Concept A CO_2 removal system. This system is the most proven of the three processes analyzed in this study. An important feature of this CO_2 recovery technology is its flexibility to operate with boilers or co-generation systems that fire fuels ranging from natural gas to high-sulfur coal and coke. The process tolerates oxygen in the flue gas as well as limited amount of sulfur dioxide. Low corrosion rates and minimal loss of the circulating solvent used to absorb CO_2 ensure economical and reliable operation.

Kerr-McGee started up their 800 TPD CO_2 recovery unit in 1978. The Trona unit has been fed flue gases from boilers fired with natural gas, coal and coke. During its first several years of operation, the installation was improved in terms of reliability and cost effectiveness. As of January 1992, three units have been licensed using this technology. They are:

1.) Applied Energy Systems, Poteau, Oklahoma. This 300 MW coal-fired co-generation plant incorporates a 200 TPD food-grade liquid CO_2 unit as the steam host. Startup was completed in January 1991. Lummus Crest was responsible for engineering, procurement, and construction of the entire facility, including the power plant. Lummus Crest made several design improvements to the AES facility.

2.) Soda Ash Botswana, Pty. Ltd., Sue Pan, Botswana. This soda ash facility, incorporating a 300 TPD CO_2 unit, started up in March 1991.

3.) AES Corporation is building a second food grade CO_2 plant in Warrior run, Pennsylvania using the Kerr-McGee / Lummus technology. This is a 150 short tons/day liquid CO_2 plant, using flue gas from coal fired circulating fluidized bed (CFB) boilers, was commissioned in 1999.

In late 1990, Kerr-McGee and Lummus Global concluded a joint licensing agreement whereby Lummus gained worldwide exclusive marketing rights to Kerr-McGee’s CO_2 recovery technology and became responsible for marketing and basic engineering. Kerr-McGee maintains a continuing role in technology transfer, process improvement, quality control of new designs, operator training, and licensing. In addition to providing an experience list for the amine part of the process, these plants also contain CO_2 compression and liquefaction facilities. These plants compress the CO_2 to a slightly higher pressure and utilize ammonia refrigeration. Many of these plants make food grade CO_2.

3.2.5.1 Introduction

The CO_2 Recovery Unit for Concept A is comprised of the following sections:

- Flue Gas Pretreatment
- Absorption
- Stripping
- CO_2 Compression and Liquefaction
- Drying

The flue gas pretreatment section cools and conditions the flue gas, which is then fed to the CO_2 Absorber (or CO_2 contactor as it is often called). In the Absorber, CO_2 is removed from the gas by contacting it, in countercurrent fashion, with monoethanolamine (MEA). The recovered CO_2 is then stripped off in the Stripper (or Regenerator) from where the lean solvent is recycled back to the Absorber. Solvent regeneration for Concept A requires about 4.7×10^6 Btu/Ton CO_2. The overhead vapor from the Stripper is cooled to condense most of the water vapor. The condensate is used as reflux in the Stripper and the wet CO_2 stream is fed forward to the CO_2 Compression and Liquefaction System. Here the CO_2 product is compressed and dried so it can be pumped to its final destination. No specific destination has been chosen for the product pipeline. It has been assumed to end at the battery limit for costing purposes.

A brief description of the processing scheme for Concept A is given in the following paragraphs. Description of the package units is indicative only and may vary for the chosen supplier of the package unit

3.2.5.2 Process Description

This section refers to the following process flow diagrams, which are shown in Section 3.2.5.3:

PROCESS FLOW DIAGRAMS - CONCEPT A:
- Figure 3.2.5:Drawing D 09484-01001R-0: Flue Gas Cooling and CO_2 Absorption
- Figure 3.2.6:Drawing D 09484-01002R-0: Solvent Stripping
- Figure 3.2.7:Drawing D 09484-01003R-0: CO_2 Compression and Liquefaction

The designs include several process trains. Only one train is shown. The note section of the PFD tells how many trains are included in the complete system. To avoid confusion, suffixes have been used to indicate parallel equipment. These are mainly for spared pumps and drier vessels in parallel. Even if there are several trains, only one drawing (typical) has been prepared to represent all of the trains. On these drawings, flow splits to the other parallel trains have been shown. Similarly, flows coming from other parallel trains and converging to a single common stream have also been shown.

A note about stream numbering convention is also necessary. The stream numbers have not been tagged with “A”, “B”, etc. to indicate which train they belong to. Instead, the flow rate given in the material balance for each stream is the actual flow rate for the stream within the train. The combined flow from all of the trains leaving a process step shows the total flow going to the next process step. As an example, stream 8 (Drawing D 09484-01001R-0) is the Rich Amine stream leaving one train of the absorber process step, and comprises 1/5 of the total rich amine. Stream 9A is the total rich amine going to the Solvent Stripping process step. Stream 9A appears on both the absorber and solvent stripper PFD’s. After the rich amine flow sheet continuation block, the stream splits 9 ways for the 9 stripping trains. Then stream 9 continues for processing on the solvent stripper PFD (Drawing D 09484-01002R-0), with 1/9 of the flow entering the rich-lean solvent exchanger (EA-2205). The numbering practice for Concept C is the same as for Concept A.

Flue Gas Pretreatment:
The pressure profile of the CO_2 capture equipment is contained in the material balance. Since the flue gas pretreatment equipment flow scheme includes a blower, the pressure profile of the existing power generation equipment does not change from today’s operation. To force the flue gas from the secondary FGD through the CO_2

Absorber, the pressure of the flue gas after sulfur removal is boosted to 1.5 psig by a motor driven fan. As the power consumption of the fan is considerable, the duct size must be chosen not to add excessive pressure drop for the 1200 feet it takes to get to the absorbers. The blower will run at constant speed. Each blower, provided as part of the boiler flue gas conditioning equipment, is equipped with its own suction and a discharge damper operated pneumatically. The suction damper controls the suction pressure to adjust for the flow variation resulting from the power plant performance. The suction pressure control will avoid any surges to blower. The discharge damper is an isolation damper.

Direct Contact Cooling (Refer to Drawing No. D 09484-01001R-0):

The Direct Contact flue gas Cooler (DCC) is a packed column where the hot flue gas flowing up is brought into an intimate contact with cold water which is fed to the top of the bed and flows down the tower. Physically, DA-2101 and DA-2102 have been combined into a single, albeit compartmentalized tower. DA-2101 is the lower one and is designed to support DA-2102 so that the top head of DA-2101 is the bottom head of DA-102. Effectively, this dividing head acts as a chimney tray with a number of upward extending chimneys which provide passages for the flue gas to flow directly from the DCC into the Absorber.

Theoretically, a direct contact cooler is capable of cooling the gas to a very close approach in a short bed. When the hot gas enters the DCC, the gas contains water but is highly superheated. At the bottom end of the bed, the gas is quickly cooled to a temperature known as the “Adiabatic Saturation Temperature” (AST). This is the temperature the gas reaches when some of its own heat content has been used to vaporize just the exact amount of water to saturate the gas.

Up to the point when the AST is reached, the mass flow of the gas stream increases due to evaporation of water. At the AST, water vapor contained in the gas begins to condense. And, as the gas travels up the column and cools further, more and more water is condensed. This internal refluxing increases the V/L traffic at the bottom end of the bed significantly beyond the external flows and must be considered in the hydraulic design.

The water stream that leaves the bottom of the DCC contains the water fed to the top as well as any water that has condensed out of the flue gas. The condensed water may be somewhat corrosive due to sulfur and nitrogen oxides that may be present in the flue gas. Therefore, instead of using the condensate in the process, it will be blown down from the system. For the DCC to be effective, the temperature of the leaving water must always be lower than the AST.

Most of the water leaving the bottom of the DCC is circulated back to the top of the direct contact cooler by DCC Water Pump GA-2102 A/B. However, before sending it back to the column the water stream is first filtered in DCC Water Filter FD-2101 and then cooled in DCC Water Cooler EA-2101 against the water from the new cooling tower. Temperature of the cooled water is controlled by a cascade loop which maintains a constant flue gas exit temperature (Absorber feed temperature). Because of the relatively low cooling water temperature at the plant, the circulating water is cooled down to 95 ^{o}F which, in turn, easily cools the gas down to 115 ^{o}F.

Filtration is necessary to remove any particulate matter that may enter the DCC in the flue gas. The blowdown is taken out after the filter but before the cooler and mixed into the return water of cooler EA-2101. This way the cooler does not have to handle the extra duty that would otherwise be imposed by the blowdown stream.

Absorption:

CO_2 Absorber DA-2102 (Refer to Drawing No. D 09484-01001R-0):

From the DCC the cooled flue gas enters the bottom of the CO_2 Absorber and flows up the tower countercurrent to a stream of 20-wt.% monoethanolamine (MEA) solution. The lean MEA solution (LAM) enters the top of the column and heats up gradually as more and more CO_2 is absorbed. By the time the stream leaves the bottom of the tower it has gained approximately 28 ^{o}F. The tower has been designed to remove 94% of the CO_2 in the incoming gas. The CO_2 loading in LAM is 0.215 mol CO_2 / mol MEA while the loading of the rich amine leaving the bottom is 0.44 mol CO_2 / mol MEA. These values are consistent with the values reported by Rochelle (2000).

To maintain water balance in the process, it is imperative that the temperature of the LAM feed be very close to that of the feed gas stream. Thus, with feed gas temperature fixed at 115 ^{o}F, the temperature of the LAM stream must also be close to 115 °F, preferably within 10 °F. If the feed gas comes in at a higher temperature than the LAM, it brings in excess moisture, which condenses in the Absorber and becomes excess water. Unless this water is purged from the system, the concentration of MEA will decrease and the performance of the system will suffer. If, on the other hand, the gas feed is colder than the LAM, it heats up in the tower and picks up extra moisture that is then carried out of the system by the vent gas. The result is a water deficiency situation because more water is removed than what comes into the system.

For the reasons explained above, it is essential that both the temperature of the flue gas and that of the LAM be accurately controlled. In fact, it is best to control one temperature and adjust the temperature of the other to maintain a fixed temperature difference. The design difference is approximately 10 °F. The LAM temperature was chosen to be the “master” and the gas temperature to be the “slave”.

The rich MEA solvent solution from the bottom of the absorber at 133 °F is heated to 204 °F by heat exchange with lean MEA solvent solution returning from the stripping column. The rich MEA solvent is then fed to the top of the stripping column. The lean MEA solvent solution thus partially cooled to 143 °F is further cooled to 105 °F by exchange with cooling water and fed back to the absorber to complete the circuit.

CO_2 Absorber DA-2102 is a packed tower which contains two beds of structured packing and a third bed, the so called “Wash Zone”, at the very top of the column. There is also a liquid distributor at the top of each of bed. The distributors for the main beds are of high-quality design. There are several reasons for selecting structured packing for this service:

- Very low pressure drop which minimizes fan horsepower
- High contact efficiency / low packing height
- Good tolerance for maldistribution in a large tower
- Smallest possible tower diameter
- Light weight

At the bottom of the tower, there is the equivalent of a chimney tray, which serves as the bottom sump for the Absorber. Instead of being flat like a typical chimney tray, it is a standard dished head with chimneys. The hold-up volume of the bottom sump is sufficient to accept all the liquid held up in the packing both in the CO_2 Absorber and in the Wash Zone. Rich Solvent Pump GA-2103 A/D takes suction from the chimney tray.

Absorber Wash Zone (Refer to Drawing. No. D 09484-01001R-0):

The purpose of the Wash Zone at the top of the tower is to minimize MEA losses both due to mechanical entrainment and also due to evaporation. This is achieved by circulating wash water in this section to scrub most of the MEA from the lean gas exiting the Absorber. The key to minimizing MEA carryover is a mist separator pad between the wash section and the Absorber. But, the demister can not stop losses of gaseous MEA carried in the flue gas. This is accomplished by scrubbing the gas with countercurrent flow of water. Wash Water Pump GA-2101 takes water from the bottom of the wash zone and circulates it back to the top of the bed. Circulation rate has been chosen to irrigate the packing sufficiently for efficient operation.

The key to successful scrubbing is to maintain a low concentration of MEA in the circulating water. The higher the concentration, the higher the vapor pressure of MEA and, consequently, the higher the MEA losses. Therefore, relatively clean water must be fed to the wash zone as make-up while an equal amount of MEA laden water is drawn out. A simple gooseneck seal accomplishes this and maintains a level on the chimney tray at the bottom of the wash section. Overflow goes to the main absorber. Make-up water comes from the overhead system of the Solvent Stripper.

The lean flue gas leaving the wash zone is released to atmosphere. The top of the tower has been designed as stack which is made high enough to ensure proper dispersion of the exiting gas.

Rich/Lean Solvent Exchanger EA-2205 (Refer to Drawing No. D 09484-01002R-0):

The Rich/Lean Solvent Exchanger is a plate type exchanger with rich solution on one side and lean solution on the other. The purpose of the exchanger is to recover as much heat as possible from the hot lean solvent from the bottom of the Solvent Stripper by heating the rich solvent feeding the Solvent Stripper. This reduces the duty of the Solvent Stripper Reboiler. This exchanger is the single most important item in the energy economy of the entire CO_2 Recovery Unit. For this study, 10 °F approach was chosen to maximize the heat recovery. An air cooler (EC-2201) was added on the lean amine stream leaving the Solvent Stripper. This was to reduce the plot space requirement (compared to placing the air cooler downstream of the rich/ lean exchanger) and overall cost of the project. A study was performed to determine that heat transfer via the plate frame type lean/ rich exchanger is relatively cheap which justifies tight temperature approaches with this type of exchanger.

Stripping:

Solvent Stripper DA-2201 (Refer to Drawing No. D 09484-01002R-0):

The solvent Stripper is a packed tower which contains two beds of structured packing and a third bed, so called the "wash zone" at the very top of the column. The purpose of the Solvent Stripper is to separate the CO_2 (contained in the rich solvent) from the bottom stream of the CO_2 Absorber that is feeding the stripper. As the solvent flows down, the bottom hot vapor from the reboiler continues to strip the CO_2 from the solution. The final stripping action occurs in the reboiler. The hot wet vapors from the top of the stripper contain the CO_2, along with water vapor, and solvent vapor. The overhead vapors are cooled by Solvent Stripper CW Condenser (EA-2206) where most of the water and solvent vapors condense. The CO_2 does not condense. The condensed overhead liquid and gaseous CO_2 are separated in a reflux drum (FA-2201). CO_2 flows to the CO_2 purification section on pressure control and the liquid (called reflux) is returned via Solvent Stripper Reflux Pump (GA-2202A/B) to the top bed in the stripper. The top bed of the stripper is a water wash zone designed to limit the amount of solvent (MEA) vapors entering the stripper overhead system.

Solvent Stripper Reboiler EA-2201 (Refer to Drawing No. D 09484-01002R-0):

The steam-heated reboiler is a vertical shell and tube thermosyphon type exchanger using inside coated high flux tubing proprietary of UOP. Circulation of the solvent solution through the reboiler is natural and is driven by gravity and density differences. The reboiler tube side handles the solvent solution and the shell side handles the steam. The energy requirement for the removal of CO_2 is about 2.6 tons of steam per ton of CO_2 for Concept A.

Solvent Reclaimer EA-2203 (Refer to Drawing No. D 09484-01002R-0):

The Solvent Stripper Reclaimer is a horizontal heat exchanger. Certain acidic gases, present in the flue gas feeding the CO_2 absorber, form compounds with the MEA in the solvent solution that cannot be regenerated by application of heat in the solvent stripper reboiler. These materials are referred to as "Heat Stable Salts" (HSS). A small slipstream of the lean solvent from the discharge of the Solvent Stripper Bottoms Pump (GA-2201A/B/C) is fed to the Solvent Reclaimer. The reclaimer restores the MEA usefulness by removing the high boiling and nonvolatile impurities, such as HSS, suspended solids, acids and iron products from the circulating solvent solution. Caustic is added into the reclaimer to free MEA up from its bond with sulfur oxides by its stronger basic attribute. This allows the MEA to be vaporized back into the circulating mixture, minimizing MEA loss. This process is important in reducing corrosion, and fouling in the solvent system. The reclaimer bottoms are cooled (EA-2204) and are supplied to a tank truck without any interim storage.

Solvent Stripper Condenser EA-2206 (Refer to Drawing No. D 09484-01002R-0):

EA-2206 is a water-cooled shell and tube exchanger. The purpose of the condenser is to completely condense all components contained in the overhead vapor stream that can condense under the operating conditions, with the use of cooling water as the condensing medium. Components that do not condense include nitrogen, carbon dioxide, oxygen, nitrogen oxides and carbon monoxide. The water vapor and MEA solvent vapor will condense and the condensed water will dissolve some carbon dioxide. This exchanger uses cooling water capacity freed up due to the reduced load on the existing surface condensers of the power plant. The same is true for the lean solvent cooler (EA-2202).

Solvent Stripper Reflux Drum, FA-2201 (Refer to Drawing No. D 09484-01002R-0):

The purpose of the reflux drum is to provide space and time for the separation of liquid and gases and also provide liquid hold-up volume for suction to the reflux pumps and also provides surge for pre-coat filter. The separation is not perfect, as a small amount of carbon dioxide is left in the liquid being returned to the stripper and the CO_2, saturated with water, is routed to the CO_2 compression and liquefaction system.

Solvent Stripper Reflux Pump, GA-2202 (Refer to Drawing No. D 09484-01002R-0):

This pump takes suction from the reflux drum and discharges on flow control to the stripper top tray as reflux.

Solvent Filtration Package, PA-2251 (Refer to Drawing No. D 09484-01002R-0):

Precoat Filter PA-2251 is no ordinary filter; it is a small system. The main component is a pressure vessel that has a number of so called “leaves” through which MEA flows. The leaves have a thin (1/8 inch) coating of silica powder which acts to filter off any solids. For the purposes of such application the powder is called “filter aid”.

To cover the leaves with the filter aid, the filter must be “precoated” before putting it into service. This is accomplished by mixing filter aid in water in a predetermined ratio (typically 10-wt %) to prepare a slurry. This takes place in an agitated tank. A pump, which takes its suction from this tank, is then operated to pump the slurry into the filter. Provided the flow rate is high enough, the filter aid is deposited on the leaves while water passes through and can be recycled back to the tank. This is continued until the water in the tank becomes clear indicating that all the filter aid has been transferred.

The volume of a single batch in the tank is typically 125% of the filter volume because there must be enough to fill the vessel and have some excess left over so level in the tank is maintained and circulation can continue. In this design, water from the Stripper overhead will be used as make-up water to fill the tank. This way the water balance of the plant is not affected.

During normal operation, it is often beneficial to add so-called “body” which is the same material as the precoat but may be of different particle size. The body is also slurried in water but is continually added to the filter during operation. This keeps the filter coating porous and prevents rapid plugging and loss of capacity. As the description suggests, an agitated tank is needed to prepare the batch. A metering pump is then used to add the body at preset rate to the filter.

When the filter is exhausted (as indicated by pressure drop), it is taken off line so the dirty filter aid can be removed and replaced with fresh material. To accomplish this, the filter must be drained. This is accomplished by pressurizing the filter vessel with nitrogen and pushing the MEA solution out of the filter. After this step, the filter is depressurized. Then, a motor is started to rotate the leaves so a set of scrapers will wipe the filter cake off the leaves. The loosened cake then falls off into a conveyor trough in the bottom of the vessel. This motor operated conveyor then pushes the used cake out of the vessel and into a disposal container (oil drum or similar). The rejected cake has the consistency of toothpaste. This design is called “dry cake” filter and minimizes the amount of waste produced.

For this application, some 2% of the circulating MEA will be forced to flow through the filter. In fact, Filter Circulating Pump GA-2203 draws the liquid through the filter as it has been installed downstream of the filter. The advantage of placing the pump on the outlet side of the filter is reduced design pressure of the filter vessel and associated piping. In spite of the restriction on its suction side, ample NPSH is still available for the pump. Flow is controlled on the downstream side of the pump.

Corrosion Inhibitor (Refer to Drawing No. D 09484-01002R-0):

Corrosion inhibitor chemical is injected into the process constantly to help control the rate of corrosion throughout the CO_2 recovery plant system. Since rates of corrosion increase with high MEA concentrations and elevated temperatures, the inhibitor is injected at appropriate points to minimize the corrosion potential. The inhibitor is stored in a tank (Part of the Package, not shown) and is injected into the system via injection pump (Part of the Package, not shown). The pump is a diaphragm-metering pump.

The selection of metallurgy in different parts of the plant is based on the performance feedback obtained from our similar commercial units in operation over a long period of time.

CO_2 Compression and Liquefaction:

(Refer to Drawing. No. D 09484-01003R-0):

CO_2 from the solvent stripper reflux drum, GA-2201, saturated with water, is compressed in a three stage centrifugal compressor using the air and cooling water from the new cooling tower for interstage and after compression cooling.

The interstage coolers for first and second stage are designed to supply 95 °F CO_2 to the compressor to minimize the compression power requirements.

Most of the water in the wet CO_2 stream is knocked out during compression and is removed from intermediate suction drums. A CO_2 drier is located after the third stage to meet the water specifications for the CO_2 product. The water-free CO_2 is liquefied after the third stage of compression at about 194 psig pressure and is further pumped with a CO_2 pump (GA-2301) to the required battery limit pressure of 2000 psig.

Drying:

CO_2 DRIER, FF-2351 (Refer to Drawing No. D 09484-01003R-0):

The purpose of the CO_2 drier is to reduce the moisture content of the CO_2 product to less than 20 vppm to meet pipeline transport specifications. The drier package, FF-2351, includes four drier vessels, three of which are in service while one is being regenerated or is on standby. The package also includes a natural gas fired regeneration heater and a cooled regeneration cooler. The exchanger will have a knock out cooler downstream for separating the condensed water. The drier size used as a basis for cost estimate is good for 10 hour run length based on 3A molecular sieve.

The drier is located on the discharge side of the 3rd Stage of the CO_2 Compressor. Considering the cost of the vessel and the performance of the desiccant, this is the location favored by vendors. The temperature of the CO_2 stream entering the drier is 90 °F.

Once a bed is exhausted, it is taken off line and a slipstream of effluent from the on line beds is directed into this drier after being boosted in pressure by a compressor. Before the slipstream enters the bed that is to be regenerated, it is heated to a high temperature. Under this high temperature, moisture is released from the bed and carried away in the CO_2 stream. The regeneration gas is then cooled to the feed gas temperature to condense any excess moisture. After this, the regeneration gas stream is mixed with the feed gas upstream of the third stage knockout drum.

All the regeneration operations are controlled by a PLC that switches the position of several valves to direct the flow to the proper drier. It also controls the regeneration compressor, heater, and cooler. Because the regeneration gas has the same composition as the feed gas, it also contains some moisture. Thus, it is primarily the heat (“temperature swing”) that regenerates the bed.

3.2.5.3 Process Flow Diagrams

The processes described above are illustrated in the following process flow diagrams:

- Figure 3.2.5: Drawing D 09484-01001R-0: Flue Gas Cooling and CO_2 Absorption
- Figure 3.2.6: Drawing D 09484-01002R-0: Solvent Stripping
- Figure 3.2.7: Drawing D 09484-01003R-0: CO_2 Compression and Liquefaction

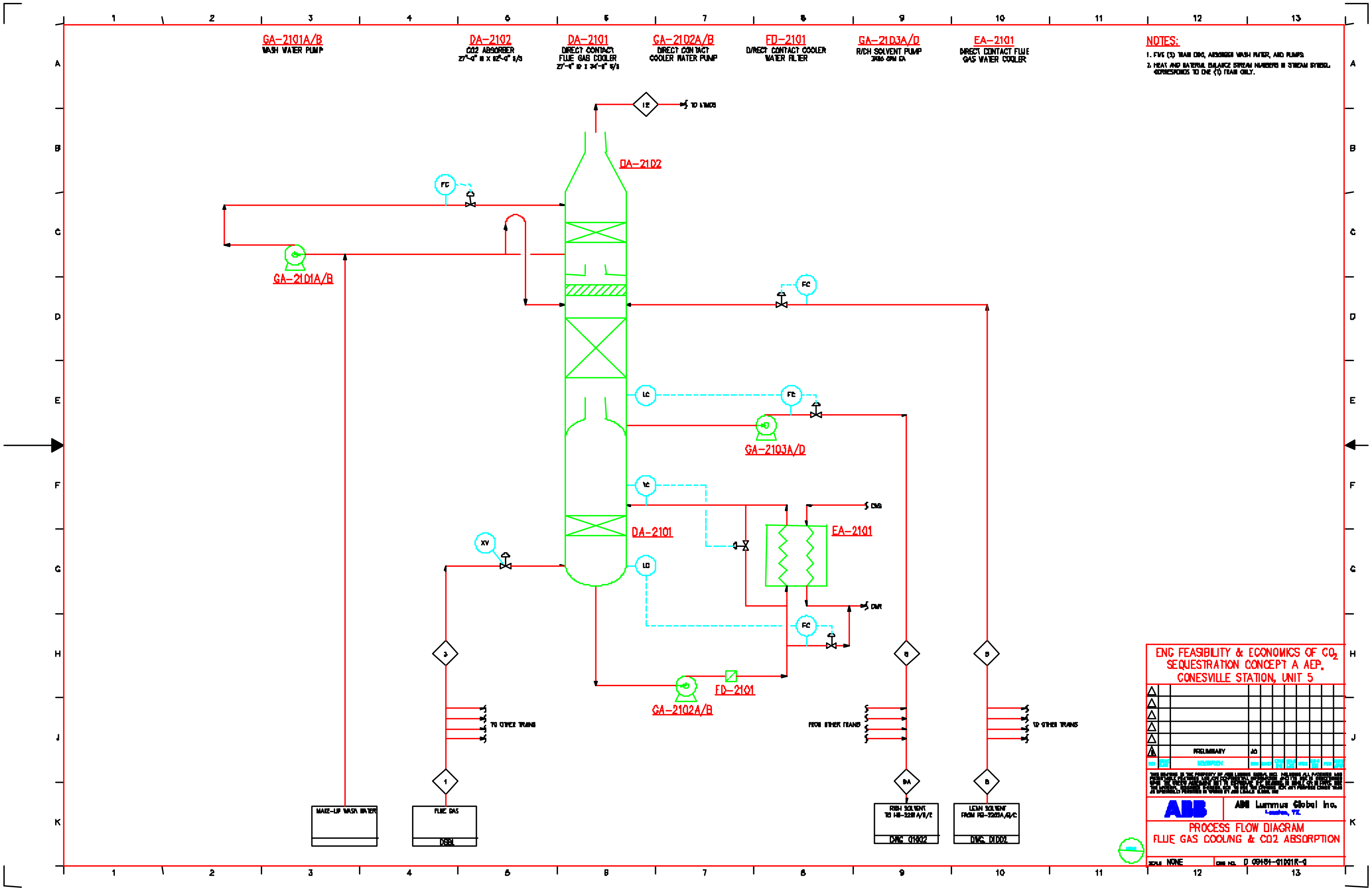


Figure 3.2. 5: Process Flow Diagram for Concept A: Flue Gas Cooling and CO_2 Absorption

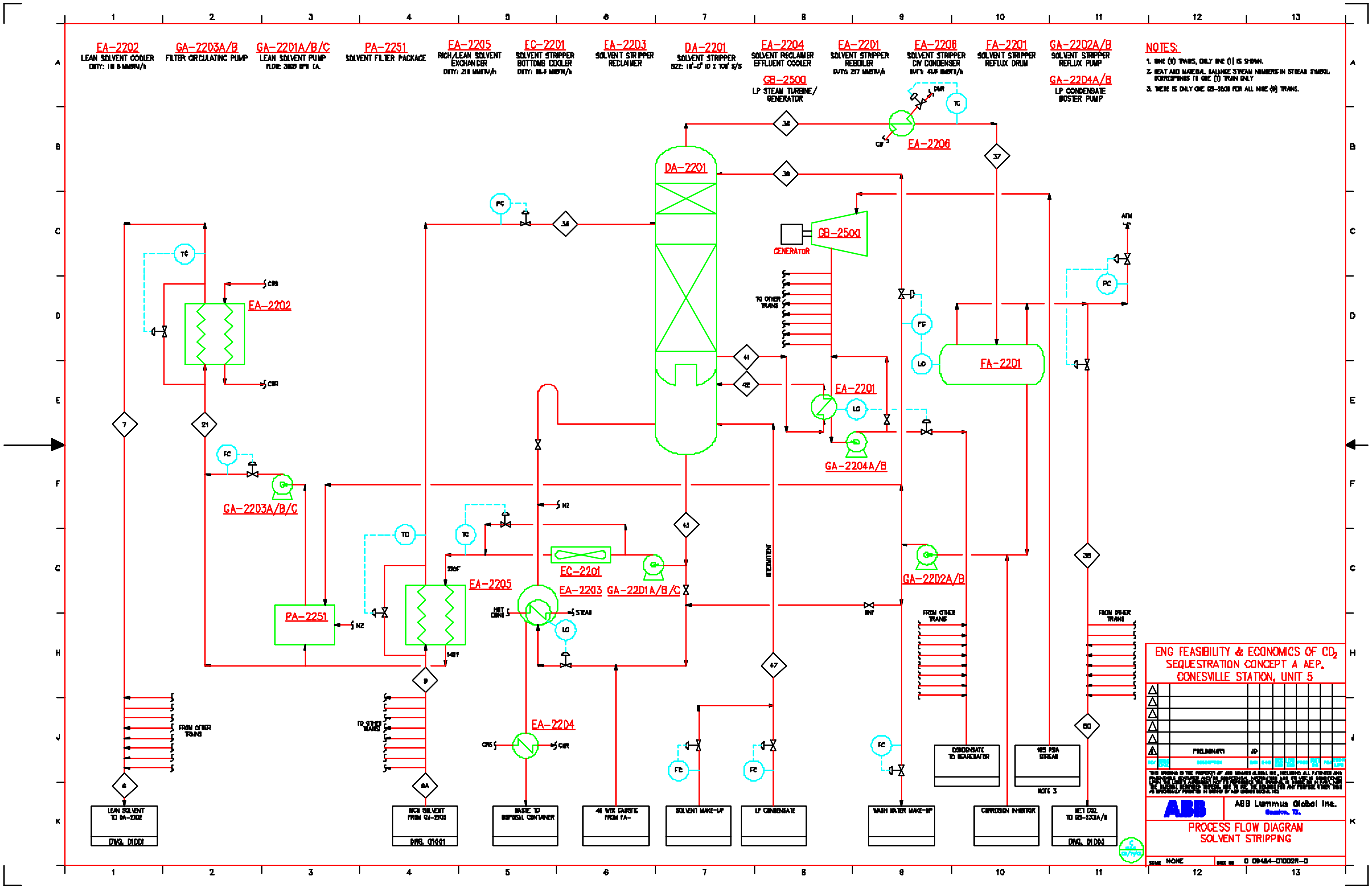


Figure 3.2. 6: Process Flow Diagram for Concept A: Solvent Stripping

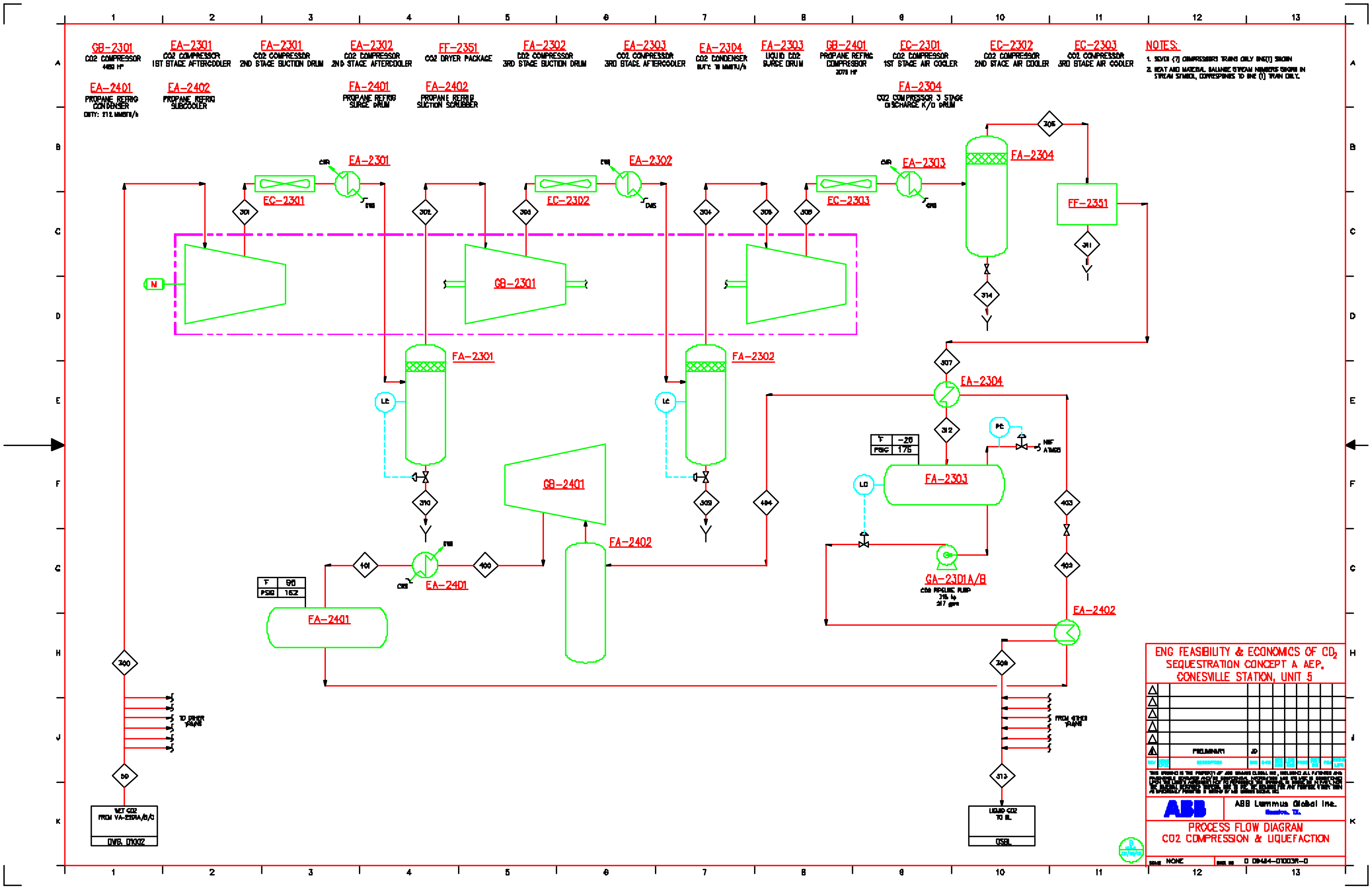


Figure 3.2. 7: Process Flow Diagram for Concept A: CO_2 Compression and Liquefaction

3.2.5.4 Overall Material and Energy Balance

The material balances (Tables 3.2.5 and 3.2.6) were run on two process simulators: Hysim and Amsim. Amsim was used for the Absorption/Stripping systems while Hysim was used for the conventional systems as follows:

Flue Gas feed	Hysim
SO_2 scrubber	ALSTOM Power Program
Absorber and Stripper	Amsim
Compression	Hysim
Heat Pumps	Hysim

The two simulators use a different reference enthalpy. They also use slightly different calculation methods for determining water saturation quantities. There is no simple way to normalize the enthalpies to the same reference. Thus, the enthalpies given in the balance are the values copied directly from the simulation. This creates a discontinuity at the interface between Hysim and Amsim simulations. Take for example the wet CO_2 flow to the CO_2 compressor. The stream comes from the Stripper overhead system, which was simulated with Amsim and enters the CO_2 compressor, which was simulated using Hysim. For this particular stream, the enthalpy value given in the balance comes from Hysim. Lastly, convergence algorithms allow the programs to slightly alter input streams. Thus some leniency and care should be exercised when using such interface streams for heat balance checks.

This section contains heat and material balances for Concept A. See the comments under “Process Flow Diagrams” (Section 3.2.5.3) for comments about stream numbering philosophies.

Table 3.2.5: Material and Energy Balance for Concept A Amine System

STREAM NAME		Total Sour Gas Feed	Sour Gas Feedto Precontactor	PrecontactorGas Outlet	Gas Feed to Absorber A	Absorber A Inlet Separator Liquid	Primary LeanAmine Feed to Absorber A	Rich Amine from Absorber A	Absorber A Total Treated Gas	Total Rich Amine	Rich Amine Feed to Flash Tank	Rich Amine to Lean/Rich Heat Exchanger	Rich Amine from Lean/Rich Heat Exchanger
STREAM NO.		**1**	**3**				**5**	**8**	**12**	**9a**	**9**	**9**	**12**
LIQUID FRACTION		0.000	0.000	0.000	0.000	1.000	1.000	1.000	0.000	1.000	1.000	1.000	0.995
TEMPERATURE	F	150	115	115	115	115	105	133	106	133	133	133	204
PRESSURE	PSIA	16.5	16.5	16.5	16.5	16.5	14.9	16.5	14.9	16.5	16.5	16.5	16.5
COMPONENTS													
CO2 (Carbon Dioxide)	LbMol/Hr	19,684.00	3,936.80	3,936.80	3,936.23	0.14	3,585.44	7,380.58	141.10	36,902.89	4,100.32	4,100.32	4,100.32
MEA	LbMol/Hr	0.00	0.00	0.00	0.00	0.00	16,765.89	16,763.07	2.82	83,815.36	9,312.82	9,312.82	9,312.82
H2O (Water)	LbMol/Hr	24,551.0	4,910.2	4,910.2	2,544.8	2,365.5	227,379.0	228,257.6	1,666.3	1,141,288.0	126,809.8	126,809.8	126,809.8
C1 (Methane)	LbMol/Hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N2 (Nitrogen)	LbMol/Hr	105,079.00	21,015.80	21,015.80	21,016.14	0.02	0.00	1.75	21,014.40	8.76	0.97	0.97	0.97
O2 (Oxygen)	LbMol/Hr	4,518.00	903.60	903.60	903.61	0.00	0.00	0.14	903.47	0.70	0.08	0.08	0.08
Total Molar Flow Rate	LbMol/Hr	153,832.0	30,766.4	30,766.4	28,400.8	2,365.6	247,730.4	252,403.2	23,728.1	1,262,016.0	140,224.0	140,224.0	140,224.0
VAPOR													
MASS FLOW RATE	Lb/Hr	446,600.625	3,572.805	3,572.805	3,397.068				2,438.328				
STD. VOL. FLOW RATE	MMSCFD	1401.1	280.22	280.22	258.66				216.1				
ACTUAL VOL. FLOW RATE	MMACFD	1378	275.6	275.6	254.5				231.72				
MOLECULAR WEIGHT	MW	285.821	57.1642	57.1642	58.9234				55.1246				
STD. DENSITY	Lb/Ft^3	0.765	0.153	0.153	0.1576				0.1354				
GAS COMPRESSIBILITY		0	0	0	0				0				
VISCOSITY	cP	0	0	0	0				0				
HEAT CAPACITY	Btu/Lb-F	0	0	0	0				0				
THERMAL CONDUCTIVITY	Btu/Hr-ft-F	127.9580	25.5916	25.5916	27.7192				1.1892				
LIQUID													
MASS FLOW RATE	Lb/Hr					85.263	10,557.848	10,923.302		273,082.551	3,371.390	3,371.390	3,371.390
STD. VOL. FLOW RATE	GPM					85.26	10252.78	10352.54		51762.70	5751.41	5751.41	5751.41
ACTUAL VOL. FLOW RATE	GPM					86.02	10308.54	10467.22		52336.10	5815.12	5815.12	5940.30
MOLECULAR WEIGHT	MW					18.02	21.31	21.64		21.64	21.64	21.64	21.64
STD. DENSITY	Lb/Ft^3					62.34	64.19	65.77		65.77	65.77	65.77	65.77
VISCOSITY	cP					0.6383	0.8608	0.6868		0.6868	0.6868	0.6868	0.3544
HEAT CAPACITY	Btu/Lb-F					0.9948	0.9357	0.9221		0.9221	0.9221	0.9221	0.9325
THERMAL CONDUCTIVITY	Btu/Hr-ft-F					0.3979	0.3557	0.3557		0.3557	0.3557	0.3557	0.3557

STREAM NAME		Rich Amine Feed to Regenerator	RegeneratorOverhead Vapor	RegeneratorCondenser Outlet	Acid Gas	RegeneratorReflux Liquid	Liquid to Regenerator Reboiler	RegeneratorReboiler Vapor	Lean Amine from RegeneratorReboiler	Lean Amine from Lean/Rich Heat Exchanger	Lean Amine to Cooler	Amine and Water Make-up	Total Acid Gas
STREAM NO.		**35**	**36**	**37**	**38**	**39**	**41**	**42**	**43**	**21**	**21**	**47**	**24**
LIQUID FRACTION		1.000	0.000	1.000	0.000	1.000	1.000	0.000	1.000	1.000	1.000	1.000	0.000
TEMPERATURE	F	209	209	105	105	105	248	250	250	173	173	68	105
PRESSURE	PSIA	28.0	26.0	23.0	23.0	23.0	29.8	30.0	30.0	30.0	30.0	30.0	23.0
COMPONENTS													
CO2 (Carbon Dioxide)	LbMol/Hr	4,100.32	2,081.06	2,081.06	2,079.81	1.27	2,701.12	680.61	2,020.51	2,020.51	2,020.51	0.00	18,718.28
MEA	LbMol/Hr	9,312.82	9.92	9.92	0.01	9.90	9,381.40	68.60	9,312.81	9,312.81	9,314.38	1.58	0.11
H2O (Water)	LbMol/Hr	126,809.8	2,128.7	2,128.7	105.7	2,023.0	137,717.9	11,013.8	126,704.0	126,704.0	126,321.8	(382.3)	951.3
C1 (Methane)	LbMol/Hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N2 (Nitrogen)	LbMol/Hr	0.97	0.97	0.97	0.97	0.00	0.00	0.00	0.00	0.00	0.00	0.00	8.76
O2 (Oxygen)	LbMol/Hr	0.08	0.08	0.08	0.08	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.70
Total Molar Flow Rate	LbMol/Hr	140,224.0	4,220.7	4,220.7	2,186.6	2,034.1	149,800.3	11,763.0	138,037.3	138,037.3	137,656.7	(380.7)	19,679.2
VAPOR													
MASS FLOW RATE	Lb/Hr		221.688		166.131			429.305					121,109.333
STD. VOL. FLOW RATE	MMSCFD		38.44		19.91			107.13					179.20
ACTUAL VOL. FLOW RATE	MMACFD		27.73		13.72			70.62					123.50
MOLECULAR WEIGHT	MW		34.37		47.50			21.97					427.46
STD. DENSITY	Lb/Ft^3		0.12		0.18			0.09					1.62
GAS COMPRESSIBILITY			0.00		0.00			0.00					0.00
VISCOSITY	cP		0.00		0.00			0.00					0.00
HEAT CAPACITY	Btu/Lb-F		0.00		0.00			0.00					0.00
THERMAL CONDUCTIVITY	Btu/Hr-ft-F		54.78		105.69			6.43					951.17
LIQUID													
MASS FLOW RATE	Lb/Hr	3,371,390		145,088		41,234	3,525,978		3,267,542	3,267,542	3,259,998	-7,547	
STD. VOL. FLOW RATE	GPM	5751.41		247.18		73.13	6116.13		5709.78	5709.78	5696.53	-13.59	
ACTUAL VOL. FLOW RATE	GPM	5951.79		248.73		73.61	6434.23		6011.14	5839.38	5825.79	-13.6	
MOLECULAR WEIGHT	MW	21.64		30.94		18.24	21.18		21.30	21.30	21.31	17.84	
STD. DENSITY	Lb/Ft^3	65.77		65.86		63.27	64.69		64.21	64.21	64.21	62.31	
VISCOSITY	cP	0.3401		0.6888		0.6655	0.2592		0.2564	0.4548	0.4549	1.2839	
HEAT CAPACITY	Btu/Lb-F	0.9324		0.4962		0.9902	0.9481		0.9491	0.9513	0.9513	0.9454	
THERMAL CONDUCTIVITY	Btu/Hr-ft-F	0.3557		0.3945		0.3944	0.3583		0.3557	0.3557	0.3557	0.3664	

Table 3.2. 6: Material and Energy Balance for Concept A CO_2 Compression and Liquifaction System

STREAM NAME		Total Acid gas from strippers	To train A 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
STREAM NO.		**300**	**300**	**301**	**302**	**310**	**303**	**304**	**309**	**306**	**305**	**314**
VAPOR FRACTION	Molar	1.000	1.000	1.000	1.000	0.000	1.000	1.000	0.000	1.000	1.000	0.000
TEMPERATURE	F	105	105	230	95	95	236	95	95	282	90	90
PRESSURE	PSIG	4	4	25	19	19	62	56	56	191	185	185
MOLAR FLOW RATE	LbMol/Hr	19,679.08	2,811.30	2,811.30	2,743.70	67.60	2,743.70	2,708.50	35.19	2,708.50	2,686.56	21.94
MASS FLOW RATE	Lb/Hr	841,192	120,170	120,170	118,951	1,219	118,951	118,315	636	118,315	117,917	398
ENERGY	Btu/Hr	8.79E+07	1.26E+07	1.58E+07	1.19E+07	-9.79E+05	1.56E+07	1.17E+07	-5.09E+05	1.64E+07	1.10E+07	-3.18E+05
COMPOSITON	Mol %											
CO2		95.12%	95.12%	95.12%	97.46%	0.09%	97.46%	98.72%	0.18%	98.72%	99.52%	0.54%
H2O		4.83%	4.83%	4.83%	2.49%	99.91%	2.49%	1.23%	99.82%	1.23%	0.42%	99.46%
Nitrogen		0.04%	0.04%	0.04%	0.05%	0.00%	0.05%	0.05%	0.00%	0.05%	0.05%	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%
Propane		0.00%	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%	0.00%
VAPOR												
MOLAR FLOW RATE	LbMol/Hr	19,679.1	2,811.3	2,811.3	2,743.7	-	2,743.7	2,708.5	-	2,708.5	2,686.6	-
MASS FLOW RATE	Lb/Hr	841,192	120,170	120,170	118,951	-	118,951	118,315	-	118,315	117,917	-
STD VOL. FLOW	MMSCFD	179.23	25.60	25.60	24.99	-	24.99	24.67	-	24.67	24.47	-
ACTUAL VOL. FLOW	ACFM	103,907.68	14,843.95	8,749.53	8,063.83	-	4,417.63	3,728.32	-	1,698.44	1,224.03	-
MOLECULAR WEIGHT	MW	42.75	42.75	42.75	43.35	-	43.35	43.68	-	43.68	43.89	-
DENSITY	Lb/Ft^3	0.13	0.13	0.23	0.25	-	0.45	0.53	-	1.16	1.61	-
VISCOSITY	cP	0.0149	0.0149	0.0187	0.0149	-	0.0193	0.0152	-	0.0212	0.0154	-
HYDROCARBON LIQUID												
MOLAR FLOW RATE	LbMol/Hr	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	Lb/Hr	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-
DENSITY	Lb/Ft^3	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-

STREAM NAME		From drier / To condenser	Water from drier	From condenser	From product pump	From Train A liquefaction	To pipeline	Refrig compressor discharge	From refrig condenser	From subcooler	Refrig to CO2 condenser	Refrig from CO2 condenser
STREAM NO.		**307**	**311**	**312**	**308**	**309**	**313**	**400**	**401**	**402**	**403**	**404**
VAPOR FRACTION	Molar	1.000	0.726	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.173	0.996
TEMPERATURE	F	90	380	-26	-12	82	82	65	95	24	-31	-31
PRESSURE	PSIG	180	180	2,003	2,000	2,000	2,000	55	162	159	5	5
MOLAR FLOW RATE	LbMol/Hr	2,675.15	11.41	2,675.15	2,675.15	2,675.15	18,726.05	2,928.57	2,928.57	2,928.57	2,928.57	2,928.57
MASS FLOW RATE	Lb/Hr	117,711	206	117,711	117,711	117,711	823,979	129,141	129,141	129,141	129,141	129,141
ENERGY	Btu/Hr	1.10E+07	2.51E+04	-8.07E+06	-7.29E+06	-1.36E+06	-9.50E+06	1.81E+07	7.63E+05	-5.17E+06	-5.17E+06	1.39E+07
COMPOSITON	Mol %											
CO2		99.95%	0.00%	99.95%	99.95%	99.95%	99.95%	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%	0.00%
Nitrogen		0.05%	0.00%	0.05%	0.05%	0.05%	0.05%	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%
Propane		0.00%	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%	0.00%
VAPOR												
MOLAR FLOW RATE	LbMol/Hr	2,675.2	8.3	-	-	-	-	2,928.6	-	-	506.5	2,915.8
MASS FLOW RATE	Lb/Hr	117,711	149	-	-	-	-	129,141	-	-	22,334	128,577
STD VOL. FLOW	MMSCFD	24.36	0.08	-	-	-	-	26.67	-	-	4.61	26.56
ACTUAL VOL. FLOW	ACFM	1,253.44	5.96	-	-	-	-	3,573.03	-	-	1,860.34	10,709.92
MOLECULAR WEIGHT	MW	44.00	18.02	-	-	-	-	44.10	-	-	44.10	44.10
DENSITY	Lb/Ft^3	1.57	0.42	-	-	-	-	0.60	-	-	0.20	0.20
VISCOSITY	cP	0.0155	0.0154	-	-	-	-	0.0082	-	-	0.0065	0.0065
HYDROCARBON LIQUID												
MOLAR FLOW RATE	LbMol/Hr	-	-	2,675.15	2,675.15	2,675.15	18,726.05	-	2,928.57	2,928.57	2,422.10	12.79
MASS FLOW RATE	Lb/Hr	-	-	117,711.33	117,711.33	117,711.33	823,979.29	-	129,141.22	129,141.22	106,807.22	563.95
STD VOL. FLOW	BPD	-	-	9,766	9,766	9,766	68,360	-	17,452	17,452	14,434	76
ACTUAL VOL. FLOW	GPM	-	-	217.05	213.53	289.79	2,028.56	-	541.52	480.49	372.27	1.97
DENSITY	Lb/Ft^3	-	-	67.61	68.73	50.64	50.64	-	29.73	33.51	35.77	35.77
MOLECULAR WEIGHT	MW	-	-	44.00	44.00	44.00	44.00	-	44.10	44.10	44.10	44.10
VISCOSITY	cP	-	-	0.1752	0.1607	0.0620	0.0620	-	0.0906	0.1332	0.1823	0.1823
SURFACE TENSION	Dyne/Cm	-	-	16.07	14.07	0.86	0.86	-	5.74	10.51	14.49	14.49

3.2.5.5 Equipment List with Data

Equipment data for Concept A, provided in Appendix II, has been presented in the so-called “short spec” format which provides adequate data for a factored cost estimate.

3.2.5.6 Consumption of Utilities

The following utilities from outside boundary limits (OSBL) are required in the CO_2 Recovery Unit.

- Steam
- High Pressure (HP) Steam
- Low Pressure (LP) Steam
- Water
- Demineralized Water
- Raw Water (Fresh Water) (Cooling tower make-up)
- Potable Water (hoses, etc.)
- Air
- Plant Air (maintenance, etc.)
- Instrument Air
- Electric Power
- Natural Gas

Note: The CO_2 Recovery Plant includes cooling water pumps that supply all the cooling water required by this unit.

Concept A utility consumption is presented in Table 3.2.7.

Table 3.2. 7: Concept A Utility Consumption

Utility	Amount Consumed	Units
Natural Gas	.42	MMSCFD
Steam (180 psig)	1,950,000	Lb/hr
Cooling water	22,000	Gpm

Number of Trains	Tag no.	Description	Number Operating per train	Power (ea) including 0.95 motor eff (kW)	Total all trains (kW)
5	GA-2101 A/B	Wash Water Pump	1	19	95
5	GA-2102 A/B	Direct Contact Cooler Water Pump	1	32	162
5	GA-2103 A/B/C/D	Rich Solvent Pump	3	146	729
9	GA-2201A/B/C	Lean Solvent Pump	2	117	1,053
9	GA-2202 A/B	Solvent Stripper Reflux Pump	1	3	28
9	GA-2203 A/B	Filter Circ. Pump	1	12	107
7	GA-2301 A/B	CO2 Pipeline Pump	1	184	1,288
9	GA-2204 A/B	LP condensate booster pump	1	74	667
3	GA-2501	Caustic metering pump	1	0	0

7	GB-2301	CO_2 Compressor (Motor driven)	1	3,557	24,901
7	GB-2401	Propane Refrig Compressor	1	2,395	16,765
1	GB-2500	LP steam turbine/ generator	NA	NA	NA
7	EC-2301	CO2 compressor 1st stage air cooler	1	9	66
7	EC-2302	CO2 compressor 2nd stage air cooler	1	10	69
7	EC-2303	CO2 compressor 3rd stage air cooler	1	15	103
9	EC-2201	Solvent stripper bottoms cooler	1	256	2,305
7	PA-2351	CO2 Drier Package	1	151	1054
1	PA-2551	Cooling Tower	1	962	962
		Total			50,355

3.2.5.7 Consumption of Chemicals and Desiccants

The consumption of chemicals and desiccants for Concept A are identified in Table 3.2.8 shown below.

Table 3.2.8: Concept A Chemicals and Desiccants Consumption

Chemical	Consumption per day (lbs.)
Caustic (100%)	3600
MEA	14000
Corrosion inhibitor	1140
Diatomaceous earth	916
Molecular sieve	257
Sodium hypochlorite	3590
Sodium bisulfite	13.8

This total does not include chemicals provided by the cooling tower service people nor disposal of waste. These are handled as a component of operating costs referred to as contracted services and waste handling, respectively.

3.2.5.8 Design Considerations

The following parameters were optimized for Concept A with the objective of reducing the overall unit cost and energy requirements.

- Solvent Concentration
- Lean Amine Loading
- Rich Amine Loading
- Absorber Temperature
- Rich /Lean Exchanger approach
- CO_2 Compressor inter-stage temperatures
- CO_2 Refrigeration Pressure and Temperature

A minimum of 90% CO_2 recovery was targeted. The above parameters were adjusted to increase the recovery until a significant increase in equipment size and/ or energy consumption was observed. AES Corporation owns and operates a 200 STPD food grade CO_2 production plant in Oklahoma. This plant was designed and built by ABB Lummus

Global as a part of the larger power station complex using coal fired boilers. This plant was started up in 1990 and has been operating satisfactorily with lower than designed MEA losses. The key process parameters from the present design for Concept A are compared with those from the AES plant (Barchas and Davis, 1992) in Table 3.2.9.

Table 3.2. 9: Concept A Key Process Parameters Comparison

PROCESS PARAMETER	AEP DESIGN (Concept A)	AES DESIGN
PLANT CAPACITY TPD	9888	200
CO_2 in Feed, % mol	13.9	14.7
O_2 in Feed, % mol	3.2	3.4
SO_2 in Feed, % mol	100 ppm	10 PPMV (Max)
SOLVENT	MEA	MEA
SOLVENT CONC. %WT	20	15 (Actual 17-18%Wt)
LEAN LOADING	0.21	0.10
RICH LOADING	0.44	0.41
STRIPPER FEED TEMP , F	210	194
STRIPPER BOTTOM TEMP F	250	245
FEED TEMP TO ABSORBER, F	105	108
CO_2 RECOVERY %	95	90 (ACTUAL 96-97%)
ABSORBER PRESSURE DROP psi	1	1.4
STRIPPER PRESSURE DROP psi	0.6	4.35
R/L EXCHANGER APPROACH,F	10	50
CO_2 COMPRESSOR I/STG TEMP F	105	115
LIQUID CO_2 TEMP F	82	-13
STEM CONSUMPTION (T steam/ T CO2)	2.6	3.45
LIQUID CO_2 PRESSURE psi (A)	2015	247

3.2.5.9 OSBL Systems

Reclaimer Bottoms (Concept A):

The reclaimer bottoms are generated during the process of recovering MEA from heat stable salts (HSS), which are produced from the reaction of MEA with SO_2 and NO_2. The HSS accumulate in the reclaimer during the lean amine feed portion of the reclaiming cycle. The volume of reclaimer bottoms generated will depend on the quantity of SO_2 and NO_2 that is not removed in the Flue Gas Scrubber. A typical composition of the waste is presented in Table 3.2.10.

Table 3.2. 10: Reclaimer Bottoms Composition

MEA	9.5 wt.%
NH_3	0.02 wt.%
NaCl	0.6 wt.%
Na_2SO_4	6.6 wt.%
Na_2CO_3	1.7 wt.%
Insolubles	1.3 wt.%
Total Nitrogen	5.6 wt.%
Total Organic Carbon	15.6 wt.%
PH	10.7
Specific Gravity	1.14

Filter Residues:
A pressure leaf filter filters a slipstream of lean amine. Diatomaceous earth is used as a filter-aid for pre-coating the leaves and as a body feed. Filter cycles depend on the rate of flow through the filter, the amount of filter aid applied, and the quantity of contaminants in the solvent. A typical composition of the filter residue is provided in Table 3.2.11 below. These will be disposed of by a contracted service hauling away the drums of spent cake.

Table 3.2. 11: Filter Residue Composition

MEA	2.6 wt.%
Total Organic Carbon	1.5 wt.%
SiO_2	0.03 wt.%
Insolubles	0.03 wt.%
PH	10.0
Specific Gravity	1.0

Excess Solvent Stripper Reflux Water:
The CO_2 Recovery Facility has been designed to operate in a manner to avoid accumulation of water in the Absorber / Stripper system. Conversely, no continuous make-up stream of water is required, either. By controlling the temperature of the scrubbed flue gas to the absorber, water balance of the MEA system can be kept in check. Excess water can accumulate in the Stripper Reflux Drum and can be reused once the system is corrected to operate in a balanced manner. Should water need to be discarded, contaminants will include CO_2 and MEA.

Secondary FGD Absorber Effluent:

The existing plant uses lime in its flue gas desulfurizer. In the cost estimate of this plant, it has been assumed that the existing plant disposal facilities can include the relatively small additional load of the secondary regenerator.

Cooling Tower Blowdown:

The composition limits on cooling tower blowdown are shown in Table 3.2.12.

Table 3.2. 12: Cooling Tower Blowdown Composition Limitations

Component	Specification
Suspended Solids	30 ppm monthly, 100 ppm daily
PH	6.5 to 9
Oil and Grease	15 ppm maximum monthly, 20 ppm maximum daily
Free Chlorine	0.035 ppm

There is a thermal limit specification for the entire river. However, the blowdown volume is too small to affect it significantly.

Relief Requirements:

The relief valve discharges from the CO_2 Recovery Unit are discharged to atmosphere. No tie-ins to any flare header are necessary.

3.2.5.10 Plant Layout

The new equipment required for Concept A covers about 7 acres of plot area. Plant layout drawings prepared for the Concept A CO_2 Recovery System are as follows:

These drawings are shown in Appendix I.

Plot Plan – Overall Site before CO_2 Unit Addition

U01-D-0208 Plot Plan – Concept A: Flue Gas Cooling & CO_2 Absorption
U01-D-0214 Plot Plan – Concept A: Solvent Stripping
U01-D-0204 Plot Plan – Concept A: CO_2 Compression & Liquefaction
U01-D-0211 Plot Plan – Concept A: Overall Layout Conceptual Plan
U01-D-0200R Plot Plan – Concept A: Modified Overall Site Plan

Plant layout has been designed in accordance with a spacing chart called “Oil and Chemical Plant Layout and Spacing” Section IM.2.5.2 issued by Industrial Risk Insurers (IRI).

When reviewing the layout, the first thing to observe is that no highly flammable materials are handled within the CO_2 Recovery Unit. The open cup flash point of MEA is 200 ^oF and, therefore, will not easily ignite. In addition to MEA, the corrosion inhibitor is the only other hydrocarbon liquid within the battery limits. The flash point of this material is higher than that of MEA and is handled in small quantities.

As the chemicals used in the process present no fire hazard, there is an opportunity to reduce the minimum spacing between equipment from that normally considered acceptable in hydrocarbon handling plants. Regardless, for the drawings that follow, standard spacing requirements, as imposed by IRI have been followed.

The plot area available for the installation of the desired equipment is small. Some equipment items are placed on structures to allow other pieces of equipment to be placed underneath them. This way pumps and other equipment associated with the Absorber can be located under the structure. Locating the pumps under the structure has been considered acceptable because the fluids being pumped are not flammable.

Noise is an issue with the flue gas fan as much as it is with compressors. Discussions with vendors suggest that it will be possible to provide insulation on the fan casing to limit noise to acceptable levels. Therefore, it has been assumed that no building needs to be provided for noise reasons.

Having economized on the required plot space as noted above, it was judged not to be practical to divide up the absorbers and strippers that are required into the relatively small plot areas initially offered for this purpose (refer to Section 2.1 and Figure 2.1.1). Eventually it was agreed that the units would be placed in an area about 1200 feet north of the Unit #5 stack (refer to Figure 3.1.1). By locating the units on a single site, the flue gas duct length and MEA piping between the absorber and stripper could be minimized, although the latter impact is not nearly as important as the flue gas routing.

The corrosion inhibitor must be protected against freezing during winter. The caustic solution will not freeze but will become very viscous when it gets cold. Therefore, a heated shed has been provided for housing the Corrosion Inhibitor and the Caustic injection packages.

The plot plan shows a substation in the Stripper area but none for the Absorber area. The assumption is that because the electrical consumption of the Absorber equipment is small (0.23 MW) compared to the Stripper equipment, the equipment can be run directly from the auxiliary power 480 volt power system.

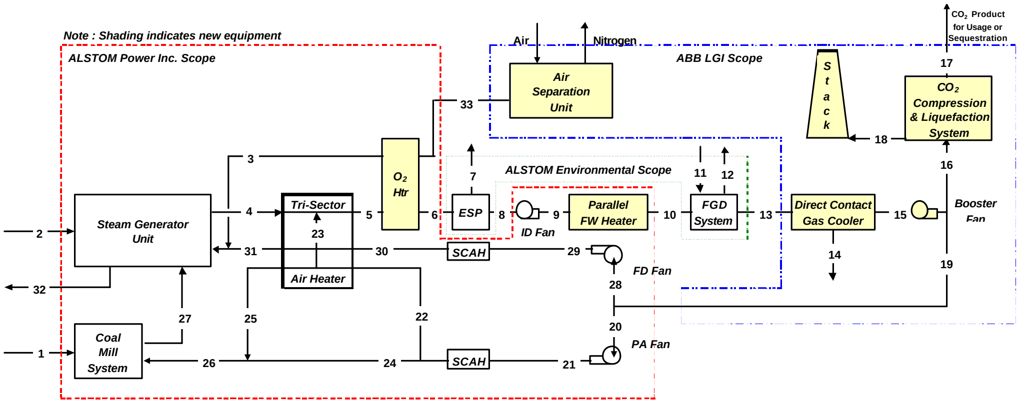
For the Rich/Lean Solvent Exchanger which is a plate and frame type exchanger, area estimates received from vendors based on similar conditions suggest that five units/ train would be sufficient for the specified service.

3.3 Concept B: CO_2 Separation with Oxygen Firing and Flue Gas Recirculation

The basic idea of the overall system for Concept B is to replace air with oxygen for combustion in the furnace in order to produces a high carbon dioxide content flue gas stream leaving the boiler island. A stream of re-circulated flue gas to the furnace is required to maintain thermal balance in the existing boiler between the lower furnace region where evaporation takes place and the convective heat transfer surfaces where steam is superheated and reheated to the required temperature level. This arrangement produces a high carbon dioxide content flue gas that after leaving the boiler system, is processed to provide high-pressure carbon dioxide liquid product for sequestration, enhanced oil recovery (EOR) or other uses.

3.3.1 Overall System Description and Material and Energy Balance

A simplified system diagram for the modified unit is shown in Figure 3.3.1. The system was designed to provide maximum flexibility of operation and to facilitate combustion of coal in either air or oxygen and recirculated flue gas mixture environment.



Material Flow Stream Identification

1	Raw Coal to Pulverizers
2	Air Infiltration Stream
3	Heated Oxygen Stream to Furnace
4	Fluegas from Economizer to Air Heater
5	Fluegas Leaving Air Heater to Oxygen Heater
6	Fluegas Leaving Oxygen Heater to ESP
7	Flyash Leaving ESP
8	Fluegas Leaving ESP to ID Fan
9	Fluegas to Parallel FW Heater
10	Fluegas Leaving Parallel FW Heater to FGD
11	Lime Feed to FGD System
12	FGD System Solids to Disposal
13	Fluegas Leaving FGD to Gas Cooler
14	Condensed Water Stream
15	Cooled Fluegas Stream to Booster Fan
16	Fluegas to CO_2 Compression System
17	Compressed CO_2 Product
18	Fluegas Vent to Stack
19	Cooled Recirculated Fluegas Stream
20	Primary Fluegas to PA Fan
21	Pressurized Primary Fluegas
22	Primary Fluegas to Air Heater
23	Air Heater Leakage Gas Stream
24	Tempering Fluegas to Pulverizers
25	Hot Primary Fluegas to Pulverizers
26	Mixed Primary Fluegas to Pulverizers
27	Pulverized Coal and Fluegas to Furnace
28	Secondary Fluegas to FD Fan
29	Secondary Fluegas to Steam Coil Air Heater
30	Secondary Fluegas to Air Heater
31	Heated Secondary Fluegas to Furnace
32	Bottom Ash from Furnace
33	Oxygen Stream from Air Separation Unit

Figure 3.3. 1: Simplified Gas Side Process Flow Diagram for CO_2 Separation with Oxygen Firing

Raw coal (Stream 1) from the coal bunkers is supplied to the existing coal mills where it is pulverized and transported with recycled flue gas (Stream 27) to the furnace. The air separation unit (ASU) supplies the oxidant for the existing boiler. There is no air supply to the boiler except for air infiltration (Stream 2). The oxygen produced, about 8,924 tons per day, (Stream 33) flows through a heat exchanger where it is preheated by cooling the flue gas leaving the existing air heater (Stream 5). The pulverized coal is combusted in the furnace with a mixture of preheated oxygen (Stream 3) and preheated recycled flue gas (Stream 31).

The products of combustion leave the furnace and are cooled in a series of heat exchangers. The gas leaving the furnace first enters the existing convection pass of the unit where steam generated in the existing furnace walls is superheated and reheated in the existing convection pass heat exchangers. The flue gas leaves the convection pass (Stream 4) and is further cooled in the existing air heater of the unit. The air heater, however, is now used to heat recycled flue gas since air is no longer supplied to the unit in this mode of operation. The flue gas stream leaving the air heater (Stream 5) flows through the oxygen heater, electrostatic precipitator, induced draft fan, parallel feedwater heater, flue gas de-sulfurization unit, gas cooler and booster fan, in series, before it is split into two streams. One stream (Stream 16) represents the exhaust gas stream leaving the boiler island. This stream provides the feed stream for the Carbon Dioxide Separation and Compression System (described in detail in Section 3.3.7). The remaining flue gas (Stream 19), which is roughly twice as large as Stream 16, is recycled back to the unit with the forced draft and primary air fans. Figure 3.3.2 shows the cooling curve for the flue gas leaving the existing boiler.

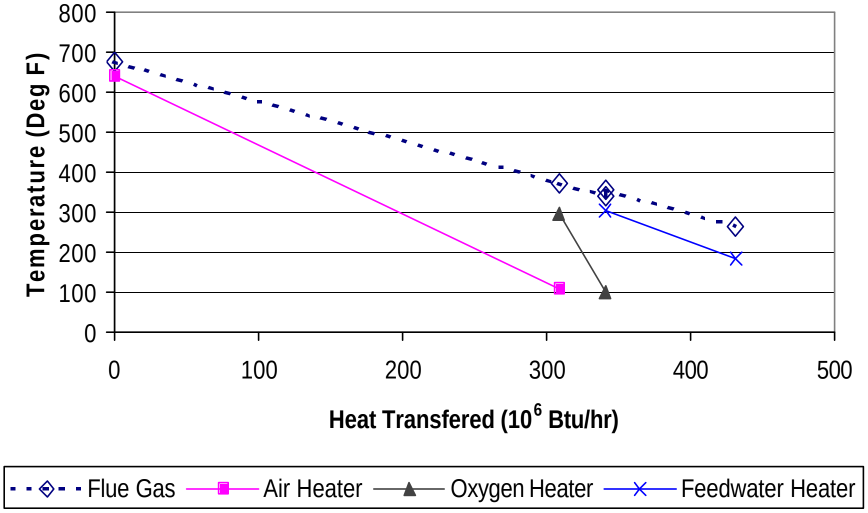
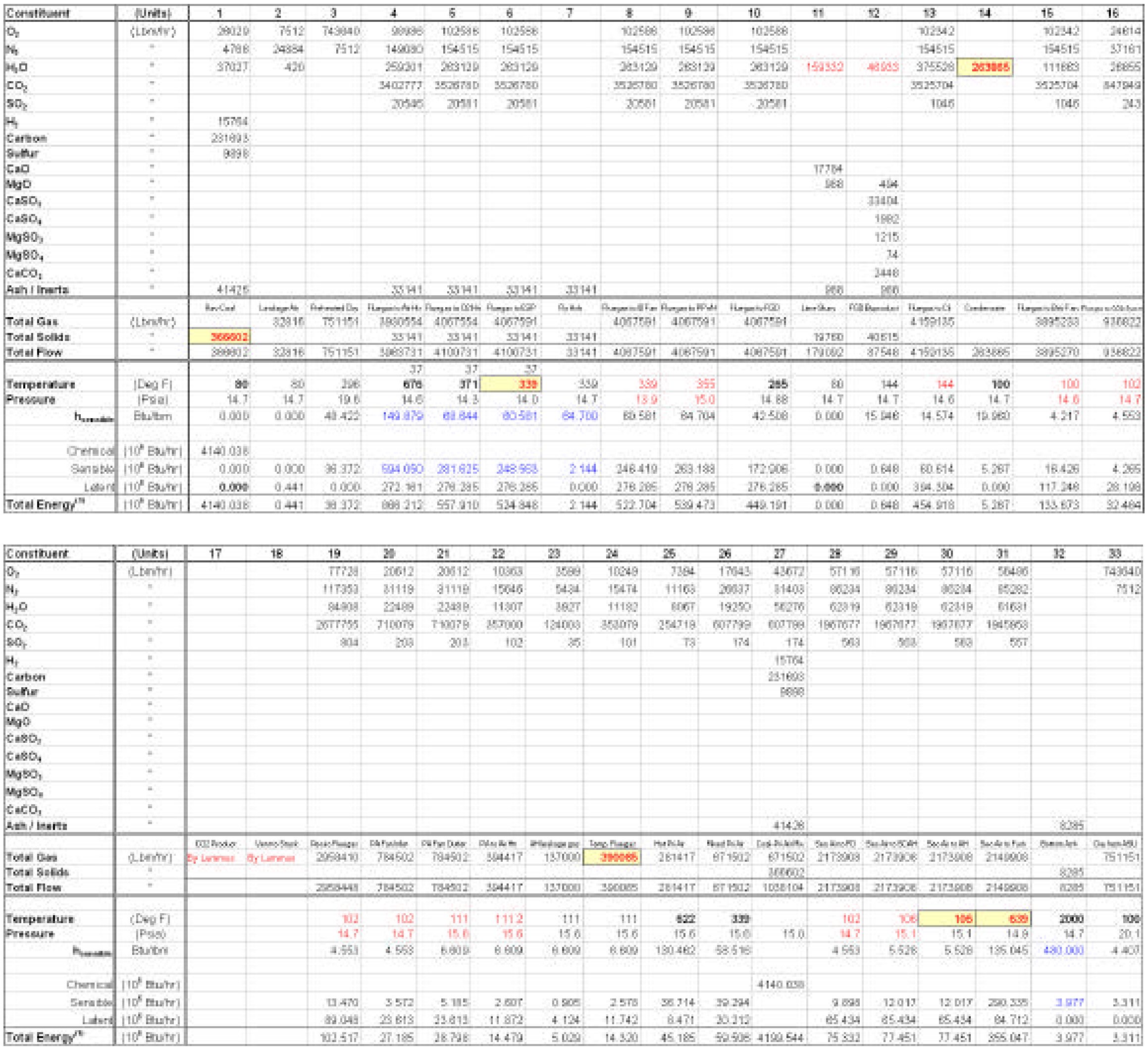


Figure 3.3. 2: Boiler Flue Gas Cooling Curve

The primary reason for the large recycle stream is to maintain the thermal balance between heat transferred in the radiant furnace and the convective heat transfer surfaces, and to generate required boiler performance. In addition, gas temperatures throughout the unit must be low enough to assure the ash, which is contained within the fuel, is maintained in a state where cleanability of the unit is not compromised. Additionally, the integrity of the existing metallurgy in the furnace walls and convective pass heat exchangers must be ensured. The recycled flue gas is supplied to the unit through a combination of new ducts and the existing air ducts. These recycle streams (Streams 29, 30, 31, 21, 22, 24, 25, and 26) provide the primary and secondary "air" streams for the air heater as well as the tempering “air” stream for outlet temperature control of the pulverizers. The modified system was designed to generate approximately 3.1 x 10^6 lbm/hr of steam, which represents the Maximum Continuous Rating (MCR) for the unit.

Table 3.3.1, below, shows constituent mass flows, temperatures, pressures, enthalpies, and energy values (chemical, sensible, latent, and total) for all the points shown in the system diagram of Figure 3.3.1 This table therefore provides the gas side energy and material balance for the system. Two of the key assumptions used in the development of this material and energy balance were an oxygen stream purity of 99 percent by weight, and an air infiltration rate equivalent to one percent of the total oxygen required for the process. For the definition of state points 17 and 18, (not defined in Table 3.3.1) refer to Section 3.3.7.4, which provides the complete material and energy balance for the CO_2 Compression and Liquifaction System.

Table 3.3. 1: Gas Side Material and Energy Balance for Concept B



Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
O_2	(Lbm/hr)	28029	7512	743640	98986	102588	102588		102588	102588	102588			102342		102342	24614
N_2	"	4788	24884	7512	149080	154515	154515		154515	154515	154515			154515		154515	37161
H_2O	"	37027	420		259301	263129	263129		263129	263129	263129	159332	48933	375528	283885	111883	26855
CO_2	"				3402777	3526780	3526780		3526780	3526780	3526780			3525704		3525704	847949
SO_2	"				20545	20581	20581		20581	20581	20581			1046		1046	243
H_2	"	15764															
Carbon	"	231893															
Sulfur	"	9898															
CaO	"											17784					
MgO	"											988	494				
$CaSO_3$	"												33404				
$CaSO_4$	"												1982				
$MgSO_3$	"												1215				
$MgSO_4$	"												74				
$CaCO_3$	"												3448				
Ash / Inerts	"	41425			33141	33141	33141	33141				988	988				
		Raw Coal	Leakage Air	Preheated Oxy	Flue gas to Air Htr	Flue gas to O2 Htr	Flue gas to ESP	Fly Ash	Flue gas to ID Fan	Flue gas to RFAH	Flue gas to FGD	Lime Slurry	FGD Byproduct	Flue gas to CI	Condensate	Flue gas to Dist Fan	Flue gas to CO2 Sys
Total Gas	(Lbm/hr)		32816	751151	3880554	4067554	4067591		4067591	4067591	4067591			4159135		3895283	938822
Total Solids	"	366602			33141	33141	33141	33141				19760	40515				
Total Flow	"	388802	32816	751151	3983731	4100731	4100731	33141	4067591	4067591	4067591	179092	87548	4159135	283885	3895270	938822
					37	37	37										
Temperature	(Deg F)	80	80	396	676	371	339	339	339	355	285	80	144	144	100	100	102
Pressure	(Psia)	14.7	14.7	19.6	14.6	14.3	14.0	14.7	13.9	15.0	14.88	14.7	14.7	14.6	14.7	14.6	14.7
$h_{sensible}$	Btu/lbm	0.000	0.000	48.422	149.879	68.844	60.581	64.700	60.581	64.704	42.508	0.000	15.946	14.574	19.980	4.217	4.553
Chemical	(10^6 Btu/hr)	4140.036															
Sensible	(10^6 Btu/hr)	0.000	0.000	36.372	594.050	281.625	248.958	2.144	246.419	263.188	172.906	0.000	0.648	60.614	5.287	16.426	4.265
Latent	(10^6 Btu/hr)	0.000	0.441	0.000	272.161	276.285	276.285	0.000	276.285	276.285	276.285	0.000	0.000	394.304	0.000	117.248	28.198
Total Energy$^{(1)}$	(10^6 Btu/hr)	4140.036	0.441	36.372	866.212	557.910	524.846	2.144	522.704	539.473	449.191	0.000	0.648	454.918	5.287	133.673	32.464

Constituent	(Units)	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33
O_2	(Lbm/hr)			77728	20612	20612	10363	3599	10249	7384	17643	43672	57116	57116	57116	56486		743640
N_2	"			117353	31119	31119	15646	5434	15474	11163	26837	31403	86234	86234	86234	85282		7512
H_2O	"			84908	22489	22489	11307	3827	11182	8067	19290	56276	62319	62319	62319	61631		
CO_2	"			2677755	710079	710079	357000	124003	353079	254718	607799	607799	1967677	1967677	1967677	1945953		
SO_2	"			804	203	203	102	35	101	73	174	174	563	563	563	557		
H_2	"											15764						
Carbon	"											231893						
Sulfur	"											9898						
CaO	"																	
MgO	"																	
$CaSO_3$	"																	
$CaSO_4$	"																	
$MgSO_3$	"																	
$MgSO_4$	"																	
$CaCO_3$	"																	
Ash / Inerts	"											41428					8285	
		CO2 Product	Vent to Stack	Recirc Flue gas	PA Fan Inlet	PA Fan Outlet	PA to Air Htr	PA leakage gas	Temp. Flue gas	Hot PA Air	Mixed PA Air	Coal-PA-Air Mix	Sec Air to FD	Sec Air to SCAH	Sec Air to AH	Sec Air to Furn	Bottom Ash	O2 from ASU
Total Gas	(Lbm/hr)	By Lummus	By Lummus	2958410	784502	784502	394417	137000	390085	261417	671502	671502	2173908	2173908	2173908	2149908		751151
Total Solids	"											366602					8285	
Total Flow	"			2958448	784502	784502	394417	137000	390085	261417	671502	1038104	2173908	2173908	2173908	2149908	8285	751151
Temperature	(Deg F)			102	102	111	111.2	111	111	622	339		102	106	106	639	2000	100
Pressure	(Psia)			14.7	14.7	15.6	15.6	15.6	15.6	15.6	15.6	15.0	14.7	15.1	15.1	14.9	14.7	20.1
$h_{sensible}$	Btu/lbm			4.553	4.553	6.609	6.609	6.609	6.609	130.462	58.516		4.553	5.528	5.528	135.045	480.000	4.407
Chemical	(10^6 Btu/hr)											4140.036						
Sensible	(10^6 Btu/hr)			13.470	3.572	5.185	2.607	0.905	2.578	36.714	39.294		9.898	12.017	12.017	290.335	3.977	3.311
Latent	(10^6 Btu/hr)			89.048	23.613	23.613	11.872	4.124	11.742	8.471	30.312		65.434	65.434	65.434	64.712	0.000	0.000
Total Energy$^{(1)}$	(10^6 Btu/hr)			102.517	27.185	28.798	14.479	5.029	14.320	45.185	69.606	4199.544	75.332	77.451	77.451	355.047	3.977	3.311

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

Boiler efficiency for the modified system is calculated to be 90.47 percent (HHV basis). This is increased significantly as compared to the Base Case (88.13 percent) due to Oxygen firing and the addition of the Oxygen Heater and Parallel Feedwater Heaters. The net plant heat rate (HHV basis) is increased significantly to 15,188 Btu/kwhr for this option as shown in Table 3.3.2, which also includes the Base Case for comparison. The plant thermal efficiency (HHV basis) for Concept B (22.47%) is about 64 percent of the Base Case value of 35.01%. This case represents the highest efficiency of the three CO_2 removal cases studied in this project. Auxiliary power is increased to 189,709 kW as a result of the added Air Separation Unit and the CO_2 Compression and Liquefaction System. Net plant output is reduced to 273,347 kW. Carbon dioxide emissions are 53,016 lbm/hr or about 0.194 lbm/kWh which is about 9.7% of the Base Case value of 1.997 lbm/kWh.

Table 3.3. 2: Overall Plant Performance Summary for Concept B

	(units)	Original Plant (Base)	Concept B O2 Fired
Fuel Paramaters			
Coal Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4140.0
Natural Gas Heat Input (HHV)	(10^6 Btu/hr)	---	11.4
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4151.5
Steam Cycle Paramaters			
Existing Steam Turbine Generator Output	(kW)	463478	463056
CO2 Removal System Turbine Generator Output	(kW)	0	0
Total Turbine Generator Output	(kW)	463478	463056
Total Auxiliary Power	(kW)	29700	189709
Net Plant Output	(kW)	433778	273347
Overall Plant Performance Paramaters			
Net Plant Efficiency (HHV)	(fraction)	0.3501	0.2247
Net Plant Efficiency (LHV)	(fraction)	0.3666	0.2354
Normalized Efficiency (HHV; Relative to Base Case)	(fraction)	1.0000	0.6419
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9749	15188
Net Plant Heat Rate (LHV)	(Btu/kwhr)	9309	14500
Overall Plant CO_2 Emissions			
Carbon Dioxide Emissions	(lbm/hr)	866102	53016
Specific Carbon Dioxide Emissions	(lbm/kwhr)	1.997	0.194
Normalized Specific CO2 Emissions (Relative to Base Case)	(fraction)	1.000	0.097
Avoided Carbon Dioxide Emissions (as compared to Base)	(lbm/kwhr)	---	1.803
Specific Carbon Dioxide Emissions	(kg/kwhr)	0.906	0.088
Avoided Carbon Dioxide Emissions (as compared to Base)	(kg/kwhr)	---	0.818

3.3.2 Air Separation Unit

Concept B uses nearly pure oxygen instead of air for combustion of coal and therefore requires an Air Separation Unit (ASU) to provide the supply of oxygen to the furnace. The ASU selected is a cryogenic type design because of the large capacity required.

3.3.2.1 Background

As stated above in Section 3.3.1, AEP's Conesville Unit #5 requires 8,924 standard tons of 99% purity O_2 per day (T/D) when fired at 100% of the maximum continuous rating (MCR). Hence it was necessary to design an Air Separation Unit (ASU) that would be capable of producing at least this amount of oxygen. Initial discussions between ABB and air separation plant vendors were about using two 3685 T/D of oxygen trains and one 1600 T/D train for a total of 8950 T/D. This is due to the fact that a 3685 T/D is the largest plant, which could be comfortably offered at this time. After factoring in capital cost considerations, it was decided to design an overall plant comprised of two-3930 T/D trains and two-550 T/D trains, for a total capacity of 8,960 T/D. A variety of practical and technical issues are discussed below. A process flow diagram of a typical cryogenic air separation unit (BOC Webpage) showing all its major components is shown in Figure 3.3.3.

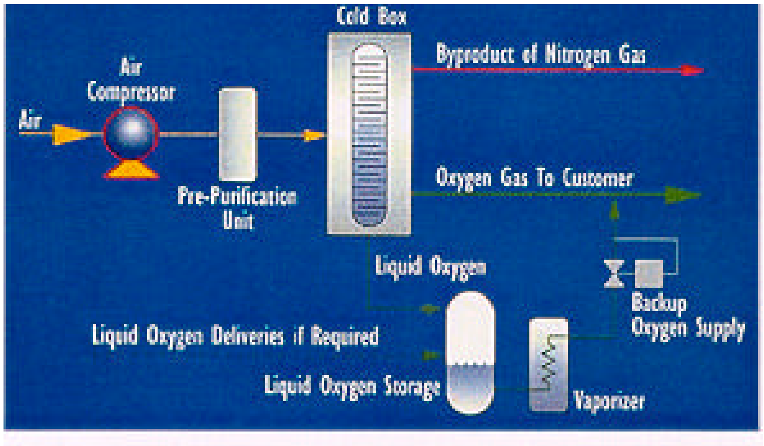


Figure 3.3. 3: Process Flow Diagram of a BOC Improved Air Separation Unit

3.3.2.2 Oxygen Purity Considerations

Oxygen purities above 98% have significant impacts on the cryogenic portion of the process. O_2 purities below 95% do not benefit in a cost reduction due to reduction in main air compressor discharge pressure requirements. The purities below 95% often require a frame size increase due to higher inlet volumetric flow rates. It also increases the size of the air adsorber system. The reduced discharge pressure requirements do save on energy of the main air compressor, but other process duties increase due to the higher flow rate of oxygen product. The overall energy consumption could increase if the product is gaseous oxygen with a significant battery limit pressure requirement.

3.3.2.3 Bottlenecks for building plants larger than 3685 ST/D

The current largest single ASU in operation is 3500 ST/D at Rozenburg, The Netherlands. However, sizes up to 4950 ST/D have been designed on paper. Usually, larger size plants are cheaper than smaller plants due to economy of scale. However, it is believed that a break point somewhere near 3685 T/D is where increased plant size will actually cost more for the following reasons:

- The compressor size, driver size, motor sizes, and line voltage drop at start-up are potential hurdles. The largest compressor sizes that ASU manufacturers have experience with are in the 40000-hp range. There may not be any significant aspects of the larger compressors that would create problems over the voltage drop to the compressor if it is motor driven, however, there is no actual experience among the major manufacturers with compressors and drivers exceeding this size.

- Columns larger than 20 feet in diameter create transportation problems for the part of the trip, which is over land. Plants located adjacent to navigable bodies of water would not have any limitation like this. The impact could translate to the need to field fabricate the column.

- Large size pipe and valves become more job specific which create the loss of pricing economies of scale which can be attained when pipe and valve sizes fit many units being built at the same time.

3.3.2.4 Major Vendors Experience

A plant size of 3685 ST/D O_2 production is larger than any vendor had operating as of May 2000. The following list represents the largest operating plant size that the three vendors had running at that time:

Vendor	Largest size operating plant
	(ST/D O_2)
Air Liquide	3190
Air Products	3500
BOC	2640

3.3.2.5 Power Consumption

The Air Separation unit (ASU) includes a cryogenic plant for air separation. Economic considerations for this application favored the selection of oxygen stream purity of 99 percent by weight. As stated above, four trains were required to produce the required oxygen mass flow rate of about 8924 tons per day. This system consumes 95,822 kW of electric power or about 21 percent of the generator output. This energy consumption is equivalent to about 260 kWh/ton of O_2 produced. By contrast, Kobayashi and Prasad (Kobayashi and Prasad, 1999) state that the thermodynamic minimum energy requirement is about 40 kWh/ton of O_2 produced. If this limit could be approached, it would represent a breakthrough technology that would enable Concept B to become nearly cost competitive with "business as usual" (coal-fired in air without CO_2 capture) electricity generation.

3.3.2.6 Plot Plan Requirements

The required plot plan areas, supplied by Air Products, are 185 feet by 250 feet and 125 feet by 150 feet for the 3930 T/D and 550 T/D plants, respectively. When the areas for storage and oxygen compressions are included, the required total plot plan amounts to 555 feet by 620 feet. Refer to Appendix I for the plant layout drawings for Concept B.

3.3.2.7 Investment and Operation and Maintenance Costs

The investment, and operation and maintenance costs of the Air Separation Unit are presented in Section 5.3.2.

3.3.3 Boiler Heat Transfer Analysis

The primary objective of the systems analysis task for Concept B was to develop a system, which would produce high carbon dioxide content flue gas from an existing coal, fired boiler without requiring major pressure part modifications to the boiler. In order to access whether pressure part modifications would be necessary an accurate heat transfer analysis of the existing boiler was required.

The first step was to set up a steady state performance model of the Conesville #5 steam generator unit. This involved calculating or obtaining all the geometric information for the unit as required by the proprietary Reheat Boiler Program (RHBP). The RHBP provides an integrated performance model of the boiler island including, in addition to the steam generator unit, pulverizers, air heater, and steam temperature control logic. The RHBP is used to size components and/or predict performance of existing components. In this study, since the boiler island component sizes are known, the RHBP was used exclusively for calculating unit performance.

The next step in the heat transfer analysis of the system was to calibrate the RHBP model of the unit. This involved obtaining a set of test data (with air firing) for the existing unit and adjusting the performance model to match the test data. The required test data includes steam temperatures entering and leaving each major heat exchanger section in the unit, steam pressures, coal analysis, flue gas oxygen content, etc. The adjustments to the model are in the form of "surface effectiveness factors" and "fouling factors".

After the model was calibrated, additional adjustments were required in order to obtain an accurate heat transfer analysis with the high carbon dioxide content flue gas of the Concept B system. The combustion process occurs in a non-conventional environment, which produces gases of different physical and thermal properties. These gas property differences cause significant differences in the heat transfer processes, which occur within the steam generator unit. Analyses were made to determine the impact of the heat transfer differences on boiler behavior.

The RHBP accounts for three modes of heat transfer in the upper furnace and convective pass of the unit (direct radiation, non-luminous radiation and convection). The direct radiation is emitted from the furnace "fire-ball" and is absorbed in various areas of the unit depending on the geometry of the tube banks and the proximity to the "fire-ball". Since the distribution of direct radiation is only a function of geometry, no formulation modifications were necessary for this component of the heat transfer analysis. Investigation of the non-luminous radiation formulations within the RHBP indicated that current equations (based on the Hottel curves, 1957) would be accurate and formulation modifications would not be required. The convection formulations, however, were not set up with the capability of accurately analyzing convective heat transfer for flue gases which were this much different than the typical range of boiler flue gases. The appropriate corrections were made to properly model the convection process in the RHBP with oxygen firing.

3.3.3.1 Furnace Analysis

Initially, it was expected that heat fluxes to the furnace walls might be increased or decreased somewhat for Concept B (oxygen firing) due to the higher CO_2 and H_2O content of the flue gas relative to the air fired Base Case for the following reasons:

- Higher gas emissivity and absorptivity of CO_2 and H_2O relative to N_2. For an equivalent local gas temperature, the higher emissivity of CO_2 and H_2O, relative to the optically transparent N_2, should serve to increase the absorption coefficient and hence, the radiation to the walls in the recycled flue gas case (Concept-B) relative to the Base Case.
- Higher specific heat of CO_2 and H_2O relative to N_2. For an equivalent local heat release and heat transfer, the higher specific heat of carbon dioxide will serve to decrease the overall flame temperatures of the recycled flue gas mixture in the Concept-B case, relative to the baseline case. This would tend to compensate for the higher gas emissivity.

In preparation for the Concept B furnace performance analysis, a review of pertinent literature and CFD furnace analysis results developed for this project (Bozzuto, et al., 2000) were completed. Explanations for the observed trends must rely on physical property differences between nitrogen, which dominates in the Base Case, and carbon dioxide, which largely supplants nitrogen in Concept-B.

Literature review:

The first step in the furnace analysis was to review applicable results obtained from the literature regarding combustion of coal in an environment where CO_2 displaces the N_2 in the combustion air. Several investigators (Thambimuthu, 1998; Kiga, et al., 1997; and Weller, at al., 1985) have found from pilot-scale testing that when coal is burned in O_2/CO_2 environment whereby CO_2 displaces the N_2 in the combustion air (i.e., in ~30% O_2/70% CO_2 mixture, by volume), the heat absorption in the lower furnace is not significantly impacted.

To illustrate this point, the results of re-analyses of Thambimuthu's (Thambimuthu, 1998) data concerning the combustion of a subbituminous coal sample from Western Canada are presented in Figure 3.3.4. As can be seen in this figure, the radiative heat fluxes into the lower furnace were, for the three cases studied (air, 28% O_2/72% CO_2, and 35%O_2/65% CO_2), roughly similar at gas temperatures greater than 1275 ^{o}C.

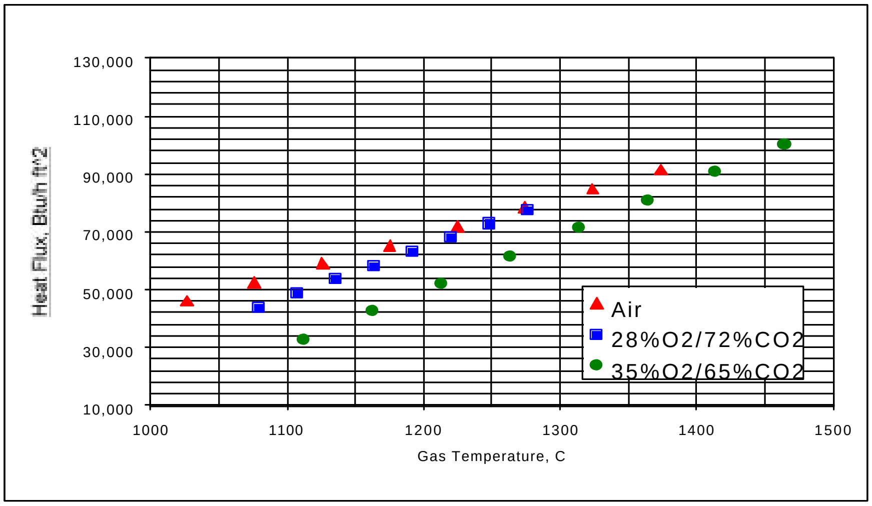


Figure 3.3. 4: Variation of Radiative Heat Flux with Gas Temperature During Combustion in Air and O_2/CO_2 Mixtures (from Thambimuthu, 1998)

Results from a similar experiment, Chui, et al., (2001), were also investigated, in somewhat more detail, in order to assess the apparent overall change in gas emissive power that would occur in a furnace when switching from air-firing to a high CO_2-O_2 environment. The test facility, the same one used previously by Thambimuthu et.al.(1998), is a pilot-scale, 0.3 MW facility and consists of a cylindrical, down-fired vertical combustor (0.61 m I.D. and 6.7 m in length). The most relevant tests from the literature source are designated as Case-1 (baseline condition with air firing and a burner swirl setting of 11) and Case-2 (dry recycle with burner swirl setting of 10). The dry recycle case is representative of an oxygen fired case similar to Concept B with 28% O_2 / 72% CO_2 by volume. The principal conclusion of the work by Chui, et al., based on both experimental and simulated results, was that the dry recycle case demonstrated a distinct decrease in the incident radiant heat flux to the wall (by about 18%) relative to the baseline case (Case-1).

The intent of the present re-analysis of this data was to utilize the incident heat flux measurements and measured centerline gas temperatures for the cases reported by Chui, et al., and then back-calculate the apparent gas emissivity for the furnace as a function of the longitudinal position. The desire was to verify whether the apparent emissivity differences between the baseline and dry recycle cases mirrored the known physical differences in radiative absorption properties between N_2 and CO_2 (corresponding to the baseline and dry recycle cases, respectively). The comparison of Case-1 and Case-2, which have similar swirl settings (and thus similar flow patterns), should isolate the effects of gas emissive properties without introducing the strong effects of flow pattern changes as observed by Chui and reconfirmed by our analysis. Based on the incident heat flux measurements and the measured centerline gas temperatures, apparent gas emissivities for the furnace were calculated. These values are tabulated in Table 3.3.3 and plotted in Figure 3.3.5.

Table 3.3. 3: Apparent Combustor Gas Emissivities

Axial Distance (m)	Case-1 (Air Firing) Gas Emissivity	Case-2 (Oxy-Firing) Gas Emissivity
0.95	0.685	0.782
1.55	0.657	0.737
2.15	0.611	0.796
2.77	0.562	0.829
3.40	0.532	0.766

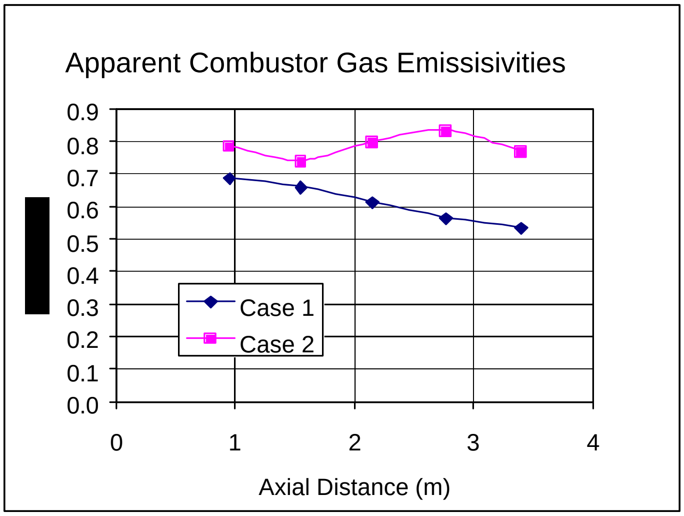


Figure 3.3. 5: Apparent Combustor Gas Emissivities as a Function of Axial Distance

As shown in Figure 3.3.5, the apparent emissivity of the dry recycle case (Case-2) is higher than that of the baseline case (Case-1). Based on the substitution of the radiating gas CO_2 for non-radiating nitrogen in the dry recycle case, the observed shift is what would be expected. It should be understood, however, that these results are only to be expected for cases with identical or similar flow patterns. The other test cases reported in Chui, et al. (2001) with swirl settings and corresponding aerodynamic and flame patterns that were different from those of Case-1 and Case-2, showed that such flow patterns have a dominant influence on the apparent gas radiative properties.

Another interesting aspect of the combustor results are the measured centerline gas temperatures. Table 3.3.4 tabulates the corresponding temperatures measured in the Chui, et al. combustor, along with the Case-1 to Case-2 shift. The temperature shifts are plotted in Figure 3.3.6. The mean gas temperature shift, averaged over all measurement locations, is -214 °F.

Table 3.3. 4: Gas Temperature Profiles

Axial Distance (m)	Case-1 Centerline Gas Temperature, °F	Case-2 Centerline Gas Temperature, °F	Temperature Shift, °F (Case 2 - Case 1)
0.95	2664	2419	-245
1.55	2552	2367	-185
2.15	2421	2215	-205
2.77	2284	2041	-243
3.40	2143	1953	-191

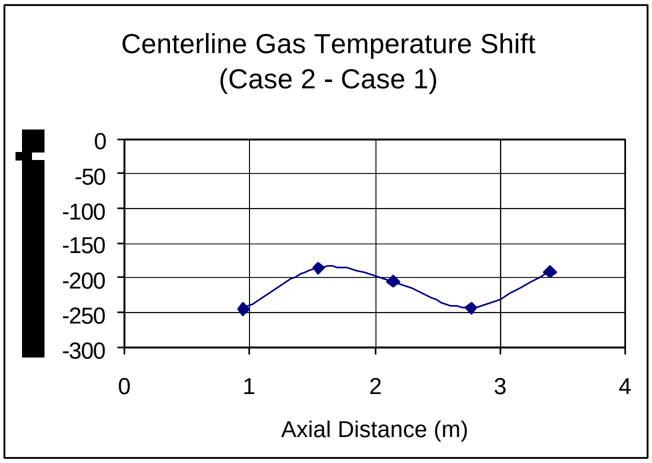


Figure 3.3. 6: Centerline Gas Temperature Shift as a Function of Axial Distance

Calculations of adiabatic flame temperature (T_{ad}) were performed for coal firing in AEP's Conesville Unit #5 at maximum continuous rating (MCR). A number of scenarios were considered, namely: (1) Base Case (i.e., coal firing in air); (2) Constant Mass Case (i.e., the mass of nitrogen in air was replaced by the mass of CO_2 in the flue gas recycle (FGR); and (3) Various CO_2/O_2 Mole Ratio Cases. Results (Figure 3.3.7) indicated that the T_{ad} for air firing was approximately 180 ^{o}F higher than for the Constant Mass Case. This difference is due to differences in specific heat (Cp) between N_2 and CO_2 (i.e., CO_2 Cp > N_2 Cp at elevated temperatures, Figure 3.3.8). One implication is that the ultimate combustion efficiency difference between air firing and O_2/FGR firing could be governed by opposing phenomena, i.e., potentially lower reaction rate, but longer residence time in O_2/FGR than in air. Thermo-gravimetric analysis (TGA) and Drop Tube Furnace System-1 (DTFS-1) results -- obtained from combustion firing of both Conesville Unit #5 and Pittsburgh Seam #8 coals other (See Volume II Report of this work, Bozzuto, et al., 2001) -- indicate that these two phenomena indeed virtually cancel each other.

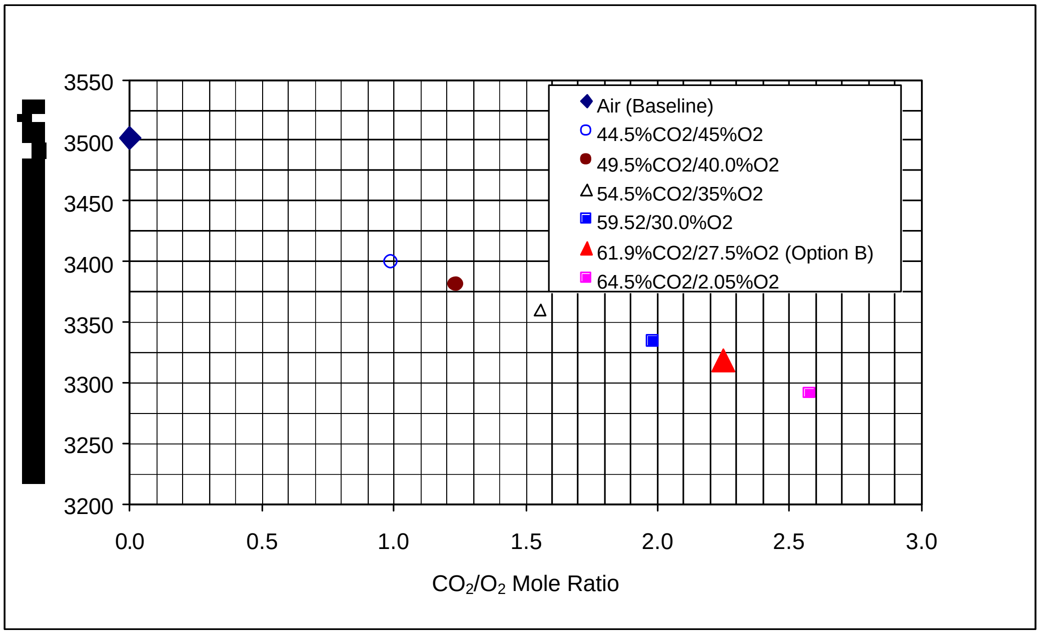


Figure 3.3. 7: **Calculated Adiabatic Flame Temperature as a Function of CO2/O2 Mole Ratio for Conevsville Unit #5**

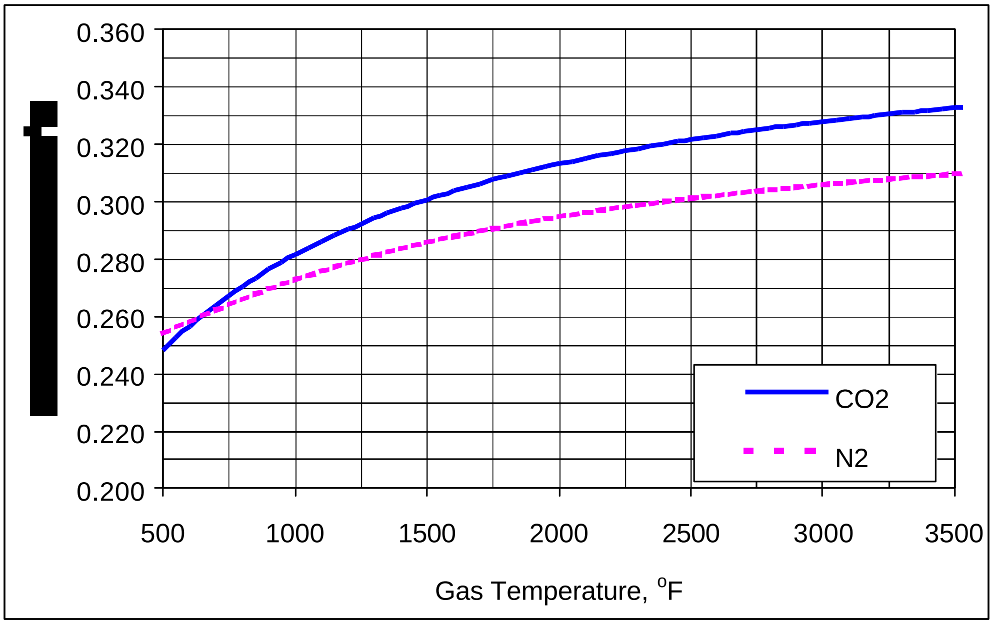


Figure 3.3. 8: **Specific Heats (Cp) of Nitrogen and Carbon Dioxide as a Function of Temperature**

CFD Furnace Model Results and Conclusions:

Review of the CFD analysis results (Bozzuto, et al., 2001) which were developed for this project indicated the following with respect to furnace heat transfer:

The baseline case (or Base Case) exhibits a slightly higher peak gas temperature (maximum difference of about 200 °F), and a correspondingly higher average (cross-sectional) gas temperature (difference of 90 °F at the HFOT), than that of the Concept-B case. This result compares favorably with the approximate -214 °F shift found by Chui et.al.

Note, however, that this close correspondence may be somewhat fortuitous, particularly since the two furnaces are so different. Nevertheless, the correspondence may indicate that the baseline air case may be expected to exhibit consistently higher peak and averaged post-flame temperatures than an oxy-fired case (with similar flow and flame patterns).

The net wall absorption in the furnace region for the baseline case is larger (by less than 1%) than that of the Concept-B case. Conversely, the net wall absorption for the baseline calculation with the proprietary in house Reheat Boiler Program (RHBP), as described below, was found to be about 6% lower than the Concept-B case calculated by the RHBP. The higher specific heat of the carbon dioxide, and the associated lower gas temperatures in the Concept-B case, will tend to offset somewhat or compensate for the expected increase in the wall absorption (i.e., anticipated due to the enhanced emissivity of the CO_2.) Furthermore, the spatial distribution of the carbon dioxide across the cross-section may be important; it is hypothesized that high concentrations of CO_2 in the cooler gas mixtures between the wall and the flame may actually act to inhibit the net radiative flux to the walls.

Performance Model (RHBP) Results:

The present study for Concept B involves coal combustion in a medium with ~31%O_2/69% CO_2 volume ratio. Heat transfer in the lower and upper furnace regions as calculated by the RHBP is compared in Figure 3.3.9. This figure compares heat fluxes (Btu/hr-ft2) in selected lower and upper furnace regions for air firing and oxygen firing. Lower furnace region results show firing zone heat flux to be about 11 percent higher with oxygen firing. Upper furnace region results show the reheat radiant wall is about 6 percent higher and the superheat division panels are about 13 percent higher with oxygen firing. Similarly, the upper furnace waterwall area is about 10 percent higher. The overall lower furnace heat absorption for Concept B was about 6% higher than for the air fired Base Case. The horizontal furnace outlet temperature (HFOT) for Concept B was calculated to be 123 °F lower than for the Base Case.

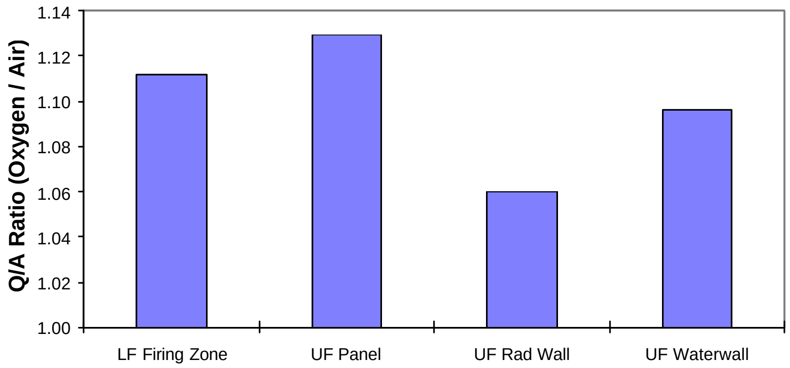


Figure 3.3. 9: Furnace Region Heat Flux Comparison

The trend calculated by the CFD code, with respect to the furnace temperature at the HFOP, appears to be qualitatively aligned with the trend calculated by the in-house RHBP code. However, the difference in wall absorption for the Concept-B case relative to the baseline case, calculated by FLUENT (change of –1%), does not fully agree with the results of the RHBP (change of +6%). Computed and/or experimental trends that indicate that oxy-firing may induce either an increase in wall absorption or a decrease in wall absorption, relative to the baseline case, are supported in the literature. It is suggested that the absorptive properties of a local CO_2-containing gas mixture in the vicinity of the wall may either act to enhance or inhibit the incident radiative flux to the wall, depending upon its relative spatial location and mixture temperature. For example, a low-temperature, CO_2-rich mixture near the wall would absorb the radiative flux from the interior “fireball” region; a portion of the energy would be re-emitted at the lower gas temperature, and the remainder of the energy would be used to heat up the gas. The result would be a net reduction in the incident radiative flux to the wall. Conversely, a high-temperature, CO_2-rich mixture near the wall would emit at its local gas temperature, thus enhancing the incident radiative flux to the wall. Therefore, it is suggested that the relative spatial relationships of the cool and hot gas mixtures, the relative composition of the absorbing media, and the proximity of that media to the wall (as induced by the furnace and firing system aerodynamics), may significantly impact the gas-phase irradiation to the wall. Indeed, the aerodynamic and flame patterns that prevail in a given air-fired or oxy-fired furnace may conceivably have as strong an impact on the net wall absorption as the physical property differences themselves (gas-phase emissivity, specific heat, etc.)

The results of computational tools (such as CFD and the RHBP) should be viewed with some skepticism, particularly when they are asked to produce calculations outside of the experience base for which they have been validated. Certainly, the gas-phase and particle-phase radiation property sub-models could be enhanced in both codes, and the impact of spatial gradients (caused by the interleaving of hot and cool regions of varying CO_2 composition) on radiative wall flux needs to be investigated more fully. CFD computations have been utilized to gain a more in-depth qualitative understanding of aerodynamic and flame patterns in combustion systems, but have not yet been integrated into routine boiler design procedures. At the present time, the ALSTOM design standards are based upon the experience and expertise built into the RHBP and other design protocols, and those standards must be adhered to in any new design project in order to mitigate risk. Therefore, the results of the RHBP, as reported in this work, must be regarded as the current standard, both from an engineering experience viewpoint and from a rules-based design viewpoint. However, additional validation work needs to be done in order to confirm the trends for oxy-firing scenarios.

Although differences in physical properties of the transport gases (e.g., air-firing versus oxy-firing) may initially induce undesirable deviations from the performance goals of an existing field unit, the designer and engineer views such alterations as challenges rather than insurmountable obstacles. Much can be done in the way of operating condition optimization (e.g., spray, nozzle tilt, and amount of gas recirculation), without major modifications to the unit, to realign steam temperatures and wall absorption with their desired, target values. Design parameters are often in conflict, and the intelligent designer must work to balance various parameters and operating conditions to achieve the desired outcome. The CFD work performed here was based on certain constraints to maintain equivalency for comparison purposes, which would not necessarily be the case in the field. For example, a potential decrease in furnace wall absorption of several percentage points could be nullified through manipulation of the various operating parameters available.

3.3.3.2 Convection Pass Analysis

Figures 3.3.10, 3.3.11 and 3.3.12 show the comparison of convective, non-luminous, and total heat transfer rates between air firing and oxygen firing for all the major sections contained in the existing convection pass of the unit at full load operating conditions as calculated by the RHBP.

Convective heat transfer in utility steam generator units is dependent upon many of the transport properties of the flue gas (viscosity, thermal conductivity, density, specific heat and others). Additionally, convection depends strongly on gas velocity. With the Concept B system there are significant changes in the flue gas analysis as compared with air firing. These gas analysis changes cause both transport property and gas velocity changes throughout the unit. The resulting convective heat transfer rates, as shown in Figure 3.3.10, ranged from about 80 to 105 percent as compared to air firing.

Significant differences in non-luminous radiant heat transfer are also expected. Of the gases produced by the complete combustion of a fuel, only carbon dioxide, water vapor and sulfur dioxide emit radiation over a sufficiently wide band of wave lengths to warrant consideration. With this system the primary change in the flue gas as compared to air firing is the large increase in the CO_2 content and decrease in N_2 content. The resulting enhancement in non-luminous heat transfer rates range from about 26 to 28 percent greater than air firing as shown in Figure 3.3.11.

Ultimately the convection pass total heat transfer rates, shown in Figure 3.3.12, were increased in the range of 1 to 8 percent over the values with air firing.

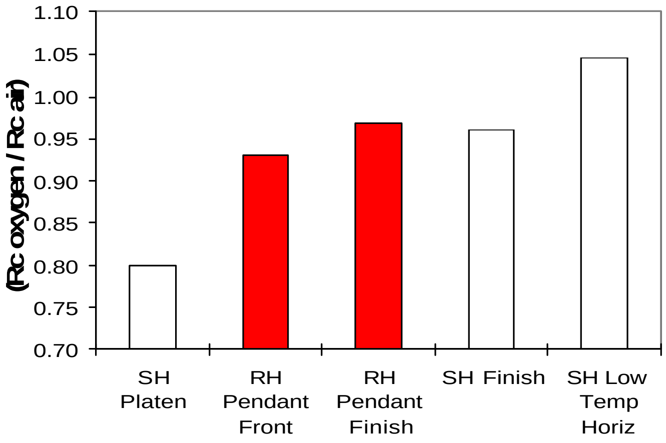


Figure 3.3. 10: Convective Heat Transfer Rate Comparison

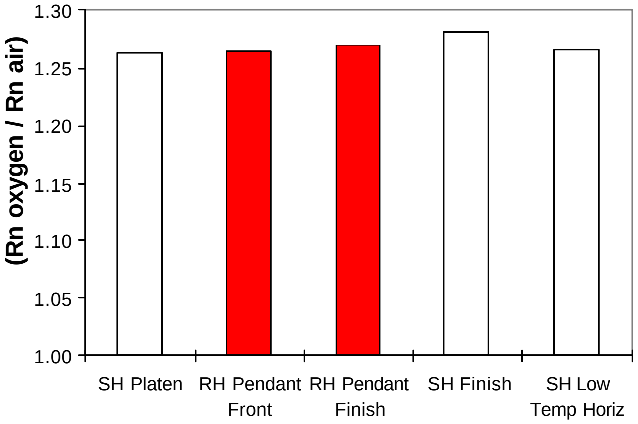


Figure 3.3. 11: Non-luminous Heat Transfer Rate Comparison

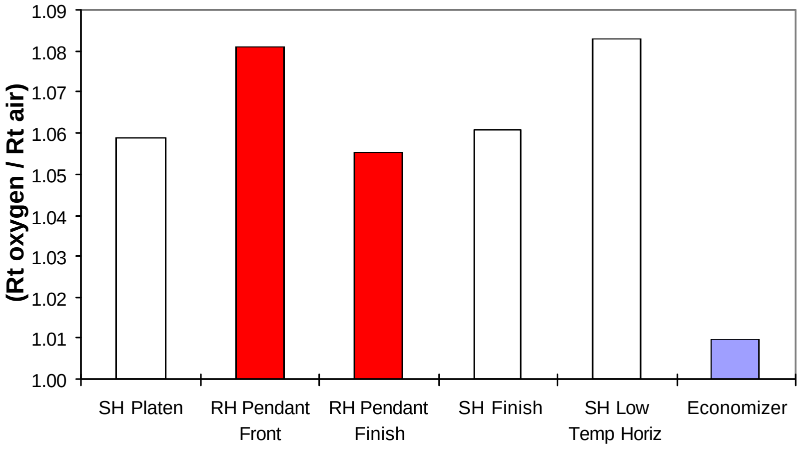


Figure 3.3. 12: Total Heat Transfer Rate Comparison

Steam temperature control was achieved through the use of burner tilt and reheat spray. The performance analysis results indicated the reheater circuit required about 1.45 percent spray to maintain the reheat outlet temperature at the design value. The superheater circuit required about 0.34 percent spray to maintain the superheat outlet temperature at the design value. The burner tilt was set at –10 degrees, the minimum value the customer uses. The overall steam conditions produced are shown in Table 3.3.5 below. Furthermore, with this recycle gas system, flue gas recirculation is also available as an additional steam temperature control variable. In this study however, we did not utilize this as an additional steam temperature control variable. We decided, as a first approximation, to set the flue gas recirculation rate at a value such that the flue gas flow to coal flow ratio was the same as with air firing. As it turned out, this was an acceptable setting for the gas recirculation rate.

Table 3.3. 5: Boiler/Turbine Steam Flows and Conditions for Concept B

	SHO	**FWI**	**ECO**	**RHO**	**RHI**
Flow (lbm/hr)	**3131619**	3131619	3131619	2808511	2808511
Pressure (psia)	**2535**	3165	3070	**590.8**	**656.5**
Temperature (F)	1000	493	622	1000	631
Enthalpy (Btu/lbm)	**1456.3**	**479.2**	**640.8**	1517.6	1305.6
			Qmain steam	(10^6 Btu/hr)	3059.9
			Qreheat	(10^6 Btu/hr)	595.4
			Q Total s/w	(10^6 Btu/hr)	3655.3

Notes:
SHO = Superheater Outlet; FWI = Feedwater Inlet; ECO = Economizer Outlet; RHO = Reheater Outlet; RHI = Reheater Inlet

With the increased heat transfer rates with oxygen firing and similar steam temperature profiles, there was concern regarding metal temperatures throughout the unit. A detailed analysis using the Metal Temperature Program (MTP) was however, beyond the scope of this study. The MTP, using thermal inputs from the RHBP, calculates steam and metal temperatures at any selected point along the length of the tube. All tubes of each heat exchanger bank are modeled. A very brief review of metal temperatures at only a few selected points was done in this study. In general, for the points investigated, the metal temperatures were found to be the same or slightly lower than with air firing. The primary reason for this result was that although the heat transfer rates were slightly higher and the steam temperature profile was similar, the gas temperatures were lower. This combination yields similar heat flux conditions and ultimately similar metal temperatures.

3.3.4 Boiler System Modifications

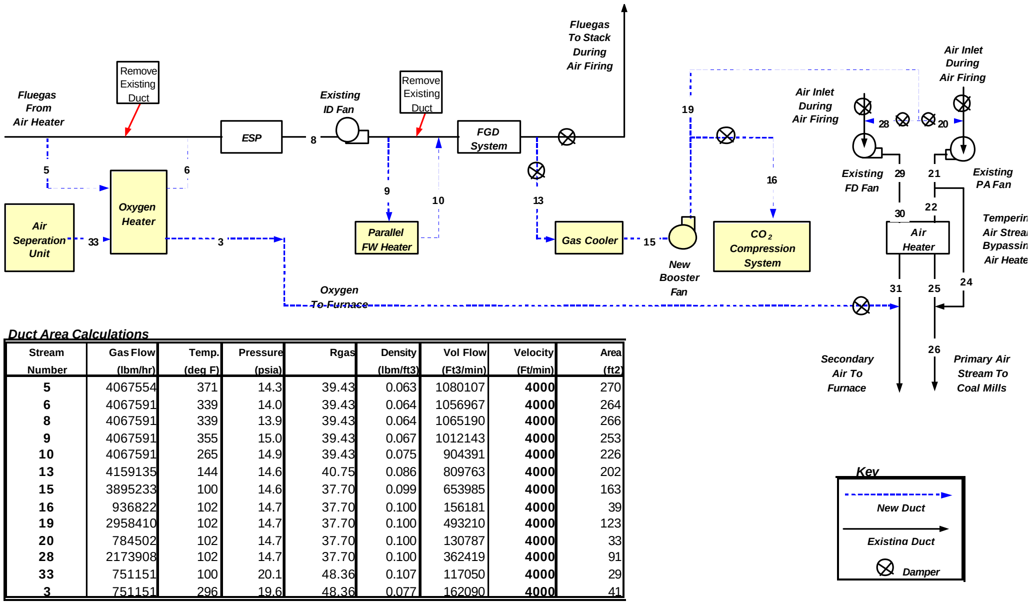
Listed below are the basic modifications required for the Conesville No.5 unit to support firing with oxygen as required in Concept B.

3.3.4.1 Boiler

The Boiler Island should be inspected for potential air leaks into the system and should be sealed to minimize any infiltration. Special attention should be given to all penetrations including seal boxes for convective surfaces, sootblowers, wallblowers, expansion joints, ductwork, fuel piping, fans and windbox.

3.3.4.2 Ducts and Dampers and Fans

Refer to the Duct and Damper Diagram (Figure 3.3.13) shown below for the required duct arrangement and associated cross-sectional areas for this system. New flue gas ductwork is required for the Oxygen Heater. Part of the existing ductwork from the Air Heater flue gas exit to the Electrostatic Precipitator inlet must be removed and replaced with new ductwork that accommodates the new Oxygen Heater. Additionally, oxygen supply ductwork from the ASU to the Oxygen Heater is required and hot oxygen supply ductwork from the Oxygen Heater to the existing secondary "air" duct must be provided. Similarly, New flue gas ductwork is required for the Parallel Feedwater Heater. Part of the existing ductwork from the Induced Draft fan outlet to the Scrubber inlet must be removed and replaced with new ductwork that accommodates the new Parallel Feedwater Heater. New ductwork and dampers are also required for the recycle flue gas streams that feed the existing Forced Draft and Primary Air fans.



Duct Area Calculations

Stream Number	Gas Flow (lbm/hr)	Temp. (deg F)	Pressure (psia)	Rgas	Density (lbm/ft3)	Vol Flow (Ft3/min)	Velocity (Ft/min)	Area (ft2)
5	4067554	371	14.3	39.43	0.063	1080107	**4000**	270
6	4067591	339	14.0	39.43	0.064	1056967	**4000**	264
8	4067591	339	13.9	39.43	0.064	1065190	**4000**	266
9	4067591	355	15.0	39.43	0.067	1012143	**4000**	253
10	4067591	265	14.9	39.43	0.075	904391	**4000**	226
13	4159135	144	14.6	40.75	0.086	809763	**4000**	202
15	3895233	100	14.6	37.70	0.099	653985	**4000**	163
16	936822	102	14.7	37.70	0.100	156181	**4000**	39
19	2958410	102	14.7	37.70	0.100	493210	**4000**	123
20	784502	102	14.7	37.70	0.100	130787	**4000**	33
28	2173908	102	14.7	37.70	0.100	362419	**4000**	91
33	751151	100	20.1	48.36	0.107	117050	**4000**	29
3	751151	296	19.6	48.36	0.077	162090	**4000**	41

Figure 3.3. 13: Duct and Damper Diagram for Concept B

A new booster fan will be required as a minimum. The booster fan is required to overcome the added draft loss of the various new heat exchanger components shown in Figure 3.3.1 previously. Specifications for the booster fan are shown in Table 3.3.6 below. The existing Induced Draft, Forced Draft and Primary Air fans should eventually be checked with the suppliers to see if they can handle the new conditions (Note: The molecular weight is now 37.7 for the FD and PA fans and 39.4 for the ID fan). For this study we have assumed that both fans are acceptable.

Table 3.3.6: Booster Fan Specification

Booster Fan Specification		
Gas Analysis		
Oxygen	(wt percent)	2.63
Nitrogen	"	3.97
Water Vapor	"	2.87
Carbon Dioxide	"	90.51
Sulfur Dioxide	"	0.03
Total	"	100.00
Mass Flow Rate	(lbm/hr)	3895233
Gas Inlet Temperature	(Deg F)	100.0
Inlet Pressure	(psia)	14.56
Outlet Pressure	(psia)	14.70
Pressure Rise	(in wg)	4.00

3.3.4.3 Oxygen Heater

The flue gas exiting the Air Heater is at 371 °F. A large quantity of sensible heat is still available in this gas stream. The Oxygen heater is used to recover additional sensible heat in the flue gas as a result of reduced air heater performance with oxygen firing. This heat is partially recovered in the Oxygen Heater where cold O2 from the Air

Separation Unit is preheated before being mixed with the gas recirculation stream prior to furnace injection. Part of the existing ductwork connecting the existing Air Heater flue gas exit stream to the Precipitator inlet is removed. New ductwork, which accommodates the addition of the new Oxygen Heater, is installed. Performance requirements and gas analyses for the Oxygen Heater are shown in Table 3.3.7. Table 3.3.8 shows the Oxygen Heater specifications. It should be noted that the flue gas stream, which provides the heat source for this heat exchanger, contains flyash, as the Oxygen Heater is located between the Air Heater and the Electrostatic Precipitator (ESP). A simple sketch of the oxygen heater, which is similar in design to a tubular air heater, is shown in Figure 3.3.14. Refer to Appendix I Concept B Boiler modification drawings for the proposed arrangement of this equipment.

Table 3.3. 7: Oxygen Heater Performance Requirements and Gas Analysis

Oxygen Heater Performance Requirements	(units)	Value
Flue Gas Mass Flow	(lbm/hr)	4067591
Flue Gas Inlet Temperature	(deg F)	371
Flue Gas Outlet Temperature	(deg F)	339
Oxygen Mass Flow	(lbm/hr)	751151
Oxygen Inlet Temperature	(deg F)	100
Oxygen Outlet Temperature	(deg F)	296
Oxygen Heater Heat Transfer	(10^6 Btu/hr)	33.062
Oxygen Pressure	(psia)	20
Flue Gas Pressure	(psia)	14.281

Gas Analysis Constituant	(units)	Oxygen Stream	Fluegas Stream
O_2	(Wt Frac)	0.9900	0.0252
N_2	"	0.0100	0.0380
H_2O	"		0.0647
CO_2	"		0.8670
SO_2	"		0.0051
Total		1.0000	1.0000

Table 3.3. 8: Oxygen Heater Specifications

Oxygen Heater Specification		
Duct Height	ft	20
Duct Width	(ft)	44.33
Tube Outside Diameter	(Inches)	2
Tube Spacings St and Sl	(Inches)	3
Tubing Thickness	(inches)	0.083
Number of tubes wide	(no.)	177
Number of tubes deep	(no.)	19
Height of tubes	(ft)	20
Total Number of Tubes	(no.)	3363
Total Lineal length of tubin	(ft)	67260
Tubing Outside Surface	(ft2)	35217
Total Weight of tubing	(tons)	61.0
Tubing type		Bare tubes
Tubing Material		SA-178

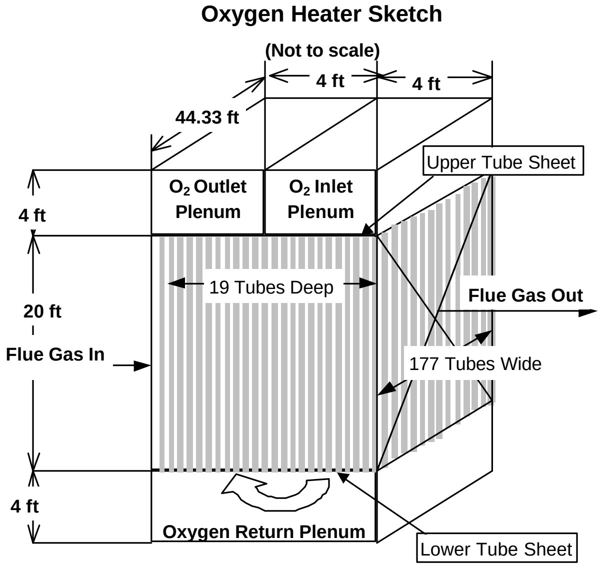


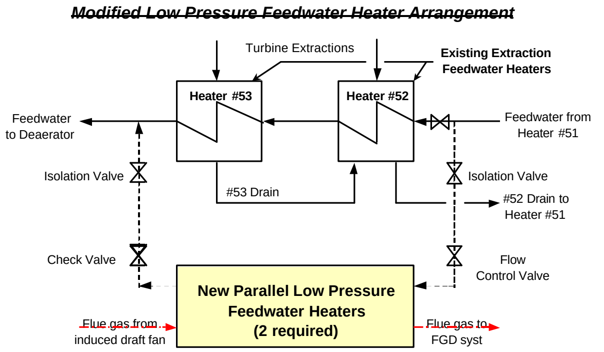
Figure 3.3. 14: Oxygen Heater Sketch

3.3.4.4 Parallel Low Pressure Feedwater Heater

A parallel low-pressure feedwater heater (PFWH) is facilitated to remove remaining useful sensible heat in the flue gas. This heater is installed in a parallel feedwater stream with the existing extraction low-pressure feedwater heaters number 52 and 53. The heat source for the PFWH is flue gas leaving the ID fan as shown on the system diagram (Figure 3.3.1). In actuality, there are two ID fans on this unit and therefore two PFWH's are used. The flue gas temperature leaving the induced draft fans is 355 F. The PFWH cools the flue gases down to 265 F by heating feedwater in a parallel stream with the feedwater being heated in the extraction feedwater heaters No. 52 and 53 as shown in Figure 3.3.15. About 29 percent of the total low-pressure feedwater flow leaving the No. 51 low-pressure feedwater heater is heated in the new PFWH. This heat exchanger is essentially a low temperature economizer section that is designed similar to Heat Recovery Steam Generator (HRSG) tube banks with finned tubing. The fin pitch can be quite high (6 fins/inch) since the flue gas at this point is particulate free.

These new components recover about 26,458 kW [90.3 x 106 Btu/hr]. This causes the steam extractions to heaters No. 52 and No. 53 to be reduced proportionally and the steam flow through the low-pressure section of the existing steam turbine to be increased by approximately two percent. This increases the generator output proportionally. Performance requirements and specifications for the PFWH are shown in Tables 3.3.9 and 3.3.10. Figure 3.3.16 shows a simple sketch of the new component. Refer to Appendix I Concept B Boiler modification drawings for the proposed arrangement of this equipment.

Table 3.3. 9: Parallel Feedwater Heater Performance Requirements



Parallel Feedwater Heater Perf. Requirements

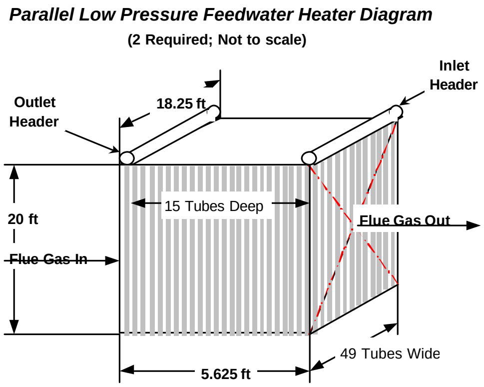
Flue Gas Mass Flow	(lbm/hr)	2033795
Flue Gas Inlet Temperature	(deg F)	355
Flue Gas Outlet Temperature	(deg F)	265
Feedwater Flow Rate	(lbm/hr)	369724
Feedwater Inlet Temperature	(deg F)	184.9
Feedwater Outlet Temperatu	(deg F)	305.3
PFWH Heat Transfer	(10^6 Btu/hr)	45.1
Feedwater Pressure	(psia)	210
Flue Gas Pressure	(psia)	14.95

Key

New Pipe	- - - - -
Existing Pipe	———
Flue gas	–·–·–►

Figure 3.3. 15: Parallel Feedwater Heater Arrangement

Table 3.3. 10: Parallel Feedwater Heater Specification



Parallel Feedwater Heater Specification

Duct Height	(ft)	20
Duct Width	(ft)	18.25
Tube Outside Diameter	(Inches)	**1.5**
Tube Spacings St and Sl	(Inches)	4.5
Tubing Thickness	(inches)	**0.095**
Number of tubes wide	(no.)	49
Number of tubes deep	(no.)	15
Height of tubes	(ft)	20
Total Number of Tubes	(no.)	735
Total Lineal length of tubing	(ft)	14700
Outside Surface Area	(ft2)	72177
Total Weight of tubing	(tons)	40.0
Tubing type		**finned**
Fins per inch		**5.2**
Fin Height	(Inches)	**0.75**
Fin Thickness	(Inches)	**0.05**
Fin Material		**Carbon Steel**
Tubing Material		**SA-178A**

Figure 3.3. 16: Parallel Feedwater Heater Sketch

Table 3.3.11 shows the specifications for the PFWH headers and piping. The piping runs from the feedwater pipe entering existing feedwater heater #52 to the two PFWH unit inlet headers and returns from the to PFWH outlet headers to the feedwater pipe leaving existing feedwater heater #53.

Table 3.3. 11: PFWH Header and Piping Specifications

	Outside Dia.	Nominal ID	Material	Weight	Length	Connections
	(inches)	(inches)		(lbm/ft)	(ft)	(number)
Headers	8.625	6.251	SA-106C	53.41	18.25	49-1.5" ; 1-6.625"
Piping	6.625	SCH 120	ASA B36.1	36.3	375 ea. 2 reqd	

3.3.4.5 Controls and Instrumentation:

Additional controls and instrumentation will be required for the new components and systems described. The transition between air firing and oxygen firing needs careful consideration.

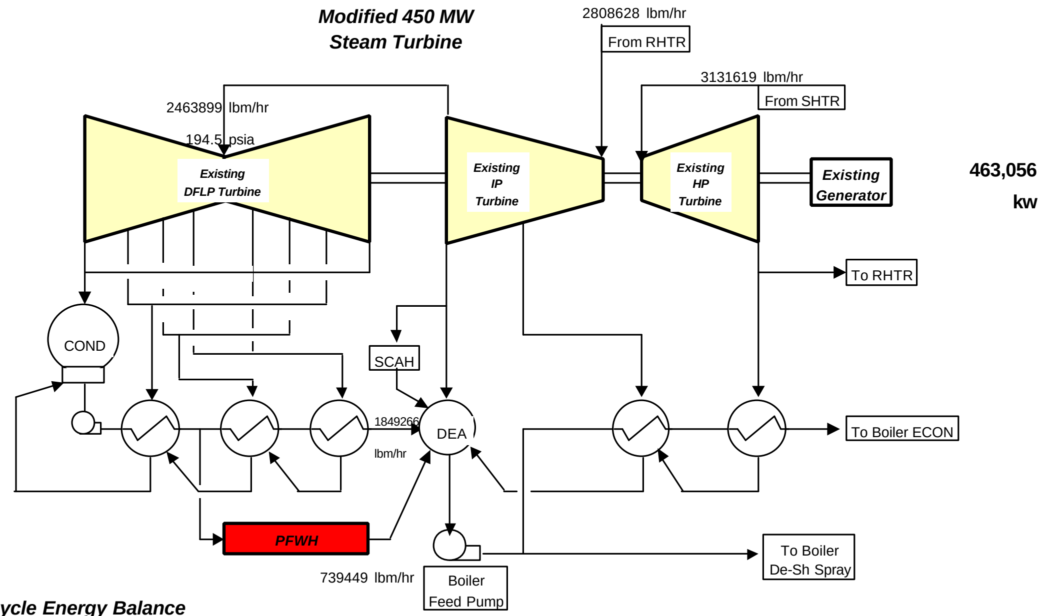
3.3.4.6 Modified Boiler Drawings

Drawings of the modified boiler showing the existing and new equipment are listed below and contained in Appendix I.

U00-E-0551R General Arrangement - Side Elevation for Concept B
U00-E-0552R General Arrangement - Plan View "B-B" for Concept B
U00-E-0585R General Arrangement - Plan View "A-A" for Concept B

3.3.5 Steam Cycle System

The steam cycle system for Concept B is modified slightly with the addition of a low-pressure feedwater heater arrangement in parallel with extraction feedwater heaters # 52 and 53 as described in Section 3.3.4.4. The parallel feedwater heaters are used to recover additional sensible heat in the flue gas as a result of reduced air heater performance with oxygen firing. The modified steam cycle is shown in Figure 3.3.17. The associated Mollier diagram showing the modified steam cycle on enthalpy vs entropy coordinates is shown in Figure 3.3.18. The modified steam cycle system produces 463,056 kw with a steam turbine heat rate of 8089 Btu/kwhr.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Power Output (Existing Turbine)	1605	Boiler Heat Input	3656
Steam Coil Air Heater Output	0	PFWH Heat Input	90
Condenser Loss	2143	Condensate Pump	2
Total Energy Output	3748	**Total Energy Input**	3748
		In - Out	0

Turbine Heat Rate
8089
(Btu/kwhr)

Figure 3.3. 17: Modified Steam Cycle Diagram and Performance for Concept B

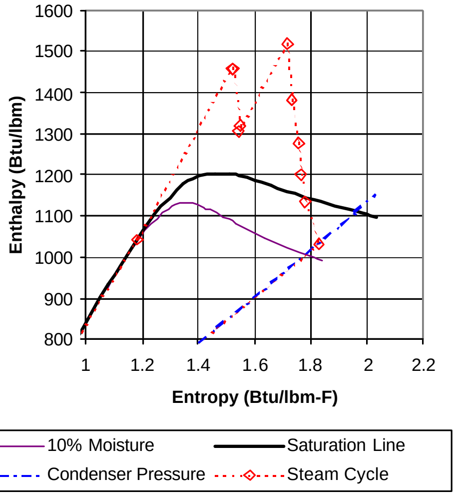


Figure 3.3. 18: Modified Steam Cycle Mollier Diagram for Concept B

3.3.6 Flue Gas Desulfurization and Electrostatic Precipitator Systems Performance Issues

Because the flue gas for Concept B was significantly different in composition from the Base Case due to combustion with oxygen as opposed to air, analysis of the Flue Gas Desulfurization System (FGD) system and the Electrostatic Precipitator (ESP) system performance with these new conditions was investigated. The most noticeable differences regarding the flue gas composition for this case is that it contains significantly more CO_2 and less N_2 than with air firing. The overall mass flow rates for both cases are nearly identical. The effects of operating the Flue Gas Desulfurization System and the ESP with the high CO_2 content flue gas of Concept B are discussed in this section.

3.3.6.1 Flue Gas Desulfurization System Performance Issues

Table 3.3.12 identifies the assumptions that were made in predicting the FGD system performance. Table 3.2.13 shows the gas constituents at the existing Absorber inlet and outlet locations. It should be noted that the CO_2/SO_2 mole ratio is 199, as opposed to the value of 63 encountered previously for the Base Case and Concept A (refer to Section 3.2.4). Hence, it was necessary to assess the impact of high CO_2 content in the flue gas on the performance of FGD System. It is important to note that SO_2 forms a much stronger acid than CO_2 (with a dissociation constant of 7.8E-3 vs. 5.2E-7). Additionally, SO_2 is about 35 times more soluble in water than CO_2. The absorbed SO_2 will drop the pH of the scrubbing solution to about 4.5. At this level, the CO_2 is basically insoluble in water. ALSTOM Power's Environmental Systems group predicted that, under these circumstances, the flue gas desulfurization systems SO_2 capture efficiency would decrease by approximately 2%.

Table 3.3. 12: FGD System Assumptions

Concept B (O_2 Fired)		
Quantity	Unit	Existing Absorber
Ca/S)	Mol Ratio	1.04
Solids	Wt.%	20
CaO	Wt.%	90
MgO	Wt.%	5
Inerts	Wt.%	5
By-pass Leakage	Wt.%	2.5
Liquid/Gas (L/G) Ratio	gpm/1000 acfm	55
SO_2 Removal Efficiency		
APC	%	94.9
Absorber	%	97.3

Table 3.3. 13: FGD System Performance

	Concept B (O_2 Fired)					
	Existing Absorber Inlet			Existing Absorber Outlet		
Species	Mol/hr	Vol.%	Unit	Mol/hr	Vol.%	Unit
O_2	3,200	3.09	Vol.%	3,190	2.87	Vol.%
N_2	5,467	5.27	Vol.%	5,467	4.92	Vol.%
H_2O	14,570	14.05	Vol.%	22,400	20.16	Vol.%
CO_2	80,031	77.20	Vol.%	80,021	72.03	Vol.%
SO_2	403	3,887	vppm	21	189	vppm
SO_2 Removal Efficiency, %					94.8	
CO_2/SO_2 Mole Ratio		199				

3.3.6.2 Electrostatic Precipitator (ESP) System Performance Issues

The effects of operating the Electrostatic Precipitator (ESP) in the high CO_2 content flue gas environment of Concept B are discussed in this section.

The electrostatic precipitation process can be characterized as four stages.

- Corona generation and flue gas ionization
- Particle charging resulting from ionization
- Migration of particles under the influence of the electric field
- Removal of particle cake from collecting electrodes

Background From White, J. “Industrial Electrostatic Precipitation” (1962):
Corona discharge is vital to the electrical separation of particles from the gas stream. The dominant ion production mechanism in the corona is ionization by electron impact, in which free electrons in the gas acquire energy from an applied electric field and collide violently with gas molecules, literally knocking electrons out of the molecules. The net result is free electrons and positively charged gas ions.

The unipolar corona, used in electrical precipitation, is a stable, self-maintaining gas discharge between an emitting electrode and a receiving electrode. The ionization processes are confined to or near the glow region in the strong electric field adjacent to the emitting electrode. Most of the ionization is produced by free electrons that are accelerated to fairly high energies in this region and ionize by collision with molecules.

With negative corona free electrons from the ionization zone, upon entering the low-field region of the corona, combine with molecules of the gas to form negative ions. The ability to form negative ions is a fundamental property of the molecular species. Some gases such as nitrogen, hydrogen, helium, neon and argon, if sufficiently pure, have no affinity for electrons and hence do not form negative ions. Negative corona does not occur in these gases. Instead, when the voltage is raised to a point that would correspond to the corona onset point, spark over occurs. On the other hand, oxygen, chlorine, sulfur dioxide, and many other gases do have strong electron affinities, and, as would be expected, they produce highly stable negative coronas. Negative corona is possible only in gases, or mixtures of gases, that exhibit appreciable electron attachment. Consequently negative corona characteristics are highly sensitive to gas composition and can range from no corona to highly stable corona.

Fortunately practically all industrial furnaces, boilers and process gases as well as air contain electro-negative gases such as oxygen, water vapor, carbon dioxide and frequently sulfur dioxide. Electron attachment is the process by which electrons combine with neutral molecules to form negative ions. The probability of attachment varies greatly with the gas composition. It is zero for gases such as nitrogen and hydrogen, very small for ammonia, and relatively high for chlorine.

Theory and experiment show that the molecular composition of the gas profoundly influences its negative-corona characteristics (Please refer to Figures 3.3.19, 3.3.20 and 3.3.21 below, extracted from White, 1962). The general conclusion is that gases such as nitrogen that in the pure state have zero electron affinity, are incapable of supporting negative corona, and gases of the CO_2 type that have moderate electron affinity do have considerable negative corona, and gases of the SO_2 type have a very wide range of negative corona.

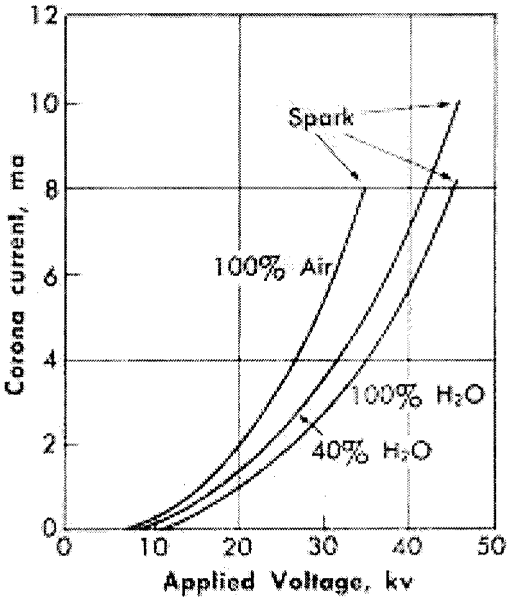


Figure 3.3. 19: Negative Corona Curves for Air-water Vapor Mixtures at 400F; Atmospheric Pressure; 3-in. Tube; 0.010-in. Wire

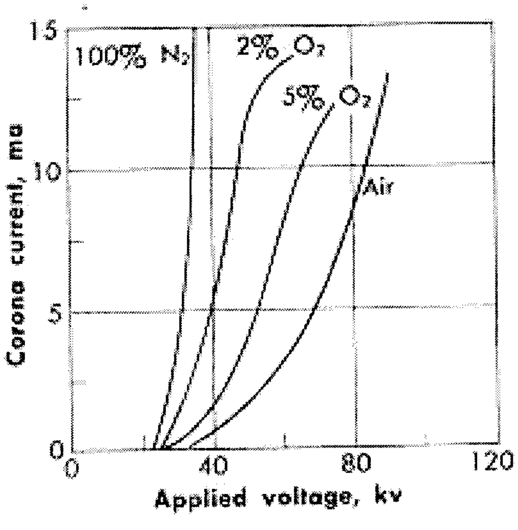


Figure 3.3. 20: Negative Corona Characteristics for Nitrogen-Oxygen Mixtures; 6-in Tube; 0.109-in Wire

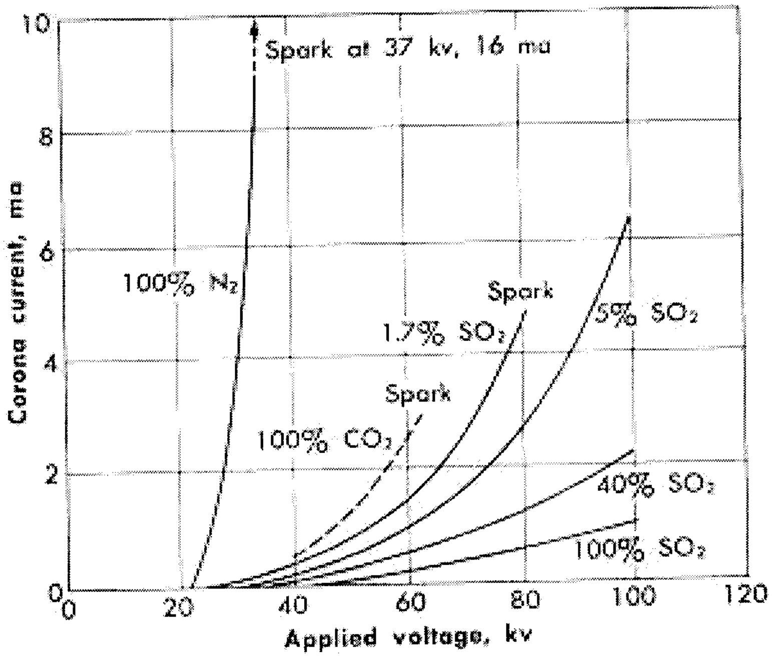


Figure 3.3. 21: Negative Corona Curves for Nitrogen-Sulfur Dioxide Mixtures; 6-in. Tube; 0.019-in. Wire

Conditions with recycled flue gas:
Table 3.3.14 gives a comparison of the flue gas analysis for air firing and O_2/Recycled flue gas firing. The high nitrogen content, 73.7%, found with the air firing is reduced to 5.3 % where as the CO_2 increases from 13.9% to 77.2%. There is also a small increase in the moisture content from 9% to 14.1% and SO_2 content from 2212 vppm to 3887 vppm.

Table 3.3. 14: Composition of Flue Gases Entering the Electrostatic Precipitator (ESP)

Species	Unit	Air Firing: (ESP Inlet Gas Composition)	O_2/Recycled Flue Gas Firing: (ESP Inlet Gas Composition)
O_2	Vol.%	3.1	3.1
N_2	Vol.%	73.7	5.27
H_2O	Vol.%	9.0	14.1
CO_2	Vol.%	13.9	77.2
SO_2	vppm	2212	3887

If the gas was 100% Nitrogen, the negative corona precipitation process could not occur. Nitrogen has no affinity for electrons and hence cannot form negative ions. When the voltage is raised to a point that would correspond to the corona onset point, spark over occurs. Figure 3.3.20 shows that the introduction of O_2 allows an increase in voltage to take place creating a stronger electrical field prior to spark over. Figure 3.3.21 shows again the 100% nitrogen gas and the effects of various levels of the electro-negative gas $S0_2$. Also a comparison curve is given for 100% CO_2. It can be seen that CO_2 has electron affinity resulting in a significant increase in voltage, beyond corona onset voltage, before spark break down occurs. The curve tends to the improved characteristic that is obtained with the presence of a small percentage of $S0_2$.

Hence with all other gas components remaining constant the replacement of nitrogen by CO_2 will result in a gas that is more beneficial for negative corona precipitation.

In Conclusion, for the gas analysis of Concept B shown above in Table 3.3.14, no ESP performance degradation is expected as a result of firing coal in O_2/Recycled flue gas environment.

3.3.7 Carbon Dioxide Separation and Compression System

This system processes the flue gas stream leaving the oxygen fired boiler system of Concept B to provide a liquid product CO_2 stream with suitable conditions for sequestration or usage.

3.3.7.1 Introduction:

Traditionally, amine scrubbing has been used on industrial scale to recover CO_2 from boiler flue gases. Monoethanolamine (MEA) is one of the most effective absorbents but reclamation of the absorbed CO_2 from MEA solution consumes large amounts of energy. Therefore, a search is on for more energy efficient methods for recovering CO_2 from flue gas streams. One proposed method is to replace combustion air of a power plant with nearly pure oxygen. Theoretically, the flue gas from such a plant would consist of only CO_2 and water vapor. Because water can be separated from this gas relatively easily (by condensation) this method seems to offer a promising way to recover CO_2 by simply compressing it to a high enough pressure so it can be disposed of or recovered for some useful purpose. Practically however, the stream leaving the Boiler Island contains many other components other than just CO_2 and water vapor, which must be fully considered in the system design for Concept B.

3.3.7.2 Process Description:

Figure 3.3.22 (Refer to Section 3.3.7.3) shows the Flue Gas Cooling process flow diagram and Figure 3.3.23 shows the Flue Gas Compression and Liquefaction process flow diagram.

The following describes a CO_2 recovery system that cools and then compresses a CO_2 rich flue gas stream from an oxygen-fired 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 an acceptable level. 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 on the process schematics and will not be repeated here except in selected instances.

Later in this report there is a section titled “Design Considerations” (Section 3.3.7.8). This section covers design issues that are not discussed in the Process Description below.

<u>Flue Gas Cooling:</u>

Please refer to Figure 3.3.22 (drawing D 09484-01005R-0).

The feed to the CO_2 Recovery System is the flue gas stream that leaves the FGD system of the Boiler Island. At this point, the flue gas is at the dew point of H_2O. All of the flue gas leaving the boiler is cooled to 100 °F in Gas Cooler DA-101 A/B/C which operates essentially under atmospheric pressure. A significant amount of water condenses out in this cooler. Approximately 2/3 of the gas is then recycled back to the boiler while 1/3 is fed forward to the CO_2 compression area. The recycle stream is required to maintain thermal balance in the existing Conesville #5 steam generator unit in order to avoid major pressure part modifications to the boiler. The Gas Cooler minimizes the volumetric flow rate to, and the resulting power consumption of, the Flue Gas Compression equipment located downstream. Excess condensate is blown down to the cooling water system. Three vessels have been provided for these coolers because a single vessel would be too large.

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 A/B/C/D 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-101 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 flue gas stream enters a series of booster blowers that are located adjacent to the Gas Cooler. 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. This arrangement also minimizes energy consumption, as it does not needlessly over boost the pressure of the stream to be recycled back to the boiler. It is only necessary to boost the pressure of the fraction of the flue gas flow that proceeds to compression and liquefaction to overcome the pressure drop of the duct which is about 1200 feet long.

<u>Three-Stage Gas Compression System:</u>

Please refer to Figure 3.3.23 (drawing D 09484-01004R-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 come from the new cooling tower. At this pressure and temperature, about 80 mole % of the stream can be condensed. The flash vapors contain approximately 80 weight % of the inlet oxygen and nitrogen, but also 12 weight % 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 %). Then the pressure of the liquid is boosted to 2000 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 80,000 ACFM to each of the two trains and only a single frame is required in each train. 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%.

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-103) cools the gas down 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 wet, some water condenses out in the three aftercoolers. The sour condensate is separated in knockout drums (FA-101/2/3) 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.

Flue gas leaving the 3rd stage discharge knockout drum (FA-103) is fed to Flue Gas Drier FF-101 A/G where additional moisture is removed.

<u>Gas Drying:</u>

Please refer to Figure 3.3.23 (drawing D 09484-01004R-0).

It is necessary to dry the CO_2 stream to meet the product specification. 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 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% 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 six vessels; FF-101 A/G. One vessel is on line while the others are being regenerated. Flow direction is down during operation and up during regeneration.

The drier is regenerated with the non-condensable 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 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.

<u>CO_2 Condensation and Stripping:</u>

Please refer to Figure 3.3.23 (drawing D 09484-01004R-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. From EA-104 the partially condensed flue gas stream continues on to CO_2 Rectifier DA-102.

At this pressure and temperature 80 mole % of the stream can be condensed. The flash vapors contain approximately 80 weight % of the inlet oxygen and nitrogen, but also 12 weight % of the CO_2. Therefore, as mentioned, a rectifier tower has been provided to reduced the loss of CO_2 to an acceptable level. The pressure of the liquid is boosted to 2000 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% 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.

<u>CO_2 Pumping and CO_2 Pipeline:</u>

Please refer to Figure 3.3.23 (drawing D 09484-01004R-0).

The CO_2 product must be increased in pressure to 2000 psig. A multistage heavy-duty pump (GA-103A/B) 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 3.3.23 (drawing D 09484-01004R-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.

3.3.7.3 Process Flow Diagrams

Two process flow diagrams are shown below for these systems:

- Figure 3.3.22 (drawing D 09484-01005R-0) Flue Gas Cooling PFD
- Figure 3.3.23 (drawing D 09484-01004R-0) CO_2 Compression and Liquefaction PFD

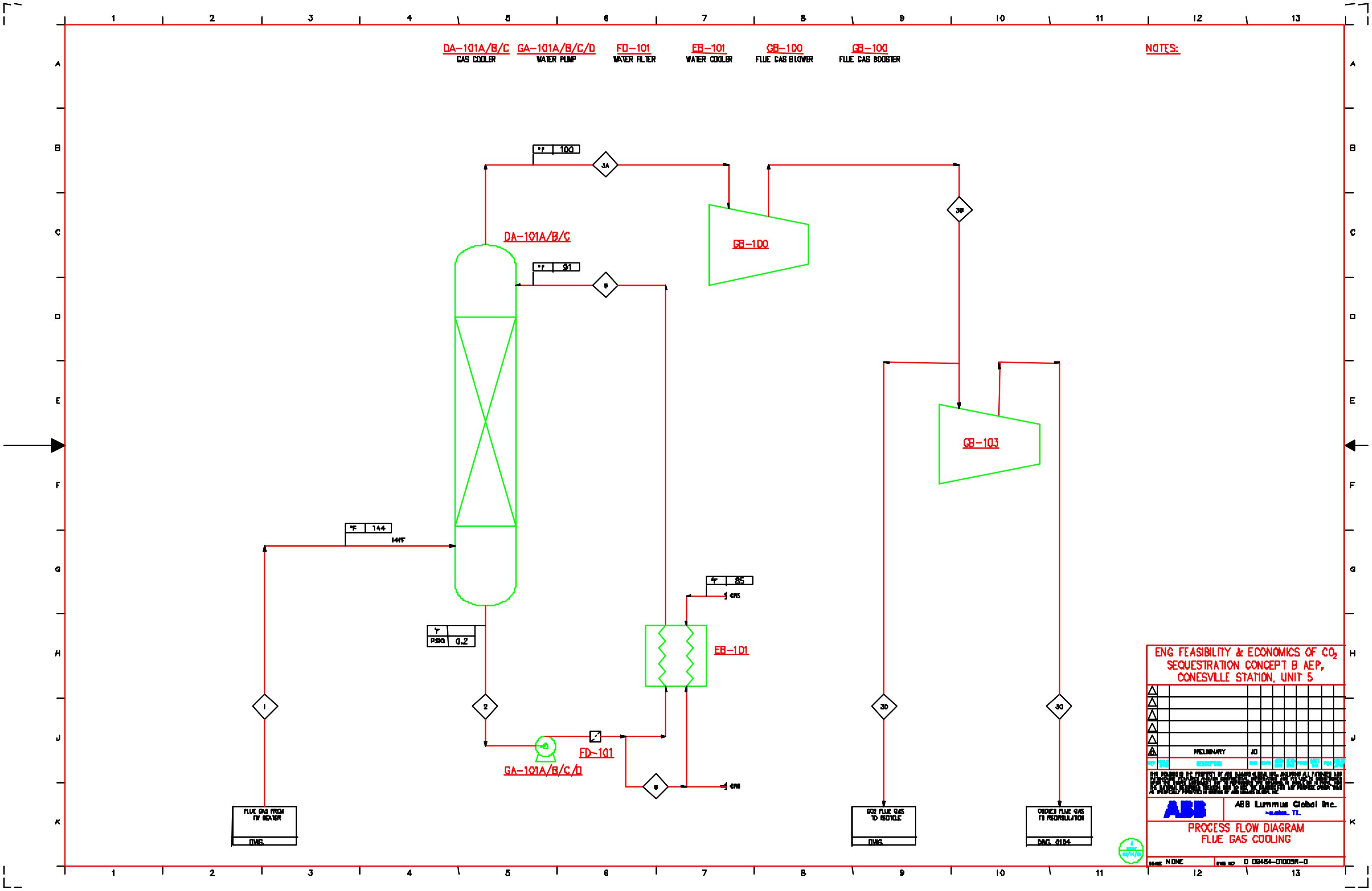


Figure 3.3. 22: Process Flow Diagram for Concept B: Flue Gas Cooling

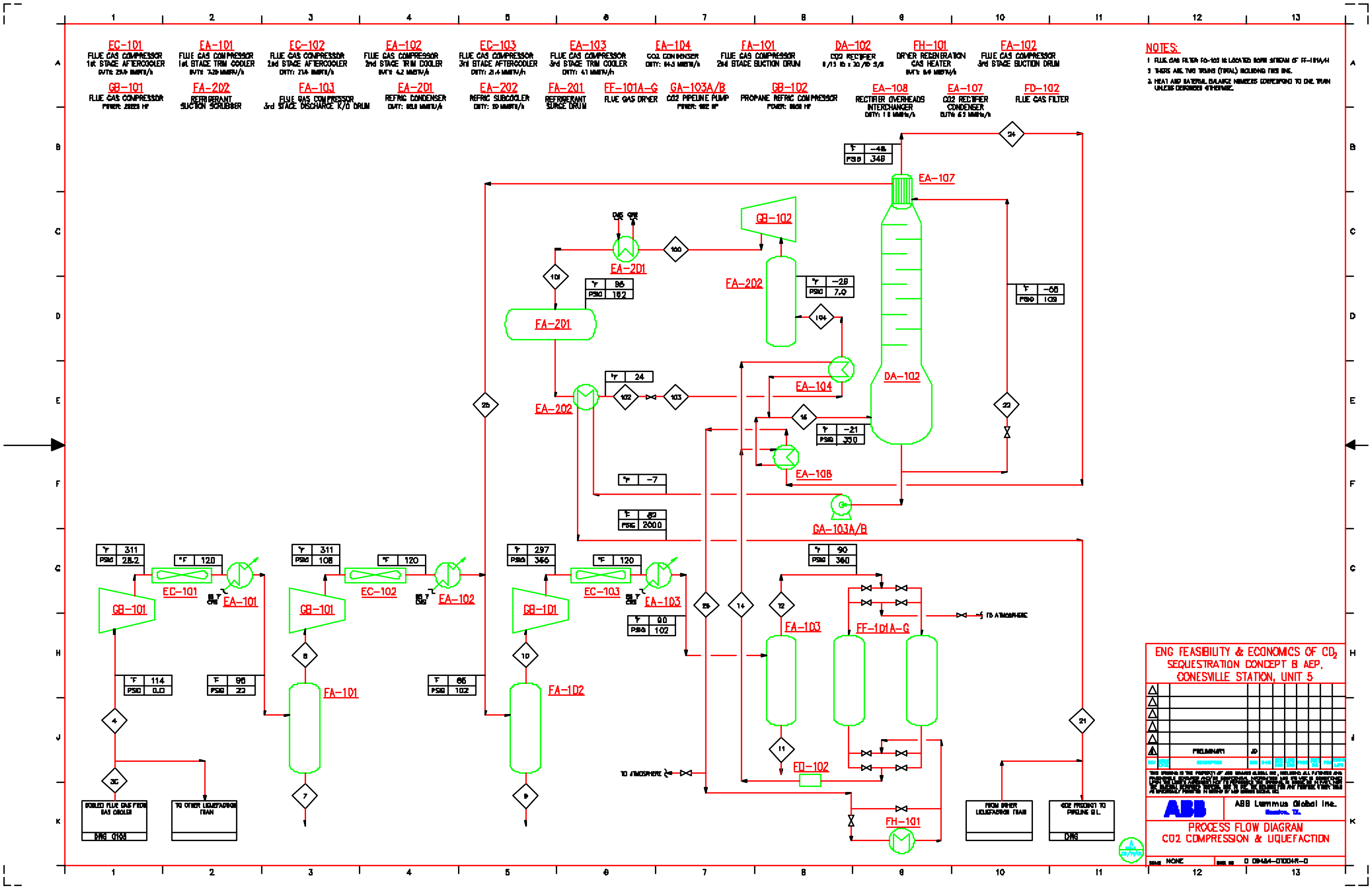


Figure 3.3. 23: Process Flow Diagram for Concept B: CO_2 Compression and Liquefaction

3.3.7.4 Overall Material and Energy Balance

Table 3.3.15 contains the overall material balance for the Flue Gas Cooling System and the CO_2 Compression and Liquefaction System. It is based on 94% recovery of CO_2.

Table 3.3. 15: Material and Energy Balance for Flue Gas Cooling, CO_2 Compression and Liquefaction

STREAM NAME		To quench columns	From Quench columns	Excess water	From Large blowers	Quench water out	Quench water in	To liquefaction trains	To boiler	To Train A liquefaction	First water KO	To 2nd stage	2nd water KO	To 3rd stage	Recycle from condenser
STREAM NO.		**1**	**3a**	**6**	**3b**	**2**	**5**	**3c**	**3d**	**4**	**7**	**8**	**9**	**10**	**25**
VAPOR FRACTION	Molar	1.000	1.000	0.000	1.000	1.000	1.000	1.000	1.000	1.000	0.000	1.000	0.000	1.000	1.000
TEMPERATURE	F	144	100	100	108	#N/A	91	114	108	114	95	95	86	86	-50
PRESSURE	PSIG	0	0	0	1	1	1	0	1	0	22	22	102	102	102
MOLAR FLOW RATE	LbMol/Hr	109,760.00	95,103.76	14,659.53	95,103.76	#N/A		22,857.50	72,246.26	11,428.75	499.76	10,928.99	15.06	11,675.19	925.00
MASS FLOW RATE	Lb/Hr	4,162,006	3,897,793	264,213	3,897,793	#N/A		936,806	2,960,987	468,403	9,015	459,389	273	496,582	40,427
ENERGY	Btu/Hr	5.24E+08	4.16E+08	-2.11E+08	4.23E+08	-		1.03E+08	3.21E+08	5.15E+07	-7.24E+06	4.70E+07	-2.20E+05	4.83E+07	2.66E+06
COMPOSITON	Mol %														
CO2		73.04%	84.29%	0.03%	84.29%	0.03%	0.03%	84.29%	84.29%	84.29%	0.09%	88.14%	0.30%	90.26%	97.82%
H2O		19.01%	6.53%	99.97%	6.53%	99.97%	99.97%	6.53%	6.53%	6.53%	99.91%	2.26%	99.69%	0.59%	0.00%
Nitrogen		5.03%	5.80%	0.00%	5.80%	0.00%	0.00%	5.80%	5.80%	5.80%	0.00%	6.07%	0.00%	5.78%	1.24%
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%	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%	0.00%
Oxygen		2.91%	3.36%	0.00%	3.36%	0.00%	0.00%	3.36%	3.36%	3.36%	0.00%	3.52%	0.00%	3.37%	0.92%
SO2		0.01%	0.02%	0.00%	0.02%	0.00%	0.00%	0.02%	0.02%	0.02%	0.00%	0.02%	0.00%	0.02%	0.02%
VAPOR															
MOLAR FLOW RATE	LbMol/Hr	109,760.0	95,103.8	-	95,103.8	-	-	22,857.5	72,246.3	11,428.8	-	10,929.0	-	11,675.2	925.0
MASS FLOW RATE	Lb/Hr	4,162,006	3,897,793	-	3,897,793	-	-	936,806	2,960,987	468,403	-	459,389	-	496,582	40,427
STD VOL. FLOW	MMSCFD	999.67	866.16	-	866.16	-	-	208.18	657.99	104.09	-	99.54	-	106.33	8.42
ACTUAL VOL. FLOW	ACFM	806,220.00	650,850.00	-	632,520.00	-	-	158,940.00	480,500.00	79,470.00	-	29,015.79	-	9,406.50	521.00
MOLECULAR WEIGHT	MW	37.92	40.98	-	40.98	-	-	40.98	40.98	40.98	-	42.03	-	42.53	43.71
DENSITY	Lb/Ft^3	0.09	0.10	0.10	0.10	0.1	0.1	0.10	0.10	0.10	-	0.26	-	0.88	1.29
VISCOSITY	cP	0.0147	0.0150	0.0150	0.0152	0.0	0.0	0.0154	0.0152	0.0154	-	0.0154	-	0.0155	0.0113
HYDROCARBON LIQUID															
MOLAR FLOW RATE	LbMol/Hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	Lb/Hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DENSITY	Lb/Ft^3	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-	-	-	-

STREAM NAME		To drier	3rd water ko	From drier / Condenser inlet	Condenser outlet	Non-condensable vent	Rectifier bottoms to condenser	Train A CO2 to pipeline	Refrig compressor discharge	Refrig condenser out	Refrig subcooler out	Refrig to CO2 condenser	Refrig from CO2 condenser	Warm non condensable
STREAM NO.		**12**	**11**	**14**	**15**	**24**	**22**	**21**	**100**	**101**	**102**	**103**	**104**	**26**
VAPOR FRACTION	Molar	1.000	0.000	1.000	0.200	1.000	0.134	0.000	1.000	0.000	0.000	0.158	0.993	1.000
TEMPERATURE	F	90	90	90	-21	-46	-56	82	144	95	24	-26	-26	81
PRESSURE	PSIG	359	359	354	349	346	105	2,000	169	162	159	8	8	341
MOLAR FLOW RATE	LbMol/Hr	11,638.57	36.62	11,606.46	11,318.96	1,421.61	925.00	9,247.50	9,750.00	9,750.00	9,750.00	9,750.00	9,750.00	1,421.61
MASS FLOW RATE	Lb/Hr	495,913	668	495,335	483,065	50,218	40,427	404,163	429,946	429,946	429,946	429,946	429,946	50,218
ENERGY	Btu/Hr	4.50E+07	-5.28E+05	4.50E+07	-1.86E+07	3.76E+06	-2.55E+06	-3.13E+06	7.13E+07	2.54E+06	-1.73E+07	-1.73E+07	4.67E+07	5.35E+06
COMPOSITON	Mol %													
CO2		90.54%	0.88%	90.79%	90.79%	40.47%	97.82%	97.82%	0.00%	0.00%	0.00%	0.00%	0.00%	40.47%
H2O		0.28%	99.11%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		5.79%	0.00%	5.81%	5.81%	38.53%	1.24%	1.24%	0.00%	0.00%	0.00%	0.00%	0.00%	38.53%
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		3.38%	0.00%	3.39%	3.39%	20.99%	0.92%	0.92%	0.00%	0.00%	0.00%	0.00%	0.00%	20.99%
SO2		0.02%	0.01%	0.02%	0.02%	0.00%	0.02%	0.02%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR														
MOLAR FLOW RATE	LbMol/Hr	11,638.6	-	11,606.5	2,263.8	1,421.6	124.2	-	9,750.0	-	-	1,539.9	9,681.4	1,421.6
MASS FLOW RATE	Lb/Hr	495,913	-	495,335	87,306	50,218	5,207	-	429,946	-	-	67,905	426,923	50,218
STD VOL. FLOW	MMSCFD	106.00	-	105.71	20.62	12.95	1.13	-	88.80	-	-	14.02	88.17	12.95
ACTUAL VOL. FLOW	ACFM	2,686.04	-	2,718.77	407.73	253.59	67.90	-	4,762.22	-	-	5,111.32	32,135.98	366.87
MOLECULAR WEIGHT	MW	42.61	-	42.68	38.57	35.32	41.94	-	44.10	-	-	44.10	44.10	35.32
DENSITY	Lb/Ft^3	3.08	-	3.04	3.57	3.30	1.28	-	1.50	-	-	0.22	0.22	2.28
VISCOSITY	cP	0.0164	-	0.0164	0.0145	0.0146	0.0116	-	0.0098	-	-	0.0066	0.0066	0.0185
HYDROCARBON LIQUID														
MOLAR FLOW RATE	LbMol/Hr	-	-	-	9,055.12	-	800.84	9,247.50	-	9,750.00	9,750.00	8,210.14	68.56	-
MASS FLOW RATE	Lb/Hr	-	-	-	395,757.44	-	35,219.85	404,163.41	-	429,945.75	429946	362042	3,023.15	-
STD VOL. FLOW	BPD	-	-	-	32,774	-	2,921	33,471	-	58,100	58,100	48,927	409	-
ACTUAL VOL. FLOW	GPM	-	-	-	748.51	-	61.47	1,008.54	-	1,802.87	1,598.64	1,268.98	10.60	-
DENSITY	Lb/Ft^3	-	-	-	65.92	-	71.43	49.96	-	29.73	33.53	35.57	35.57	-
MOLECULAR WEIGHT	MW	-	-	-	43.71	-	43.98	43.71	-	44.10	44.10	44.10	44.10	-
VISCOSITY	cP	-	-	-	0.1610	-	0.2221	0.0558	-	0.0906	0.1336	0.1771	0.1771	-
SURFACE TENSION	Dyne/Cm	-	-	-	15.05	-	20.06	0.85	-	5.74	10.54	14.13	14.13	-

3.3.7.5 Equipment List with Data

Equipment data for Concept B, summarized in Appendix II, has been tabulated in the so-called "short spec" format which provides adequate data for a factored cost estimate.

3.3.7.6 Consumption of Utilities

In addition to the primary utilities listed below (Table 3.3.16), the plant will also need a supply of other utilities such as instrument air, plant air, process water, nitrogen etc. However, these needs are minor and in many cases are for maintenance purposes only. Consumption of these utilities has not been estimated.

Table 3.3. 16: Concept B Utility Consumption

Utility	Amount Consumed	Units
Natural Gas	0.26	MMSCFD
Steam	0	Lb/hr
Cooling water	93,200	Gpm

Number of Trains	Item Number	Service	Number Operating per train	Power (ea) including 0.95 motor eff (kW)	Total all trains (kW)
2	EC-101	Flue Gas Compressor 1st Stage Aftercooler	1	81	163
2	EC-102	Flue Gas Compressor 2nd Stage Aftercooler	1	69	138
2	EC-103	Flue Gas Compressor 3rd Stage Aftercooler	1	68	137
1	PA-101A/B	Large Air Separation Unit	2	40255	80511
1	PA-102	Small Air Separation Unit	1	15311	15311
1	PA-103	Cooling Tower	1	4074	4074
2	GB-101	1 Stage	1	6416	12831
2		2 Stage	1	6675	13349
2		3 Stage	1	6718	13436
2	GB-102	1 stage	1	2362	4724
2		2 stage	1	5204	10408
1	GB-100	Flue Gas Blower	1	2173	2173
1	GB-103	Flue Gas Booster	1	396	396
1	GA-101 A/B/C/D	Water pump	3	104	311
2	GA-103A/B	CO2 Pipeline pump	1	1045	2089
		Total			160051

3.3.7.7 Consumption of Chemical and Desiccants

The CO_2 compression system does not need any chemicals for its operation. Naturally, there will be a minor demand of lube oil and similar supplies. These have not been estimated.

It has been assumed that the mole sieve desiccant in the CO_2 Drier must be replaced once every 3 years. The estimated consumption of mole sieve and other chemicals can be found Table 3.3.17.

Table 3.3. 17: Concept B Chemical and Desiccant Consumption

Chemical	Consumption per day (lbs.)
Sodium Hypochlorite	15300
Sodium Bisulfite	86
Mole Sieves	136

The totals shown in Table 3.3.17 do not include chemicals provided by the cooling tower service personnel. This is handled as a component of operating costs referred to as contracted services. It also does not include air separation plant chemicals and lubricants that were presented as monetary value only. They will appear in the operating expense break down.

3.3.7.8 Design Considerations

The process and equipment specifications presented for Concept B were the result of a number of optimizations. The areas of optimizations in the CO_2 Compression and Liquifaction System were:

- O_2 purity
- CO_2 purity
- CO_2 condensation pressure and temperature

O_2 Purity:
Initial analysis included an Air Separation Unit (ASU) producing oxygen of 95% purity. Additionally, 5% of the total oxygen demand was assumed to be from infiltration of ambient air into the boiler which is typical for a boiler of this type and age. These operating conditions combined with the need for nominally 15% excess oxygen for combustion, resulted in almost 15-wt% of nitrogen and oxygen in the flue gas. This concentration is much higher than typically accepted for EOR applications. The conclusion then was that the bulk of the inerts must be separated from CO_2 before it can be used for EOR. Thus, direct compression to the pipeline is not an option. Instead, CO_2 would have to be condensed so the inerts could be separated.

It soon became obvious that it is impossible to condense CO_2 from such a mixture at pressures below the critical pressure of CO_2 and temperatures achievable with cooling water. This is illustrated by Figure 3.3.24 which shows the achievable CO_2 recovery (the ratio of pure CO_2 product recovered to the CO_2 in the flue gas entering the system) as a function of pressure and temperature. Pressure was limited to 1000 psig, which is just slightly below the critical pressure of CO_2 (1070 psia).

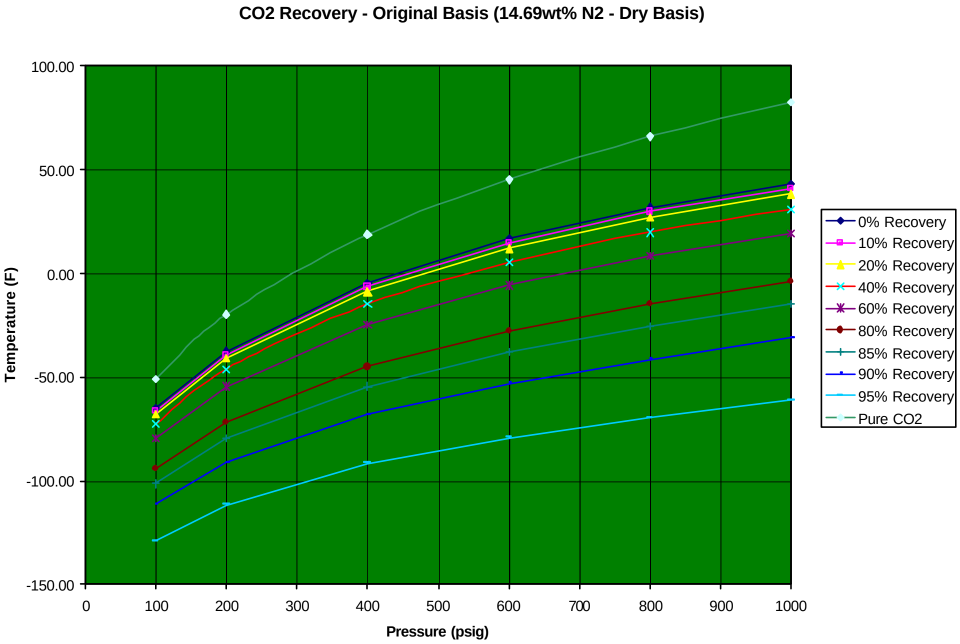


Figure 3.3. 24: CO_2 Recovery from Boiler Flue Gas (Combustion with 95% Pure O_2 and 5% Air Infiltration)

Because the combustion process requires excess oxygen (typically 15-20% with coal firing), the flue gas will always contain at least some oxygen. Furthermore, it is not easy to completely eliminate nitrogen. Some infiltration of air will always occur as long as the combustion chamber of the boiler operates under slight vacuum, as is the current practice with coal firing, thus some N_2 will also be present. Additionally, increasing oxygen purity from the ASU beyond 99% is expensive because the remaining 1% is mainly argon and the separation becomes more difficult.

Because the inert gases have a strong effect on the dew point of the flue gas, the only way to condense a substantial fraction of the CO_2 is by refrigeration. One should keep in mind that the critical temperature of CO_2 is approximately 88 F and condensation of CO_2 without some form of refrigeration may not be feasible in the warmer regions of the world even when inerts are not present in the CO_2 stream.

Figure 3.3.25 shows the achievable CO_2 recovery as a function of pressure and temperature using 99% pure O_2 and 1% air infiltration as combustion process assumptions. One comparison which emphasizes the impact of flue gas purity is that with 95% O_2 purity, the temperature for 95% recovery at 1000 psig is -60°F (Figure 3.3.24) whereas it is 30 °F (Figure 3.3.25) with 99% pure oxygen.

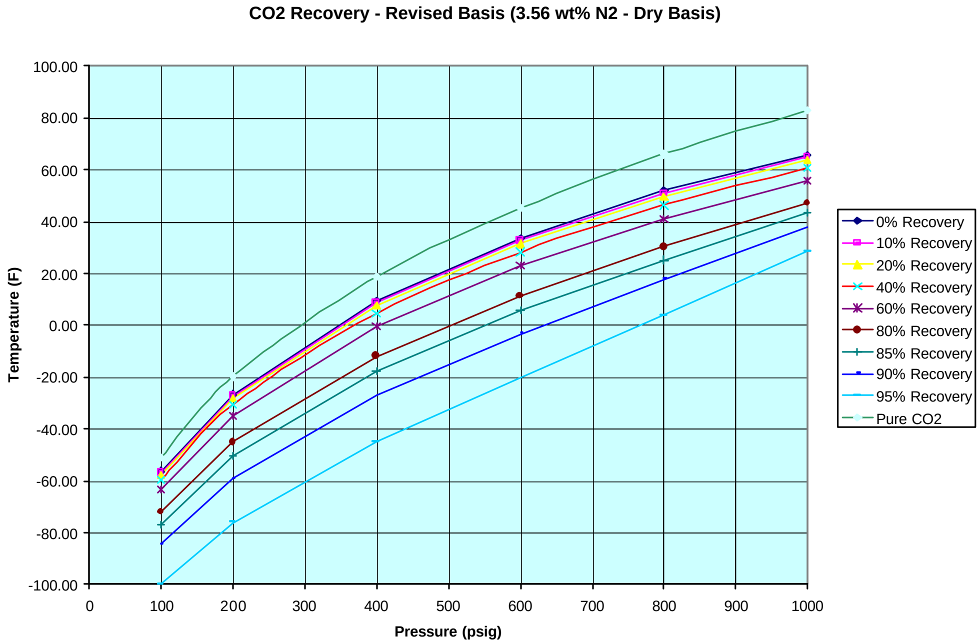


Figure 3.3. 25: CO_2 Recovery Boiler Flue Gas
(Combustion with 99% Pure O_2 and 1% Air Infiltration)

CO_2 purity:

Simulations were made to evaluate modifications to the liquefaction flow scheme presented in Sections 3.3.7.2 & 3.3.7.3 for this concept. The alternate schemes evaluated included total product condensation and having a reboiler on the rectifier (turning it into a complete distillation column). The result was that total condensation, even with the use of 99% pure oxygen, left the product with approximately 11% inerts, which cast doubt according to experts that it could be sequestered in a body of water. The result of having a reboiler on the bottom of the rectifier showed that it improved the CO_2 product purity, but could never be feasibly made to produce a product which contained less than 10 ppm O_2 which would be required for EOR applications. The reason that a reboiler was not presented in the final design was that all of the heat of reboiling was added to the condenser, which required refrigeration at a much lower temperature than the refrigeration which was recovered in the reboiler. The result was a significant work of liquefaction increase.

CO_2 condensation pressure and temperature:

The range of condensation pressures evaluated ranged from 995 psia to 70 psia. The range of condensation temperatures evaluated ranged from 41 °F to –69 °F. The optimum was found to be 350 psig and –22 °F.

3.3.7.9 OSBL Systems

Concept B has equipment associated with the liquefaction of CO_2 in four different areas. The areas include the cooling tower, flue gas cooling section, air separation plant, and the compression and liquefaction equipment.

For this Concept, all of the cooling water must come from a new cooling tower since there is no diversion of steam to supply process heat. The clarifier producing make-up water for the cooling tower produces a blowdown sludge that must be sent to the existing clarifier blowdown handling system. The blowdown from the cooling tower itself can be discharged to the river after the free chlorine is removed by injection of sodium bisulfite.

The air separation plant utilizes the bulk of the cooling water. It also consumes electric power. The original design of the air separation plant involved the consumption of steam for molecular sieve drier regeneration. This duty was converted to an electric load to get the CO_2 recovery process isolated from swings in steam pressures that can occur with changes in throughput. The only other utility it takes in is instrument air.

The CO_2 compression and liquefaction section requires less supplemental utilities and chemicals than its amine counterparts. Only cooling water and electric power are required. Once every three years the molecular sieve in the dryers may need to be replaced.

3.3.7.10 Plant Layout

The gas from the flue gas desulfurizer arrives at the nozzles of the flue gas coolers at atmospheric pressure. In order to avoid any additional inflow of oxygen, the flue gas coolers must be placed as close to the power boilers as possible. The coolers are located just west of the existing FGD system.

Initially there were concerns about the piping between the air separation plant and the power boiler. The oxygen is produced at the air separation plant at 5 psig. Therefore it should not be subject to more than 3 psi of frictional pressure loss in the duct connecting it to the power boiler. Calculations reveal that two 36-inch supply lines can carry the oxygen from the ASU to the Boiler, a distance of about 1500 feet. The wall thickness of standard pipe should provide more than enough protection from most incidents that could be envisioned for a pipe spanning this distance. There may be even scope to reduce the wall thickness to less than standard wall if calculations permit. Air Products also had concerns about putting the Air Separation plant too close to the power plant where the inlet air may be high in CO_2. This CO_2 could affect the drier operation. Due to the fact that this CO_2 liquefaction unit reduces gases going up the stack, there should be no problem.

The layout of the compression and liquefaction unit offers few areas to comment on. At the low-pressure section of the plant, elbows are quite large making spacing allowances on the plot plan difficult without a rigorous design being made.

Plant layout drawings for Concept B (listed below) are included in Appendix I. This new equipment requires about 5.5 acres of plot area.

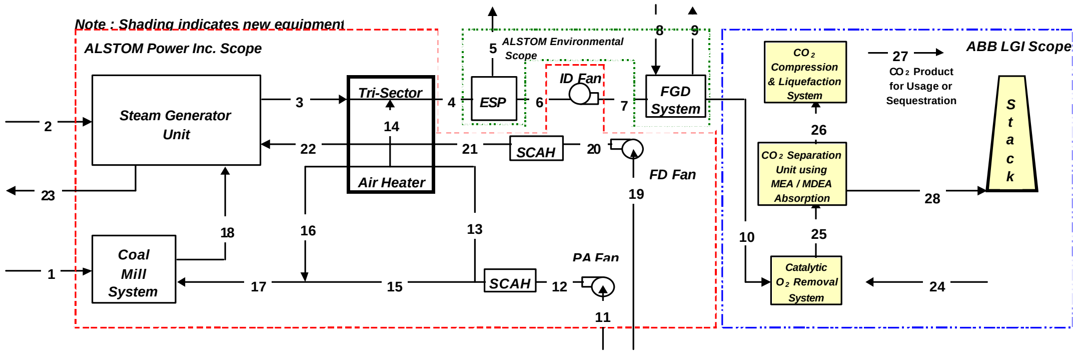
U01-D-0203 Plot Plan – Concept B: Air Separation Plants
U01-D-0209 Plot Plan – Concept B: Flue Gas Cooling
U01-D-0205 Plot Plan – Concept B: CO_2 Compression & Liquefaction
U01-D-0212 Plot Plan – Concept B: Overall Layout Conceptual Plan
U01-D-0201R Plot Plan – Concept B: Modified Overall Site Plan

3.4 Concept C: CO_2 Separation by MEA/MDEA Absorption and De-Oxy Catalyst

Concept C is defined as a system designed for the removal and recovery of CO_2 from the boiler flue gas using MEA/MDEA blended amine technology. The difference between the amines used in this case and in Case A is that the mixture of amines are not oxygen resistant. Therefore, the oxygen in the flue gas is converted to CO_2 by combustion with methane over a de-oxy catalyst upstream of the amine contactor.

3.4.1 Overall System Description and Material and Energy Balance

A simplified process flow diagram for the modified unit is shown in Figure 3.4.1. The operation and performance of the existing Boiler, ESP, and FGD systems are identical to the Base Case and are not affected by the addition of the MEA/MDEA based CO_2 removal system. The gas side material and energy balance for the Boiler Island system is shown in Table 3.4.1 below. The flue gases leaving the existing FGD system are ducted to the new de-oxy system. The de-oxy system converts the remaining oxygen contained in the flue gas to CO_2 using two stages of catalytic combustion with natural gas as the fuel source. High temperature heat recovery is provided in the de-oxy system by generation of high pressure superheated steam, which is expanded through a new steam turbine for additional power generation. The exhaust from this turbine provides part of the feed for solvent regeneration in the re-boilers of the MEA/MDEA system. The de-oxygenated flue gas leaving the de-oxy system is supplied to the MEA/MDEA system where about 91 percent of the CO_2 is removed, compressed, and is available for usage or sequestration. The remaining flue gases leaving the new MEA/MDEA system absorber, consisting of primarily, nitrogen, water vapor, carbon dioxide, and relatively small amounts of sulfur dioxide and methane, is discharged to the atmosphere through stacks above the absorbers.



Material Flow Stream Identification

1	Raw Coal to Pulverizers
2	Air Infiltration Stream
3	Fluegas from Economizer to Air Heater
4	Fluegas Leaving Air Heater to ESP
5	Flyash Leaving ESP
6	Fluegas Leaving ESP to Induced Draft Fan
7	Fluegas to Fluegas De-Sulfurization System
8	Lime feed to FGD System
9	FGD System Solids to Disposal
10	Fluegas to O_2 Removal System
11	Air to Primary Air Fan
12	Primary Air to Steam Coil Air Heater
13	Primary Air to Air Heater
14	Air Heater Leakage Air Stream
15	Tempering Air to Pulverizers
16	Hot Primary Air to Pulverizers
17	Mixed Primary Air to Pulverizers
18	Pulverized Coal and Air to Furnace
19	Air to Forced Draft Fan
20	Secondary Air to Steam Coil Air Heater
21	Secondary Air to Air Heater
22	Heated Secondary Air to Furnace
23	Bottom Ash from Furnace
24	Methane for Oxygen Consumption
25	Oxygen Free Gas to CO_2 Removal
26	Separated CO_2 to Compressor
27	Compressed CO_2 Product
28	CO_2 Depleted Fluegas to Stack

Figure 3.4. 1: Simplified Gas Side Process Flow Diagram for CO_2 Separation by MEA/MDEA Absorption for Concept C

Table 3.4. 1: Gas Side Material and Energy Balance for Concept C

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11	12	13
O2	(Lbm/hr)	26586	42147	101097	144817		144817	144817	5355		144578	203237	203237	112918
N2	"	4868	139626	2797385	2942220		2942220	2942220			2942220	673283	673283	374075
H2O	"	37820	2357	228849	231294		231294	231294	250709	45979	436024	11365	11365	6314
CO2	"			867210	867210		867210	867210			866156			
SO2	"			20202	20202		20202	20202			1063			
H2	"	16102												
Carbon	"	236655												
Sulfur	"	10110												
Ca	"								12452					
Mg	"								584					
MgO	"									484				
MgSO3	"									1293				
MgSO4	"									94				
CaSO3	"									35179				
CaSO4	"									2468				
CaCO3	"									2398				
Ash / Inerts	"	42313		33851	33851	33851			968	968				
		Raw Coal	Leakage Air	Fluegas to AH	Fluegas to ESP	Flyash	Fluegas to ID Fan	Fluegas to FGD	Lime Slurry	FGD Disposal	Fgas to CO2 Sep	Pri Air to PA Fan	PA from PA Fan	Pri Air to AH
Total Gas	(Lbm/hr)		184130	4014743	4205743		4205743	4205743			4390042	887885	887885	493308
Total Solids	"	**374455**		33851	33851	33851			14003	42884				
Total Flow	"	374455	184130	4048594	4239594	33851	4205743	4205743	270067	88863	4390042	887885	887885	493308
Temperature	(Deg F)	**80**	80	**706**	**311**	311	311	325	80	136	136	**80**	92	92
Pressure	(Psia)	14.7	14.7	14.6	14.3	14.7	14.2	15.0	14.7	14.7	14.7	14.7	15.6	15.6
hsensible	(Btu/lbm)	0.000	0.000	161.831	57.924	57.750	57.924	61.384	0.000	14.116	14.543	0.000	2.899	2.899
Chemical	106 Btu/h	4228.715												
Sensible	106 Btu/h	0.000	0.000	655.007	245.567	1.955	243.612	258.166	0.000	3.314	63.916	0.000	2.574	1.430
Latent	106 Btu/h	**0.000**	2.475	240.291	242.858	0.000	242.858	242.858	**0.000**	0.000	464.020	11.933	11.933	6.630
Total Energy(1)	106 Btu/h	4228.715	2.475	895.298	488.425	1.955	486.470	501.024	0.000	3.314	527.936	11.933	14.507	8.060

Constituent	(Units)	14	15	16	17	18	19	20	21	22	23	24	25	26
O2	(Lbm/hr)	43720	90319	66680	156999	183585	641283	641283	641283	643801				
N2	"	144835	299208	220899	520107	524975	2124443	2124443	2124443	2132785				
H2O	"	2445	5051	3729	8779	46599	35860	35860	35860	36001				
CO2	"													
SO2	"													
H2	"					16102								
Carbon	"					236655								
Sulfur	"					10110								
Ca	"													
Mg	"													
MgO	"													
MgSO3	"													
MgSO4	"													
CaSO3	"													
CaSO4	"													
CaCO3	"													
Ash / Inerts	"					42313					8463			
		Air Htr Lkg Air	Tempering Air	Hot Pri Air	Mixed Pri Air	Coal-Pri Air Mix	Sec Air to FD	Sec Air to SCAH	Sec Air to AH	Hot Sec Air	Bottom Ash	CO2 to Comp	CO2 Product	Vent Stream
Total Gas	(Lbm/hr)	191000	**394577**	291308	685885		2801587	2801587	2801587	2812587				
Total Solids	"										8463			
Total Flow	"	191000	394577	291308	685885	1060340	2801587	2801587	2801587	2812587	8463			
Temperature	(Deg F)	92	92	**666**	**339**		**80**	86.4	**86.4**	**616**	**2000**			
Pressure	(Psia)	15.6	15.6	15.6	15.6	15.0	14.7	15.2	15.1	14.9	14.7			
hsensible	(Btu/lbm)	2.899	2.899	145.249	63.358		0.000	1.549	1.549	132.582	480.000			
Chemical	106 Btu/hr)					4228.715								
Sensible	106 Btu/h	0.554	1.144	42.312	43.456		0.000	4.341	4.341	372.898	4.062			
Latent	106 Btu/h	2.567	5.303	3.915	9.218		37.653	37.653	37.653	37.801	0.000			
Total Energy(1)	106 Btu/h	3.121	6.447	46.227	52.674	4281.389	37.653	41.994	41.994	410.699	4.062			

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

Boiler efficiency is calculated to be 88.13 percent, the same as for the Base Case and Case A. The net plant heat rate is increased significantly to 15,223 Btu/kwhr for this case as shown in Table 3.4.2 which also includes the Base Case for comparison. The plant thermal efficiency for Case C (22.42%) is about 64 percent of the Base Case value (35.01%). Auxiliary power is increased to 95,317 kW and the net plant output is reduced to 335,973 kw. Fuel heat input to the overall system is increased by about 21 percent as compared to the Base Case due to the natural gas consumption of the de-oxy system. The fuel heat input to the boiler is the same as in the Base Case and Concept A. Carbon dioxide emissions are 92,153 lbm/hr or about 0.274 lbm/kWh which is about 13.7% of the Base Case value.

Table 3.4. 2: Overall Plant Performance Summary for Concept C

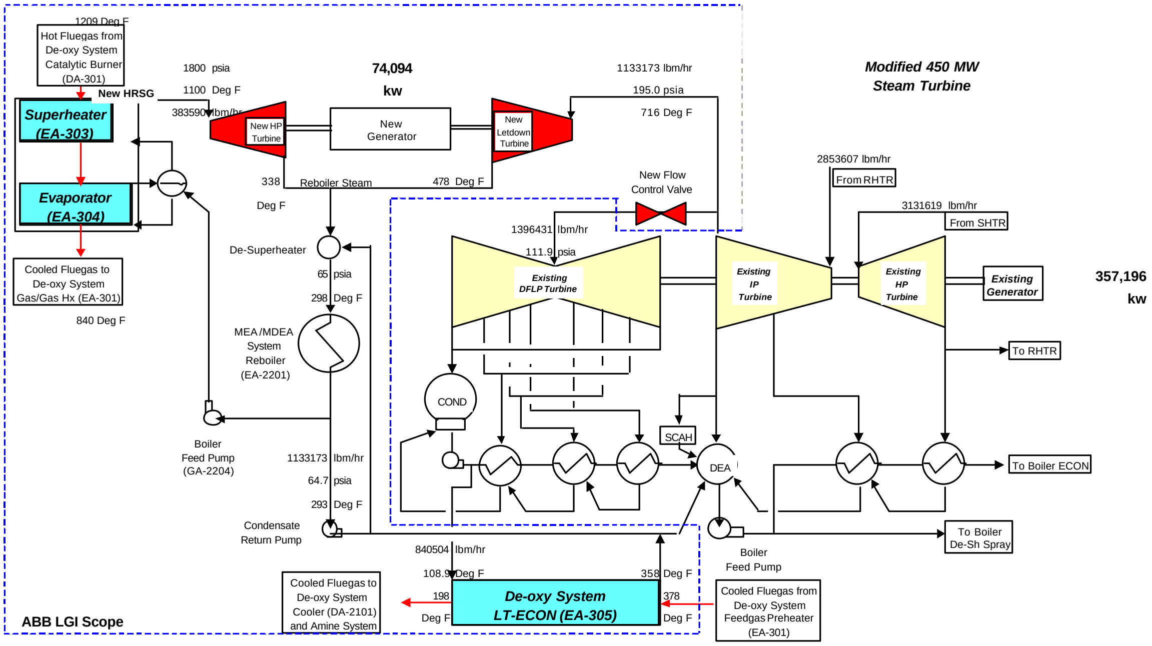
	(units)	**Original Plant (Base)**	**Concept C MEA-MDEA**
Fuel Paramaters			
Coal Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4228.7
Natural Gas Heat Input (HHV)	(10^6 Btu/hr)	---	885.9
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	4228.7	5114.6
Steam Cycle Paramaters			
Existing Steam Turbine Generator Output	(kW)	463478	357196
CO2 Removal System Turbine Generator Output	(kW)	0	36343
Total Turbine Generator Output	(kW)	463478	431290
Total Auxiliary Power	(kW)	29700	95317
Net Plant Output	(kW)	433778	335973
Overall Plant Performance Paramaters			
Net Plant Efficiency (HHV)	(fraction)	0.3501	0.2242
Net Plant Efficiency (LHV)	(fraction)	0.3666	0.2371
Normalized Efficiency (HHV; Relative to Base Case)	(fraction)	1.0000	0.6404
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9749	15223
Net Plant Heat Rate (LHV)	(Btu/kwhr)	9309	14395
Overall Plant CO_2 Emissions			
Carbon Dioxide Emissions	(lbm/hr)	866102	92153
Specific Carbon Dioxide Emissions	(lbm/kwhr)	1.997	0.274
Normalized Specific CO2 Emissions (Relative to Base Case)	(fraction)	1.000	0.137
Avoided Carbon Dioxide Emissions (as compared to Base)	(lbm/kwhr)	---	1.722
Specific Carbon Dioxide Emissions	(kg/kwhr)	0.906	0.125
Avoided Carbon Dioxide Emissions (as compared to Base)	(kg/kwhr)	---	0.782

3.4.2 Boiler Analysis

The existing boiler system performance in Case C is identical to that of the Base Case described previously in Section 2.3 and will not be repeated here.

3.4.3 Steam Cycle Modifications and Performance

The steam cycle system for Concept C is modified as shown in Figure 3.4.2. Figure 3.4.3 shows the associated Mollier diagram for this system. About 45 percent of the IP turbine exhaust is extracted from the IP/LP crossover pipe. This steam is expanded to about 65 psia through a new letdown steam turbine generating 36,343 kw. The exhaust from the letdown turbine, at about 478 °F, is de-superheated and then provides most of the heat requirement for solvent regeneration in the re-boilers of the MEA/MDEA CO_2 removal system. The condensate leaving the re-boilers is split. About 20 percent of the condensate provides feedwater for the de-oxy system heat recovery steam generator system. The remainder is pumped to the Deaerator of the existing steam cycle. High temperature heat recovery is provided in the de-oxy by the generation of high pressure superheated steam. This steam is then expanded through a second new steam turbine for additional power generation. This turbine generates 37,751 kw. The exhaust from this turbine provides about 20 percent of the feed for the re-boilers of the MEA/MDEA system. Low temperature heat recovery is provided in the de-oxy system with a low pressure feedwater heater which is located in a feedwater stream which is in parallel with the three existing low pressure extraction feedwater heaters. This heat exchanger is located in the gas stream leaving the gas to gas heat exchanger. About 92 percent of the low-pressure feedwater leaving the main condenser flows through this new heat exchanger with the remainder flowing through the existing extraction feedwater heaters. The modified existing steam cycle system produces 357,196 kw. The total output from the modified steam cycle is 431,290 kw. This represents a gross output reduction of 32,188 kW (about 7%) as compared to the Base Case.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Steam Turbine Power Output (Existing and New Turbines)	1496	Boiler Heat Input	3707
Steam Coil Air Heater Heat Output	0	De-Oxy System Heat Recovery	703
MEA / MDEA System Reboiler Heat Output	1503	Condensate Pump	1
Condenser Heat Loss	1412	**Total Energy Input**	4412
Total Energy Output	4411	**In - Out**	0

Figure 3.4. 2: Modified Steam Cycle Diagram and Performance for Concept C

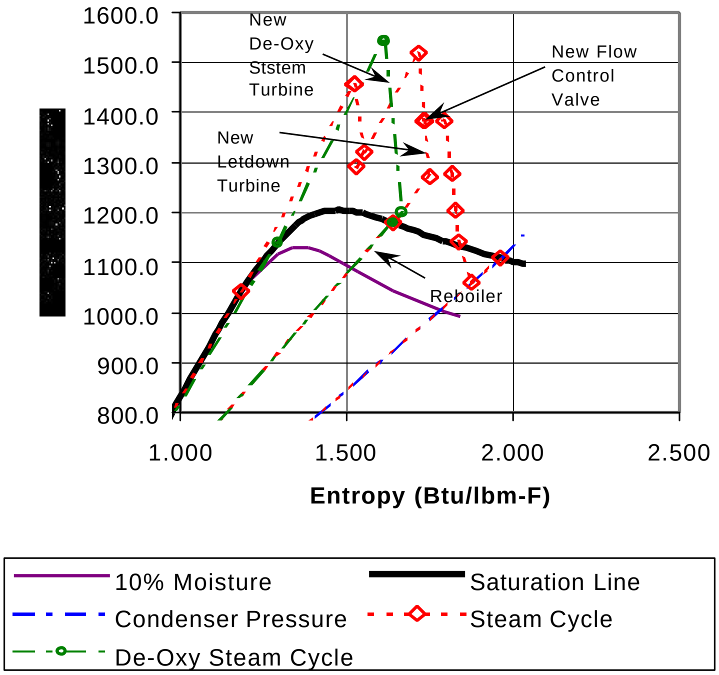


Figure 3.4. 3: Modified Steam Cycle Mollier Diagram for Concept C

3.4.4 Flue Gas Desulfurization System Analysis

The Flue Gas Desulfurization System performance in Case C is identical to that of the Base Case described previously in Section 2.3.5 and will not be repeated here. Table 3.4.3 identifies the assumptions that were made in predicting the FGD performance. Table 3.4.4 shows the gas constituents at the existing Absorber inlet and outlet locations. Results show a CO_2/SO_2 mole ratio of 63 and an SO_2 removal efficiency of 94.8%, corresponding to a value of 104 vppm at the outlet of the absorber.

Table 3.4. 3: FGD Assumptions

Concept C (MEA/MDEA)		
Quantity	Unit	Existing Absorber
Ca/S)	Mol Ratio	1.04
Solids	Wt.%	20
CaO	Wt.%	90
MgO	Wt.%	5
Inerts	Wt.%	5
Bypass Leakage	Wt.%	2.5
Liquid/Gas (L/G) Ratio	gpm/1000 acfm	55
SO_2 Removal Efficiency		
APC	%	94.8
Absorber	%	97.2

Table 3.4. 4: FGD Performance

	Concept C (MEA/MDEA)					
	Existing Absorber Inlet			Existing Absorber Outlet		
Species	Mol/hr	Vol.%	Unit	Mol/hr	Vol.%	Unit
O_2	4,469	3.14	Vol.%	4,461	2.91	Vol.%
N_2	105,018	73.74	Vol.%	105,018	68.44	Vol.%
H_2O	12,863	9.03	Vol.%	24,228	15.79	Vol.%
CO_2	19,743	13.86	Vol.%	19,720	12.85	Vol.%
SO_2	315	2,212	vppm	16	104	vppm
SO_2 Removal Efficiency, %					94.9	
CO_2/SO_2 Mole Ratio		63				

3.4.5 Carbon Dioxide Separation and Compression System

One important aspect of this project was to evaluate a new technology to separate the CO_2 with the intent of finding a more cost-effective process. Initially triethanolamine (TEA) was proposed for the solvent due to its low energy requirement to remove the absorbed CO_2. Preliminary calculations made on TEA showed that it had poor performance removing CO_2 at atmospheric pressure. Calculations made on compressing the flue gas stream in order to use TEA showed that energy requirements were excessive and that a different solvent should be found. The result was that a 10/25-weight % mixture of monoethanolamine (MEA) and methyldiethanolamine (MDEA) was to be employed. This combination has the benefit of the low energy requirement for CO_2 stripping of a tertiary amine with the aggressiveness to capture CO_2 at atmospheric pressure of MEA. The energy requirement for the removal of CO_2 (about $3.4\text{x}10^6$ Btu per ton of CO_2) is about 72% of what was required for Concept A with MEA. However, MDEA lacked both oxygen resistance and SO_2 resistance. Thus a process had to be developed to reduce both of these components to very low levels.

Burner manufacturers, Corporate Development Labs, and Catalyst Vendors were contacted. In the end only one catalyst vendor was willing to put forth the resources to demonstrate that they could meet our process requirements. CRI Catalyst, affiliated with the Shell Oil Company, modified their NOx reduction catalyst to create a low-pressure drop surface to promote the reaction between the excess oxygen present in the flue gas and natural gas added to the mixture. In reality the bench tests were done with ethylene instead of natural gas due to availability at the testing facility. However, identical results for oxygen removal (to below 100 ppmv) with natural gas were promised by CRI. The resulting process, reflected in the following pages, shows the most thermally efficient process to get the flue gas heated above its minimum inlet temperature, 550 ^{o}F, by the use of a feed/effluent exchanger.

3.4.5.1 Process Description

The numbering practice for Concept C is the same as for Concept A. The process description for the amine absorption and stripping in Concept C is essentially the same as of Concept A except that Concept C uses a less energy intensive amine than Concept A. Thus the reboiler duty will be less per unit of CO_2 absorption. Also prior to selecting which amine was to be used for Concept C, it was decided to remove essentially all of the oxygen in the flue gas stream. Tertiary amines cannot tolerate oxygen and form oxalic, acetic, propionic, glycolic, and formic acids, which are corrosive. Thus, there will be no chemical injection package for Concept C. Besides the process differences, the oxygen removal steps have an impact on the material balance. The de-oxy catalyst reactor enables the injection of methane to combine with the 2.94% of the flue gas stream that is oxygen. The process of combusting methane creates 11 % more CO_2 than in Concept A and also requires the addition of additional heat recovery equipment. The hydraulics of the equipment sized in Concept A were such that this volume of gas can be tolerated in the same size equipment. The hydraulics in this Concept are also helped by the fact that there is less liquid circulation required due to the higher amine concentration.

Thus the process flow for Concept C proceeds as follows:

Flue Gas De-Oxygenation and Cooling Systems:

Please refer to Figures 3.4.4 and 3.4.5 (drawings D 09484-01006R-0; D 09484-01007R-0)

Flue gas supplied from the existing Flue Gas Desulfurizer (FGD) is at 14.7 psia and 136 F. It goes through the Flue Gas Blower, GB-301 where it is boosted by 2.4 psi which increases its temperature to 173 °F. The flue gas is then delivered to the Bottom Direct Contact Cooler DA-2103 to be cooled to 140 °F. The heat from this process stream goes to the new cooling tower. This is done because the flow to one train is at the maximum capacity of the largest skid built by the Flue Gas Scrubber manufacturer with which there is the most experience. The gas then proceeds to Flue Gas Scrubber, D-301X, a two stage scrubbing system. The active component of the scrubbing solution is sodium sulfite, which reacts with sulfur dioxide to form sodium bisulfite. The resulting bisulfite is converted sodium sulfite with make-up caustic prior to recirculation to the scrubber. It is also equipped with a pressure control damper that maintains a constant inlet pressure to the scrubber. The scrubbing liquid used in this stage is fed to a common manifold that feeds the quench and the first stage of scrubbing. Flue gas traverses through the two-stage scrubber. The first stage includes three baffles and a Chevron type mist eliminator. In the second stage, three more baffles are encountered and scrubbing with sodium sulfite is introduced at a constant pH. The liquid level controller is used to control blowdown from the second stage to the first stage. Each section has separate pH controllers for caustic addition. The water spray in spray washing the mist eliminators also serves as make-up. The purpose of this equipment is to reduce the inlet SO_2 level from 108 ppmv down to 1 ppmv.

The gas then flows to a feed/ effluent exchanger, EA-301, where it exchanges heat with the de-oxy catalyst reactor effluent. Next the gas proceeds to a mixing chamber where the stochiometric quantity of natural gas required for the complete combustion of the oxygen is flow ratio controlled into the flue gas stream. A temperature of 628 °F is sufficient to start the reaction with the natural gas. The heat from the exothermic reaction is not available until the reaction has been established, so a start-up heater is also required. This heater, FH-302 is placed in series with the feed/ effluent exchanger, but it is normally turned off. It has 33 % of the duty of the feed/ effluent exchanger and is designed to initially heat the reactors using 33% of the design flow rate. The other 67 % of the flow continues to flow up the stack until the reactors reach operating temperature. However, as the gas leaving the de-oxy catalyst reactor increases in temperature as the reactors absorb the heat from the start-up heater, a larger gas flow can be put through the heater. Note that during the start-up mode natural gas is not mixed in with the flue gas.

All of the piping downstream of the feed/effluent exchanger through the start-up heater and reactor loop back to the other side of the feed effluent exchanger is refractory lined. The De-oxy Catalyst Reactor, DA-301, is refractory lined also. It is actually two reactors in parallel. This reactor is in a vertical orientation and contains two catalyst beds, each 5 feet deep, which create a pressure drop of 0.5 psi for each reactor vessel. The flow goes into the middle of each reactor and is split into to 50% streams by having symmetrical external piping and identical internals in the reactors. Half of the flow goes through the upper bed and half of the flow goes through the lower bed. The catalyst is a hollow cylinder 6-mm OD x 2.5-mm ID x 6-mm long. The de-oxygenated flue gas exits the reactor at 1209 °F. The outlet of the exchanger is cooled back to 840°F in the EA-303 and EA-304, Reactor Effluent/ Steam Superheater & Steam Generator, respectively. These heat exchangers are similar to a Heat Recovery Steam Generator (HRSG) type heat exchanger and only have about 6 inches of water pressure drop. The pressure of the steam generated is 1800 psia due to the desire to generate power. The effluent from the steam turbine, which drives the generator, provides part of the reboiler steam for solvent regeneration in the amine stripper. The flow passes from the steam generating heat exchangers to the Feed/ Effluent Exchanger, EA-301, where it cools by exchanging heat with the flue gas coming from the Flue Gas Scrubber. The effluent outlet temperature from EA-301supplies 42.4 MM-Btu/h heat to preheat the boiler feed water in each of the five trains. This heat is used to preheat low-pressure feedwater before supplying the power plant deaerator. The heat exchange takes place in the Dry Flue Gas Cooler EA-305. The gas is further cooled in Direct Contact Cooler (DCC) Flue Gas Cooler DA-2101. From the DCC cooler the gas then continues on to the CO_2 Absorber.

The DCC Flue Gas Cooler DA-2101 is a packed column where the hot flue gas flowing up is brought into an intimate contact with cold water which is fed to the top of the bed and flows down the tower. Physically, DA-2101, DA-2102, and DA-2103 have been combined into a single, albeit compartmentalized tower. DA-2103 is the bottom one and will be designed to support the top portion of this column. However, since there is intervening equipment between this equipment and the DCC Flue Gas cooler, DA-2101, the top head must have an outlet nozzle added to it so that the outlet pipe can exist through the section of skirt between these two vessel sections. The top of the DCC Flue Gas Cooler is configured differently. Since the flow from it proceeds directly to the amine absorption section above it, the top head of this section serves as the bottom head for the CO_2 Absorber. Effectively, this dividing head acts as a chimney tray with a number of upward extending chimneys which provide passages so the flue gas may flow directly from the DCC into the Absorber.

Theoretically, a direct contact cooler is capable of cooling the gas to a very close approach in a short bed. When the hot gas enters the DCC, it contains water but is highly superheated. At the bottom end of the bed, the gas quickly cools down to a temperature called the “Adiabatic Saturation Temperature” (AST). The AST is the temperature the gas reaches when some of its own heat content has been used to vaporize just the exact amount of water to saturate the gas.

Up to the point when the AST is reached, the mass flow of the gas stream increases due to evaporation of water. At the AST, water begins to condense. And, as the gas travels up the column and cools down further, more and more water is condensed. This internal refluxing increases the V/L traffic at the bottom end of the bed significantly beyond the external flows and must be considered in the hydraulic design.

The water stream which leaves the bottom of the DCC contains the water fed to the top as well as any water which has condensed out of the flue gas. The condensed water may be somewhat corrosive due to sulfur and nitrogen oxides that may be present in the flue gas. Therefore, instead of using the condensate in the process, it will be blown down from the system. For the DCC to be effective, the temperature of the leaving water must always be lower than the AST.

DCC Water Pump GA-2102A/B circulates most of the water leaving the bottom of the DCC back to the top of the direct contact cooler. However, before sending it back to the column the water stream is first filtered in DCC Water Filter FD-2101 and then cooled in DCC Water Cooler EA-2101. The source of this cooling water is the new cooling tower. Temperature of the cooled water is controlled by a cascade loop which maintains a constant flue gas exit temperature (Absorber feed temperature). The circulating water is cooled to 95 ^{o}F, which in turn, easily cools the gas to 115 ^{o}F.

Filtration is necessary to remove any particulate matter that may enter the DCC in the flue gas. The blowdown is taken out after the filter but before the cooler and mixed into the return water of cooler EA-2101. This way the cooler does not have to handle the extra duty that would otherwise be imposed by the blowdown.

The process description of all other equipment (CO_2 absorption, solvent stripping, and CO_2 compression and liquefaction) is essentially the same as in Concept A and will not be repeated here. Refer to Section 3.2.5.2 for the process description of this equipment. One small exception to note is a difference in the lean amine cooling. In Concept C the lean amine is cooled in two steps against cooling water downstream of the lean-rich exchanger. The warmest cooling water exchanger (EA-2208) is fed from the new cooling tower. The final cooler (EA-2202) is fed from the existing plant cooling tower. The solvent stripper CW condenser (EA-2206) also gets water from the existing plant cooling tower. Another exception for Concept C is the lean and rich amine loading. For Concept C the loadings are 0.148 mole CO_2/mole of amine and 0.445 mole CO_2/mole of amine for the lean and rich streams respectively. Please refer to Figures 3.4.6 and 3.4.7 (drawings D 09484-01008R-0; D 09484-01009R-0) for process flow diagrams of these systems for Concept C.

3.4.5.2 Process Flow Diagrams:

The processes described above for Concept C are illustrated in the following Process Flow Diagrams.

Figure 3.4.4: Drawing D 09484-01006R-0: De-Oxy System
Figure 3.4.5: Drawing D 09484-01007R-0: Flue Gas Cooling and Absorption
Figure 3.4.6: Drawing D 09484-01008R-0: Solvent Stripping
Figure 3.4.7: Drawing D 09484-01009R-0: CO_2 Compression and Liquefaction

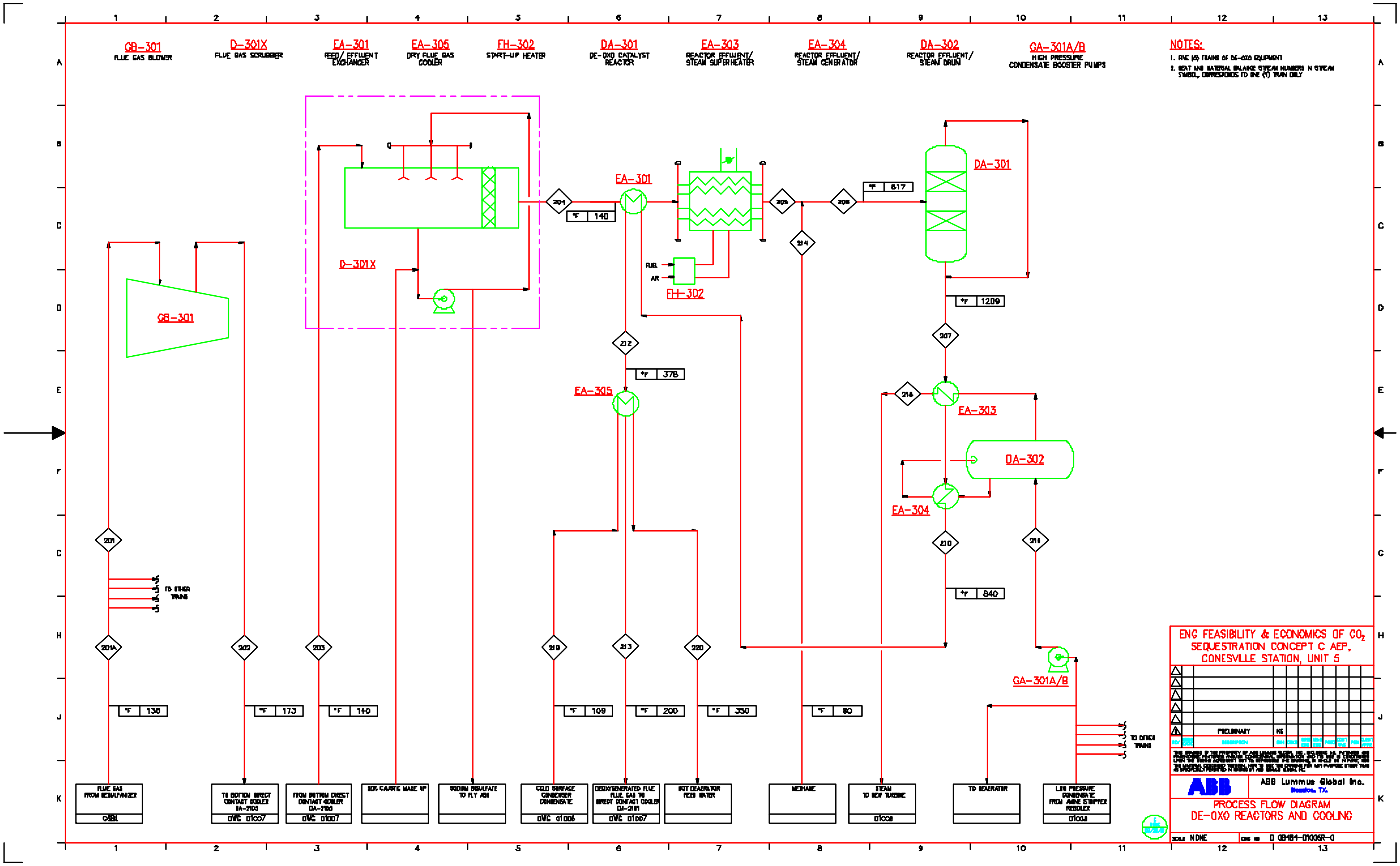


Figure 3.4. 4: De-Oxy System Process Flow Diagram for Concept C

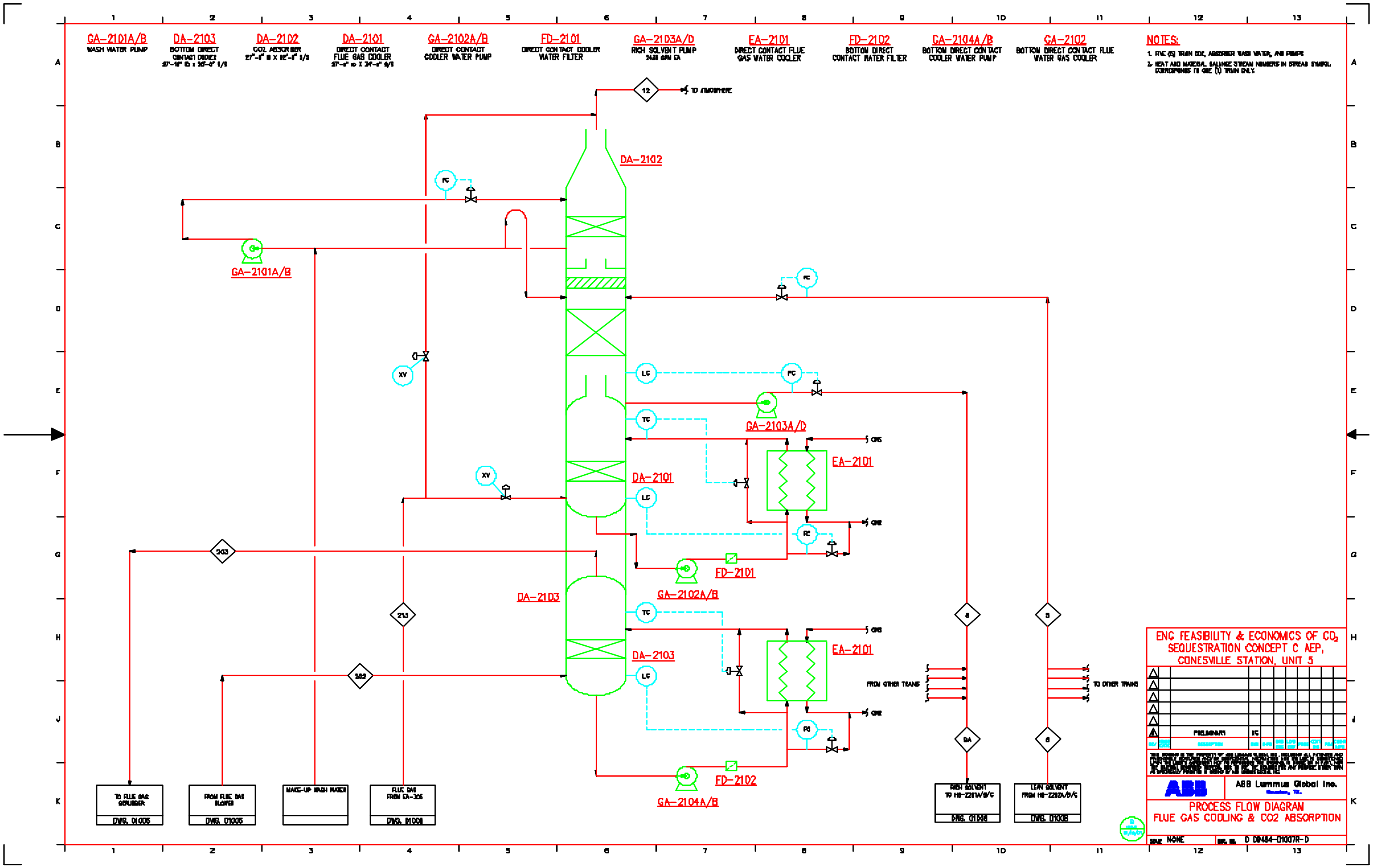


Figure 3.4. 5: Flue Gas Cooling and Absorption System Process Flow Diagram for Concept C

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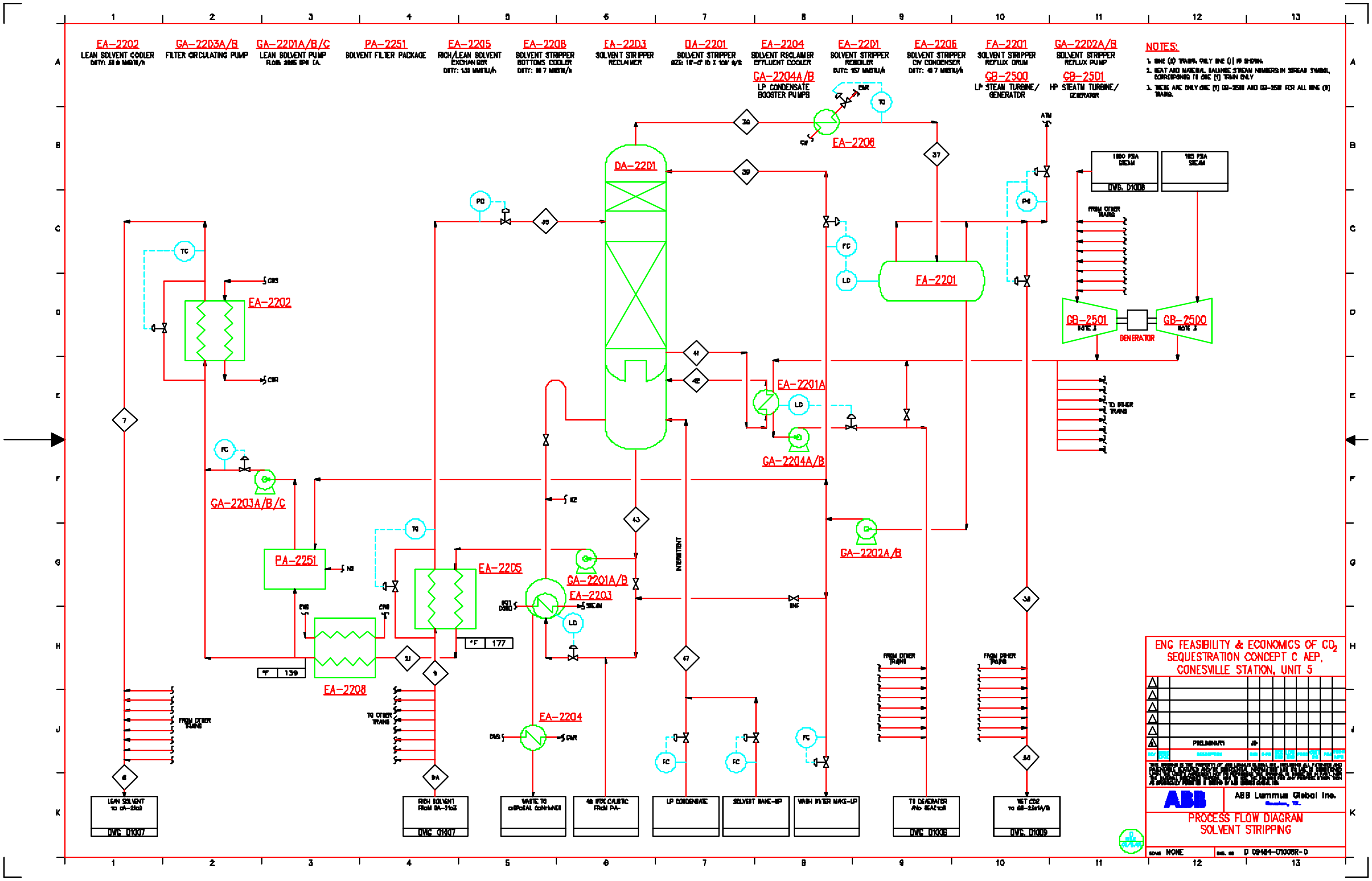


Figure 3.4. 6: Solvent Stripping System Process Flow Diagram for Concept C

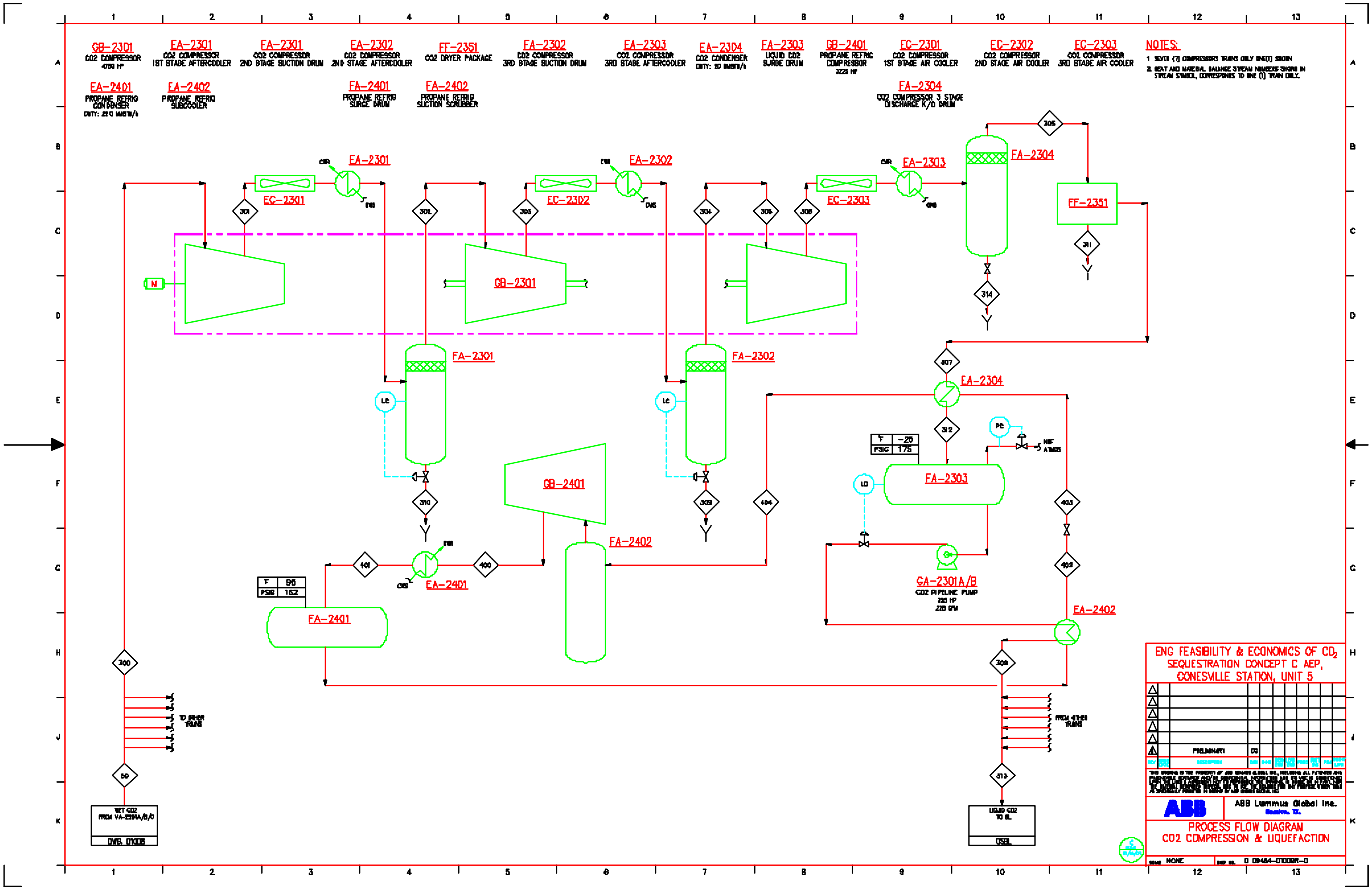


Figure 3.4. 7: CO_2 **Compression and Liquefaction System Process Flow Diagram for Concept C**

3.4.5.3 Overall Material and Energy Balance

This section contains energy and material balances for the CO_2 capture system for Concept C.

Table 3.4. 5: De-Oxy System Material and Energy Balance for Concept C

STREAM NAME		Flue gas from power plant	Flue gas to train A de-oxygenation	Flue gas from booster blower	Flue gas to desulfurization	Flue gas from desulfurization	Flue gas from feed / effluent exchanger	Methane for first deoxo reactor	First de-oxo reactor inlet	First de-oxo reactor outlet	
STREAM NO.		**201A**	**201**	**202**	**203**	**204**	**205**	**214**	**206**	**207**	**215**
VAPOR FRACTION	Molar	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	#N/A
TEMPERATURE	F	136	136	173	140	140	628	80	618	1,209	#N/A
PRESSURE	PSIG	0	0	2	2	2	2	2	2	1	#N/A
MOLAR FLOW RATE	LbMol/Hr	153,440.00	30,688.00	30,688.00	30,688.00	30,686.00	30,686.00	451.80	31,138.00	31,138.00	#N/A
MASS FLOW RATE	Lb/Hr	4,390,041	878,008	878,008	878,008	877,796	877,796	7,248	885,044	885,042	#N/A
ENERGY	Btu/Hr	6.60E+08	1.32E+08	1.41E+08	1.33E+08	1.33E+08	2.48E+08	1.96E+06	2.50E+08	4.06E+08	#N/A
COMPOSITON	Mol %										
Methane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	1.45%	0.00%	#N/A
Oxygen		2.94%	2.94%	2.94%	2.94%	2.94%	2.94%	0.00%	2.90%	0.01%	#N/A
Nitrogen		68.45%	68.45%	68.45%	68.45%	68.45%	68.45%	0.00%	67.46%	67.46%	#N/A
H2O		15.77%	15.77%	15.77%	15.77%	15.77%	15.77%	0.00%	15.55%	18.44%	#N/A
CO2		12.83%	12.83%	12.83%	12.83%	12.83%	12.83%	0.00%	12.64%	14.09%	#N/A
SO2		0.01%	0.01%	0.01%	0.01%	0.00%	0.00%	0.00%	0.00%	0.00%	#N/A
SO3		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	#N/A
VAPOR											
MOLAR FLOW RATE	LbMol/Hr	153,440.0	30,688.0	30,688.0	30,688.0	30,686.0	30,686.0	451.8	31,138.0	31,138.0	#N/A
MASS FLOW RATE	Lb/Hr	4,390,041	878,008	878,008	878,008	877,796	877,796	7,248	885,044	885,042	#N/A
STD VOL. FLOW	MMSCFD	1,397.55	279.51	279.51	279.51	279.48	279.48	4.11	283.59	283.59	#N/A
ACTUAL VOL. FLOW	ACFM	1,109,800.0	221,960.00	203,080.00	193,174.00	196,044.00	359,640.00	2,620.60	361,460.00	577,460.00	#N/A
MOLECULAR WEIGHT	MW	28.61	28.61	28.61	28.61	28.61	28.61	16.04	28.42	28.42	#N/A
DENSITY	Lb/Ft^3	0.07	0.07	0.07	0.08	0.07	0.04	0.05	0.04	0.03	#N/A
VISCOSITY	cP	0.0173	0.0173	0.0184	0.0175	0.0175	0.0281	0.0113	0.0277	0.0379	#N/A

STREAM NAME			From reactor effluent cooling		Effluent from feed / effluent exchanger	De-oxygenated flue gas to amine	High pressure BFW to steam generator	Superheated steam from steam superheater	Condensate to Dry Gas Cooler	Hot condensate from dry gas cooler
STREAM NO.		**208**	**210**	**211**	**212**	**213**	**216**	**218**	**219**	**220**
VAPOR FRACTION	Molar	#N/A	1.000	#N/A	1.000	1.000	0.000	1.000	0.000	0.000
TEMPERATURE	F	#N/A	840	#N/A	378	200	297	1,102	109	350
PRESSURE	PSIG	#N/A	1	#N/A	1	1	1,785	1,785	195	192
MOLAR FLOW RATE	LbMol/Hr	#N/A	31,138.00	#N/A	31,138.00	31,138.00	4,300.00	4,300.00	9,224.09	9,224.09
MASS FLOW RATE	Lb/Hr	#N/A	885,046	#N/A	885,046	885,046	77,465	77,465	166,173	166,173
ENERGY	Btu/Hr	#N/A	3.08E+08	#N/A	1.92E+08	1.50E+08	-4.56E+07	5.25E+07	-1.31E+08	-8.87E+07
COMPOSITON	Mol %									
Methane		#N/A	0.01%	#N/A	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		#N/A	0.01%	#N/A	0.01%	0.01%	0.00%	0.00%	0.00%	0.00%
Nitrogen		#N/A	67.46%	#N/A	67.46%	67.46%	0.00%	0.00%	0.00%	0.00%
H2O		#N/A	18.44%	#N/A	18.44%	18.44%	100.00%	100.00%	100.00%	100.00%
CO2		#N/A	14.09%	#N/A	14.09%	14.09%	0.00%	0.00%	0.00%	0.00%
SO2		#N/A	0.00%	#N/A	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
SO3		#N/A	0.00%	#N/A	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR										
MOLAR FLOW RATE	LbMol/Hr	#N/A	31,138.0	#N/A	31,138.0	31,138.0	-	4,300.0	-	-
MASS FLOW RATE	Lb/Hr	#N/A	885,046	#N/A	885,046	885,046	-	77,465	-	-
STD VOL. FLOW	MMSCFD	#N/A	283.60	#N/A	283.60	283.60	-	39.16	-	-
ACTUAL VOL. FLOW	ACFM	#N/A	455,320.00	#N/A	299,000.00	236,160.00	-	620.17	-	-
MOLECULAR WEIGHT	MW	#N/A	28.42	#N/A	28.42	28.42	-	18.02	-	-
DENSITY	Lb/Ft^3	#N/A	0.03	#N/A	0.05	0.06	-	2.08	-	-
VISCOSITY	cP	#N/A	0.0314	#N/A	0.0229	0.0186	-	0.0335	-	-

Table 3.4. 6: Amine System Material and Energy Balance for Concept C

STREAM NAME		Gas Feed to Absorber A	Lean Amine from Stripping Section	Primary LeanAmines Feed toAbsorber A	Rich Amines from Absorber A	Rich Amine from Absorption Section	Absorber A Total Treated Gas	Rich Aminesto Lean/Rich Heat Exchanger	Rich Aminesfrom Lean/RichHeat Exchanger	RegeneratorOverhead Vapor	RegeneratorCondenser Outlet
STREAM NO.		**213**	**6**	**5**	**8**	**9A**	**12**	**9**	**35**	**36**	**37**
LIQUID FRACTION		0.000	1.000	1.000	1.000	1.000	0.000	1.000	0.989	0.000	1.000
TEMPERATURE	F	115	105	105	137	137	109	137	201	211	105
PRESSURE	PSIA	15.5	14.9	14.9	15.8	15.8	14.9	15.6	15.6	26.0	23.0
COMPONENTS											
CO2 (Carbon Dioxi	LbMol/Hr	4,383.84	9,889.26	1,977.85	5,953.08	29,765.42	408.61	3,307.27	3,307.27	2,207.25	2,207.25
MEA	LbMol/Hr	0.00	29,344.65	5,868.93	5,867.05	29,335.26	1.88	3,259.47	3,259.47	6.71	6.71
MDEA	LbMol/Hr	0.00	37,607.29	7,521.46	7,521.38	37,606.91	0.08	4,178.54	4,178.54	0.41	0.41
H2O (Water)	LbMol/Hr	2,479.20	646,705	129,341	130082	650,410	1,738.26	72,267.71	72,267.71	2,345.92	2,345.92
C1 (Methane)	LbMol/Hr	1.54	0.00	0	0	0.00	1.54	0.00	0.00	0.00	0.00
N2 (Nitrogen)	LbMol/Hr	21,005.66	0.00	0	1	4.81	21,004.70	0.58	0.58	0.58	0.58
O2 (Oxygen)	LbMol/Hr	3.09	0.00	0	0	0.00	3.09	0.00	0.00	0.00	0.00
Total Molar Flow	LbMol/Hr	27,873.34	723,546	144,709	149424	747,122	23,158.16	83,013.58	83,013.58	4,560.86	4,560.86
VAPOR											
MASS FLOW RATE	Lb/Hr	3,152,095					2,365,819			236,573	
STD. VOL. FLOW RATE	MMSCFD	253.86					210.92			41.5444444	
ACTUAL VOL. FLOW RATE	MMACFD	265.88					227.42			30.0666667	
MOLECULAR WEIGHT	MW	59.28					55.0962			34.0783333	
STD. DENSITY	Lb/Ft^3	0.149					0.1346			0.123	
GAS COMPRESSIBILITY		1.9978					1.999			1.10211111	
VISCOSITY	cP	0.035					0.0356			0.01822222	
HEAT CAPACITY	Btu/Lb-F	0.4922					0.5038			0.29622222	
THERMAL CONDUCTIVITY	Btu/Hr-ft-F	0.0284					0.0298			0.0157	
LIQUID											
MASS FLOW RATE	Lb/Hr		183,618,326	7,344,733	7721129	193,028,218		2,383,064	2,383,064		155,429
STD. VOL. FLOW RATE	GPM		35184.1	7036.82	7169.36	35846.8		3982.97778	3982.97778		279.655556
ACTUAL VOL. FLOW	GPM		35498.9	7099.78	7296.84	36484.2		4053.8	4152.81111		281.522222
MOLECULAR WEIGHT	MW		25.3776	25.3776	25.8363	25.8363		25.8363	25.8363		30.6705
STD. DENSITY	Lb/Ft^3		65.0654	65.0654	67.1353	67.1353		67.1353	67.1353		62.3637
VISCOSITY	cP		1443.164	1443.164	4263.43	4263.43		4263.431	4263.431		40525.96
HEAT CAPACITY	Btu/Lb-F		1.3621	1.3621	0.9361	0.9361		0.9361	0.5233		0.6489
THERMAL CONDUCTIVITY	Btu/Hr-ft-F		0.8545	0.8545	0.838	0.838		0.838	0.8554		0.4488

STREAM NO.		**38**	**50**	**39**	**41**	**42**	**43**	**21**	**47**	**47**
LIQUID FRACTION		0.000	0.000	1.000	1.000	0.000	1.000	1.000	1.000	1.000
TEMPERATURE	F	105	105	105	248	250	250	177	68	105
PRESSURE	PSIA	23.0	23.0	23.0	29.8	30.0	30.0	30.0	30.0	14.9
COMPONENTS										
CO2 (Carbon Dioxi	LbMol/Hr	2,205.90	19,853.09	1.40	1,852.30	750.93	1,101.37	1,101.37	0.00	1,098.81
MEA	LbMol/Hr	0.01	0.07	6.70	3,291.65	32.18	3,259.46	3,259.46	1.05	3,260.52
MDEA	LbMol/Hr	0.00	0.00	0.41	4,180.89	2.35	4,178.54	4,178.54	0.04	4,178.59
H2O (Water)	LbMol/Hr	112.29	1,010.61	2,233.58	80,554.30	8,398.87	72,155.42	72,155.42	(299.34)	71,856.12
C1 (Methane)	LbMol/Hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N2 (Nitrogen)	LbMol/Hr	0.58	5.23	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O2 (Oxygen)	LbMol/Hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Molar Flow	LbMol/Hr	2,318.78	20,868.99	2,242.09	89,879.13	9,184.33	80,694.80	80,694.80	(298.25)	80,394.03
VAPOR										
MASS FLOW RATE	Lb/Hr	176236	128475917			344216				
STD. VOL. FLOW RATE	MMSCFD	21.12	190.10			83.64				
ACTUAL VOL. FLOW RATE	MMACFD	14.54	130.90			55.18				
MOLECULAR WEIGHT	MW	47.50	427.47			22.58				
STD. DENSITY	Lb/Ft^3	0.1802	1.6220			0.0889				
GAS COMPRESSIBILITY		1.1026	9.9230			1.0958				
VISCOSITY	cP	0.0174	0.1570			0.0162				
HEAT CAPACITY	Btu/Lb-F	0.2383	2.1450			0.3553				
THERMAL CONDUCTIVITY	Btu/Hr-ft-F	0.0114	0.1030			0.0171				
LIQUID										
MASS FLOW RATE	Lb/Hr			45295	2480296		2272931	2272931	-5917	2266893
STD. VOL. FLOW RATE	GPM			79.73	4239.72		3919.88	3919.88	-10.66	3909.34
ACTUAL VOL. FLOW	GPM			80.26	4521.07		4189.88	4050.54	-10.66	3944.32
MOLECULAR WEIGHT	MW			18.18	24.84		25.35	25.35	17.85	25.38
STD. DENSITY	Lb/Ft^3			63.7434	65.6431		65.0634	65.0634	62.3123	65.0654
VISCOSITY	cP			90.1959	2243.225		1442.643	1442.643	0	1443.164
HEAT CAPACITY	Btu/Lb-F			0.6638	0.3745		0.3776	0.613	0.9817	1.3621
THERMAL CONDUCTIVITY	Btu/Hr-ft-F			0.9903	0.8891		0.8918	0.8799	0.9826	0.8545

Table 3.4. 7: CO_2 Compression and Liquefaction System Material and Energy Balance for Concept C

STREAM NAME		Total acid gas from strippers	To train A 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
STREAM NO.		**50**	**300**	**301**	**302**	**310**	**303**	**304**	**309**	**306**	**305**	**314**
VAPOR FRACTION	Molar	1.000	1.000	1.000	1.000	0.000	1.000	1.000	0.000	1.000	1.000	0.000
TEMPERATURE	F	105	105	230	95	95	236	95	95	282	90	90
PRESSURE	PSIG	4	4	25	19	19	62	56	56	191	185	185
MOLAR FLOW RATE	LbMol/Hr	20,642.53	2,948.93	2,948.93	2,877.78	71.16	2,877.78	2,840.86	36.92	2,840.86	2,817.85	23.01
MASS FLOW RATE	Lb/Hr	882,394	126,056	126,056	124,773	1,283	124,773	124,106	667	124,106	123,688	418
ENERGY	Btu/Hr	9.22E+07	1.32E+07	1.66E+07	1.25E+07	-1.03E+06	1.63E+07	1.22E+07	-5.34E+05	1.72E+07	1.16E+07	-3.33E+05
COMPOSITON	Mol %											
CO2		95.13%	95.13%	95.13%	97.48%	0.09%	97.48%	98.74%	0.18%	98.74%	99.55%	0.54%
H2O		4.84%	4.84%	4.84%	2.49%	99.91%	2.49%	1.23%	99.82%	1.23%	0.42%	99.46%
Nitrogen		0.03%	0.03%	0.03%	0.03%	0.00%	0.03%	0.03%	0.00%	0.03%	0.03%	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%
Propane		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	20,642.5	2,948.9	2,948.9	2,877.8	-	2,877.8	2,840.9	-	2,840.9	2,817.8	-
MASS FLOW RATE	Lb/Hr	882,394	126,056	126,056	124,773	-	124,773	124,106	-	124,106	123,688	-
STD VOL. FLOW	MMSCFD	188.00	26.86	26.86	26.21	-	26.21	25.87	-	25.87	25.66	-
ACTUAL VOL. FLOW	ACFM	108,994.53	15,570.65	9,177.60	8,457.57	-	4,633.12	3,910.14	-	1,781.35	1,283.82	-
MOLECULAR WEIGHT	MW	42.75	42.75	42.75	43.36	-	43.36	43.69	-	43.69	43.89	-
DENSITY	Lb/Ft^3	0.13	0.13	0.23	0.25	-	0.45	0.53	-	1.16	1.61	-
VISCOSITY	cP	0.0149	0.0149	0.0187	0.0149	-	0.0193	0.0152	-	0.0212	0.0154	-
HYDROCARBON LIQUID												
MOLAR FLOW RATE	LbMol/Hr	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	Lb/Hr	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-
DENSITY	Lb/Ft^3	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-

STREAM NAME		From drier / To condenser	Water from drier	From condenser	From product pump	From Train A liquefaction	To pipeline	Refrig compressor discharge	From refrig condenser	From subcooler	To chiller	From chiller
STREAM NO.		**307**	**311**	**312**	**308**	**309**	**313**	**400**	**401**	**402**	**403**	**404**
VAPOR FRACTION	Molar	1.000	0.726	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.174	0.996
TEMPERATURE	F	90	380	-26	-11	82	82	149	95	24	-31	-31
PRESSURE	PSIG	180	180	175	2,003	2,000	2,000	169	162	159	5	5
MOLAR FLOW RATE	LbMol/Hr	2,805.88	11.97	2,805.88	2,805.88	2,805.88	19,641.16	3,071.43	3,071.43	3,071.43	3,071.43	3,071.43
MASS FLOW RATE	Lb/Hr	123,473	216	123,473	123,473	123,473	864,308	135,441	135,441	135,441	135,441	135,441
ENERGY	Btu/Hr	1.15E+07	2.63E+04	-8.44E+06	-7.62E+06	-1.43E+06	-9.99E+06	2.28E+07	8.00E+05	-5.39E+06	-5.39E+06	1.46E+07
COMPOSITON	Mol %											
CO2		99.97%	0.00%	99.97%	99.97%	99.97%	99.97%	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%	0.00%
Nitrogen		0.03%	0.00%	0.03%	0.03%	0.03%	0.03%	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%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%
VAPOR												
MOLAR FLOW RATE	LbMol/Hr	2,805.9	8.7	-	-	-	-	3,071.4	-	-	534.9	3,058.8
MASS FLOW RATE	Lb/Hr	123,473	156	-	-	-	-	135,441	-	-	23,589	134,886
STD VOL. FLOW	MMSCFD	25.55	0.08	-	-	-	-	27.97	-	-	4.87	27.86
ACTUAL VOL. FLOW	ACFM	1,314.67	6.25	-	-	-	-	1,522.13	-	-	1,964.87	11,235.55
MOLECULAR WEIGHT	MW	44.01	18.02	-	-	-	-	44.10	-	-	44.10	44.10
DENSITY	Lb/Ft^3	1.57	0.42	-	-	-	-	1.48	-	-	0.20	0.20
VISCOSITY	cP	0.0155	0.0154	-	-	-	-	0.0099	-	-	0.0065	0.0065
HYDROCARBON LIQUID												
MOLAR FLOW RATE	LbMol/Hr	-	-	2,805.88	2,805.88	2,805.88	19,641.16	-	3,071.43	3,071.43	2,536.49	12.59
MASS FLOW RATE	Lb/Hr	-	-	123,472.59	123,472.59	123,472.59	864,308.12	-	135,440.78	135,440.78	111,851.67	555.16
STD VOL. FLOW	BPD	-	-	10,244	10,244	10,244	71,706	-	18,304	18,304	15,116	75
ACTUAL VOL. FLOW	GPM	-	-	227.86	224.14	303.64	2,125.46	-	567.94	504.18	389.85	1.94
DENSITY	Lb/Ft^3	-	-	67.56	68.68	50.70	50.70	-	29.73	33.49	35.77	35.77
MOLECULAR WEIGHT	MW	-	-	44.01	44.01	44.01	44.01	-	44.10	44.10	44.10	44.10
VISCOSITY	cP	-	-	0.1746	0.1602	0.0621	0.0621	-	0.0906	0.1330	0.1823	0.1823
SURFACE TENSION	Dyne/Cm	-	-	16.01	14.00	0.86	0.86	-	5.74	10.48	14.49	14.49

3.4.5.4 Equipment List with Data

Equipment data has been presented in the so-called “short spec” format as shown in Appendix II, which provides adequate data for a factored cost estimate.

3.4.5.5 Consumption of Utilities

Consumption of utilities for Concept C is defined in Table 3.4.8.

Table 3.4. 8: Utility Consumption for Concept C

Utility	Amount Consumed	Units
Natural Gas	21.0	MMSCFD
Steam (180 psig)	1,130,000	Lb/hr
Cooling water	67800	Gpm

Number Of Trains	Tag no.	Description	Number Operating per train	Power (ea) including 0.95 motor eff (kW)	Total all trains (kW)
7	EC-2301	CO2 compressor 1st stage air cooler	1	10	71
7	EC-2302	CO2 compressor 2nd stage air cooler	1	10	73
7	EC-2303	CO2 compressor 3rd stage air cooler	1	16	109
5	GA-2101 A/B	Wash Water Pump	1	19	95
5	GA-2102 A/B	Direct Contact Cooler Water Pump	1	55	274
5	GA-2103 A/B/C/D	Rich Solvent Pump	3	103	1542
5	GA-2104 A/B	Bottom Direct Contact Cooler Water Pump	1	10	48
9	GA-2201A/B/C	Lean Solvent Pump	2	82	1473
9	GA-2202 A/B	Solvent Stripper Reflux Pump	1	3	28
9	GA-2203 A/B	Filter Circ. Pump	1	12	107
7	GA-2301 A/B	CO2 Pipeline Pump	1	184	1288
3	GA-2501	Caustic metering pump	1	0	0
5	GA-301 A/B	HP condensate booster pumps	1	148	742
9	GA-2204 A/B	LP condensate booster pump	1	57	514
5	GB-301	Flue Gas Blower	1	2628	13142
7	GB-2301	CO2 Compressor (Motor driven)	1	3730	26110
7	GB-2401	Propane Refrig Compressor	1	2532	17727
1	GB-2500	LP steam turbine/ generator	NA	NA	NA
1	GB-2501	HP steam turbine/ generator	NA	NA	NA
7	PA-2351	CO2 Drier Package	1	166	1160
1	PA-2551	Cooling Tower	1	2972	2972
		Total			67477

3.4.5.6 Consumption of Chemicals and Desiccants

Consumption of Chemicals and Desiccants for Concept C is defined in Table 3.4.9.

Table 3.4. 9: Chemicals and Desiccants Consumption f or Concept C

Chemical	Consumption per day (lbs.)
Caustic (100%)	25200
MEA	700
Corrosion inhibitor	0
Diatomaceous earth	916
Molecular sieve	282
Sodium hypochlorite	10300
Sodium bisulfite	30

This total does not include chemicals provided by the cooling tower service people nor disposal of waste. These are handled as a component of operating costs referred to as contracted services and waste handling, respectively.

3.4.5.7 Design Considerations

The following parameters were optimized for Concept C with the objective of reducing the overall unit cost and energy requirements.

- Solvent Concentration
- Lean Amine Loading
- Rich Amine Loading
- Stripper Feed Temperature
- Absorber Temperature
- Rich /Lean Exchanger approach
- CO_2 Compressor interstage temperatures
- CO_2 Refrigeration Pressure and Temperature

A minimum of 90% recovery was targeted. The above parameters were adjusted to increase the recovery until a significant increase in equipment size and/ or energy consumption was observed. AES Corporation owns and operates a 200 STPD food grade CO_2 production plant in Oklahoma. This plant was designed and built by ABB Lummus Global as a part of the larger power station complex using coal fired boilers. This plant was started up in 1990 and has been operating satisfactorily with lower than designed MEA losses. Table 3.4.10 compares key process parameters for the Concept C design with the AES design.

Table 3.4. 10: Key Process Parameters Comparison

PROCESS PARAMETER	AEP DESIGN (Concept C)	AES DESIGN
PLANT CAPACITY TPD	10372	200
CO_2 in Feed, % mol	15.7	14.7
O_2 in Feed, % mol	100 ppm	3.4
SO_2 in Feed, % mol	1 ppm	10 PPMV (Max)
SOLVENT	MEA/ MDEA	MEA
SOLVENT CONC. %WT	10/25	15 (Actual 17-18%Wt)
LEAN LOADING	0.15	0.10
RICH LOADING	0.44	0.41
STRIPPER FEED TEMP , F	201	194
STRIPPER BOTTOM TEMP F	250	245
FEED TEMP TO ABSORBER, F	105	108
CO_2 RECOVERY %	90.5	90 (ACTUAL 96-97%)
ABSORBER PRESSURE DROP psi	1	1.4
STRIPPER PRESSURE DROP psi	0.6	4.35
R/L EXCHANGER APPROACH,F	10	50
CO_2 COMPRESSOR I/STG TEMP F	105	115
LIQUID CO_2 TEMP F	82	-13
LIQUID CO_2 PRESSURE psi (A)	2015	247

3.4.5.8 OSBL Systems

Filter Residues:
A pressure leaf filter filters a slipstream of lean amine. Diatomaceous earth is used as a filter-aid for pre-coating the leaves and as a body feed. Filter cycles depend on the rate of flow through the filter, the amount of filter aid applied, and the quantity of contaminants in the solvent. A typical composition of the filter residue is provided in Table 3.4.11 below. These will be disposed of by a contracted service hauling away the drums of spent cake.

Table 3.4. 11: Filter Residue Composition

MEA	2.6 wt.%
Total Organic Carbon	1.5 wt.%
SiO2	0.03 wt.%
Insolubles	0.03 wt.%
PH	10.0
Specific Gravity	1.0

Excess Solvent Stripper Reflux Water:
The CO_2 Recovery Facility has been designed to operate in a manner to avoid accumulation of water in the Absorber / Stripper system. Conversely, a continuous make-up stream of water is not required either. By controlling the temperature of the scrubbed flue gas to the absorber, water balance of the MEA/MDEA system can be kept in check. Excess water can accumulate in the Stripper Reflux Drum and can be reused once the system is corrected to operate in a balanced manner. Should water need to be discarded, contaminants will include CO_2 and MEA/MDEA.

Secondary Absorber Effluent:
The existing plant uses lime in its flue gas desulfurizer. In the cost estimate of this plant, it has been assumed that the existing plant disposal facilities can include the additional load of the secondary absorber (D-301X).

Cooling Tower Blowdown:
The composition limits on cooling tower blowdown are shown in Table 3.4.12:

Table 3.4. 12: Filter Residue Composition

Component	Specification
Suspended Solids	30 ppm monthly, 100 ppm daily
pH	6.5 to 9
Oil and Grease	15 ppm maximum monthly, 20 ppm maximum daily
Free Chlorine	0.035 ppm

There is a thermal limit specification for the entire river. However, the blowdown volume is too small to affect it significantly.

The relief valve discharges from the CO_2 Recovery Unit are discharged to atmosphere. No tie-ins to any flare header are necessary.

3.4.5.9 Plant Layout

The Concept C Plant Layout Drawings listed below are shown in Appendix I. This equipment requires about 8 acres of plot area.

U01-D-0207 Plot Plan – Concept C: De-Oxy Catalyst & Cooling
U01-D-0210 Plot Plan – Concept C: Flue Gas Cooling & CO_2 Absorption
U01-D-0215 Plot Plan – Concept C: Solvent Stripping
U01-D-0206 Plot Plan – Concept C: CO_2 Compression & Liquefaction
U01-D-0213 Plot Plan – Concept C: Overall Layout Conceptual Plan
U01-D-0202R Plot Plan – Concept C: Modified Overall Site Plan

Plant layout has been designed in accordance with a spacing chart called “Oil and Chemical Plant Layout and Spacing” Section IM.2.5.2 issued by Industrial Risk Insurers.

When discussing layout, the first thing to observe is that no highly flammable materials are handled within the CO_2 Recovery Unit. The open cup flash point of MEA is 200 °F and, therefore, will not easily ignite. In addition to MEA, the corrosion inhibitor is the only hydrocarbon liquid within the battery limits. The flash point of this material is higher than that of MEA and is handled in small quantities.

As the chemicals used in the process present no fire hazard, there is an opportunity to reduce the minimum spacing between equipment from that normally considered acceptable in hydrocarbon handling plants. Regardless, for the drawings that follow, standard spacing requirements, as imposed by IRI have been followed.

The plot area available for the installation of the desired equipment is small. Some equipment items are placed on structures to allow other pieces of equipment to be placed underneath them. This way pumps and other equipment associated with the Absorber can be located under the structure. Locating the pumps under the structure has been considered acceptable because the fluids being pumped are not flammable.

Noise is an issue with the flue gas fan as much as it is with compressors. Discussions with vendors suggest that it will be possible to provide insulation on the fan casing to limit noise to acceptable level. Therefore, it has been assumed that no building needs to be provided for noise reasons.

Having economized on the required plot space as noted above, it was judged not to be practical to divide up the absorbers and strippers that are required into the plot areas initially offered for this purpose (refer to Section 2.1, Figure 2.1.1). Eventually it was agreed that the units would be placed in the northeast corner of the site, about 1200-ft from the Unit 5/6 stack. By having the units all together, the flue gas duct length and MEA piping between the absorber and stripper could be minimized, although the latter impact is not nearly as important as the flue gas routing.

The caustic solution will not freeze but will become very viscous when it gets cold. It will be electric traced.

The plot plan shows a substation in the Stripper area but none for the Absorber area. The assumption is that because the electrical consumption of the Absorber equipment is small (0.23 MW) compared to the Stripper equipment, the equipment can be run directly from the auxiliary power 480 volt power system.

For the Rich/Lean Solvent Exchanger which is a plate and frame type exchanger, area estimates received from vendors based on similar conditions suggest that five units/ train be sufficient for the specified service.

4.0 SUMMARY OF SYSTEMS PERFORMANCE AND CARBON DIOXIDE EMISSIONS

This section summarizes overall system performance and CO_2 emissions. Table 4.0.1 shows a comparison of plant performance and emissions for the three CO_2 recovery concepts and the Base Case that has no CO_2 recovery system.

Table 4.0. 1: Overall Plant Performance and CO_2 Emissions Comparison

	(units)	Original Plant (Base)	Concept A MEA	Concept B O2 Fired	Concept C MEA-MDEA
~~Boiler Parameters~~					
Main Steam Flow	(lbm/hr)	3131619	3131651	3131651	3131651
Reheat Steam Flow (to IP turbine)	(lbm/hr)	2853607	2853607	2808612	2853607
Main Steam Pressure	(psia)	2535	2535	2535	2535
Main Steam Temp	(Deg F)	1000	1000	1000	1000
Reheat Steam Temp	(Deg F)	1000	1000	1000	1000
Boiler Efficiency [1]	(percent)	88.13	88.13	90.47	88.13
Flue Gas Flow leaving Economizer	(lbm/hr)	4014743	4014743	3930554	4014743
Flue Gas Temperature leaving Air Heater	(Deg F)	311	311	371	311
Coal Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4228.7	4140.0	4228.7
[1] (Includes New Oxygen Heater and Parallel FW Heater for Case B) (LHV)	(10^6 Btu/hr)	4037.9	4037.9	3953.2	4037.9
CO_2 Removal Steam System Parameters					
CO_2 Removal System Steam Pressure	(psia)	---	65	---	65
CO_2 Removal System Steam Temp	(Deg F)	---	478	---	478
CO_2 Removal System Steam Extraction Flow	(lbm/hr)	---	1935690	0	1133173
CO_2 Removal System Condensate Pressure (from reboilers)	(psia)	---	64.7	---	64.7
CO_2 Removal System Condensate Temperature	(Deg F)	---	292.7	---	292.7
Natural Gas Heat Input (HHV) [2]	(10^6 Btu/hr)	---	17.7	11.4	885.9
[2] (Includes Desicant Regeneration Case A,B,C and De-Oxy system for Case C) (LHV)	(10^6 Btu/hr)	---	16.0	10.3	798.4
Steam Cycle Parameters					
Total Heat Input to Steam Cycle [3]	(10^6 Btu/hr)	3707.4	3707.4	3745.8	4410.4
Heat Output to CO_2 Removal System Reboilers	(10^6 Btu/hr)	---	1953.0	---	1503.0
Existing Condenser Pressure	(psia)	1.23	1.23	1.23	1.23
Existing Condenser Condensate Flow	(lbm/hr)	2617295	640768	2588715	1522923
Existing Condenser Heat Loss	(10^6 Btu/hr)	2102.8	603.3	2142.6	1412.1
Existing Steam Turbine Generator Output	(kW)	463478	269341	463056	357196
CO_2 Removal System Turbine Generator Output	(kW)	0	62081	0	36343
De-Oxy System Turbine Generator Output (Concept C)	(kW)	0	0	0	37751
Total Turbine Generator Output	(kW)	463478	331422	463056	431290
~~Auxiliary Power Requirements~~					
Condensate Pump Power	(kW)	563	481	557	530
Condenser Cooling Water Pump Power	(kW)	5562	1596	5667	3735
Boiler Island Auxiliary Power (Fans & Pulverizers)	(kW)	7753	7753	7804	7753
Coal & Ash Handling System	(kW)	1020	1020	998	1020
FGD & ESP System Auxiliary Power	(kW)	8157	8157	7986	8157
Misc. Auxiliary Power (Lighting, HVAC, Trans, etc)	(kW)	6645	6645	6645	6645
Air Separation Unit Power Reqirement (Case B)	(kW)	0	0	95822	0
CO_2 Removal System Auxiliary Power	(kW)	0	50355	64229	67477
Total Auxiliary Power	(kW)	29700	76007	189709	95317
Plant Performance Parameters					
Net Plant Output	(kW)	433778	255414	273347	335973
Normalized Net Plant Output (Relative to Base Case)	(fraction)	1.00	0.59	0.63	0.77
Net Plant Efficiency (HHV)	(fraction)	0.3501	0.2053	0.2247	0.2242
Net Plant Efficiency (LHV)	(fraction)	0.3666	0.2150	0.2354	0.2371
Normalized Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.59	0.64	0.64
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9749	16626	15188	15223
Net Plant Heat Rate (LHV)	(Btu/kwhr)	9309	15872	14500	14395
~~Plant CO_2 Emissions~~					
Carbon Dioxide Produced	(lbm/hr)	866102	868137	849255	967806
Carbon Dioxide Recovered	(lbm/hr)	0	835053	796238	875653
Carbon Dioxide Emissions	(lbm/hr)	866102	33084	53016	92153
Fraction of Carbon Dioxide Recovered	(fraction)	0	0.962	0.938	0.905
Specific Carbon Dioxide Emissions	(lbm/kwhr)	1.997	0.130	0.194	0.274
Normalized Specific CO_2 Emissions (Relative to Base Case)	(fraction)	1.00	0.065	0.097	0.137
Avoided Carbon Dioxide Emissions (as compared to Base)	(lbm/kwhr)	---	1.867	1.803	1.722

[3] (Includes Boiler Heat Output for the Base Case and Case A, Boiler + PFWH for Case B, and Boiler + De-Oxy System Heat Recovery for Case C)

Several comparisons have been made in this table and throughout the report. Some of the more important comparisons are summarized in this section.

Boiler Performance

All four cases were analyzed and designed based on the boiler producing a main steam flow of 3,131,619 lbm/hr at conditions of 1000 F and 2535 psia at the turbine. This main steam flow represents the maximum continuous rating (MCR) for the unit. All four cases also provided reheat steam to the turbine at conditions of 1000 ^{o}F and 590.8 psia. The boiler performance for the Base Case, Case A, and Case C was identical. Case B, while producing the same main steam flow as the other cases had a slightly lower reheat steam flow because there was less reheat spray required for this case. Boiler efficiency for Case B (90.47%) was somewhat higher than the other cases (88.13%) due primarily to the oxygen firing and the addition of the new oxygen heater and parallel feedwater heater.

Steam Cycles

The steam cycles for the four cases were all very different. The Base Case steam cycle is a typical steam cycle with nominal conditions of 2535 psia / 1000 ^{o}F / 1000 ^{o}F. Six extraction feedwater heaters are used. The generator in this case produces 463,478 kW.

Case A differs from the Base Case in that it extracts a large portion of steam from the IP/LP crossover pipe. The extracted steam, at about 200 psia is expanded through a new turbine generating power before exhausting into the reboilers of the CO_2 recovery plant at 65 psia. The modified existing steam turbine generator produces 269,341 kW and the new turbine produces 62,081 kW for a total of 331,422 kW. The gross output is reduced by 132,056 kW, about 28 percent, for this case.

The steam cycle for Case B is quite similar to the Base Case. The system is modified slightly with the addition of a new low-pressure feedwater heater in parallel with existing extraction feedwater heaters # 52 and #53 as described in Section 3.1.2. The parallel feedwater heater (PFWH) was used to recover additional sensible heat in the flue gas as a result of reduced air heater performance with oxygen firing. The modified steam cycle system of Case B produces 463,056 kW. This is 422 kW lower than for the Base Case. The gross output for Case B is slightly lower than the Base Case due to less reheat spray being required. Nearly offsetting this is the increased output resulting from the addition of the PFWH.

Case C differs from the Base Case in that a large portion of steam is extracted from the IP/LP crossover pipe that is ultimately used for the solvent regeneration in the reboilers similar to Case A. The amount of steam extracted is however, less for Case C due to two reasons. First, less steam is required per unit of CO_2 recovered and secondly, additional steam is being provided to the reboilers from the de-oxy system. The steam extracted from the existing steam turbine is expanded through a new turbine generating 36,343 kW before exhausting into the reboilers of the CO_2 recovery plant at 65 psia. The high-pressure steam generated in the de-oxy system is expanded through a new turbine providing 37,751 kW of output. A parallel low-pressure feedwater heater, arranged similar to the one used in Case B, is used for low level heat recovery in the de-oxy system. The modified existing steam turbine generator produces 357,196 kW for a total gross output of 431,290 kW. The gross output is reduced by 32,188 kW, about 7 percent, for Case C as compared to the Base Case.

Auxiliary Power and Net Plant Output

The auxiliary power required for the Base Case is 29,700 kW or about 6.4 percent of the gross output. Net plant output is 433,778 kW. All three CO_2 capture options require large amounts of additional auxiliary power. For Concept A (MEA) auxiliary power is increased to 76,007 kW due primarily to the addition of the CO_2 compression and liquefaction system which delivers the CO_2 as a liquid at 2000 psig. This system consumes 50,355 kW. The total amount of auxiliary power represents about 23 percent of the gross output. Net plant output is reduced to 255,414 kW or about 59 percent of the Base Case output. For Concept B (oxygen fired) auxiliary power is increased to 189,709 kW due primarily to the addition of the air separation unit, which consumes 95,822 kW, and the CO_2 compression and liquefaction system which consumes 64,229 kW. The total amount of auxiliary power represents about 41 percent of the gross output. Net plant output is reduced to 273,347 kW or about 64 percent of the Base Case output. For Concept C (MEA/MDEA) auxiliary power is increased to 95,317 kW due primarily to the addition of the CO_2 compression and liquefaction system which consumes 67,477 kW. The total amount of auxiliary power for Concept C represents about 22 percent of the gross output. Net plant output is reduced to 335,973 kW or about 77 percent of the Base Case output. Net Plant Output values are compared below in Figure 4.0.1.

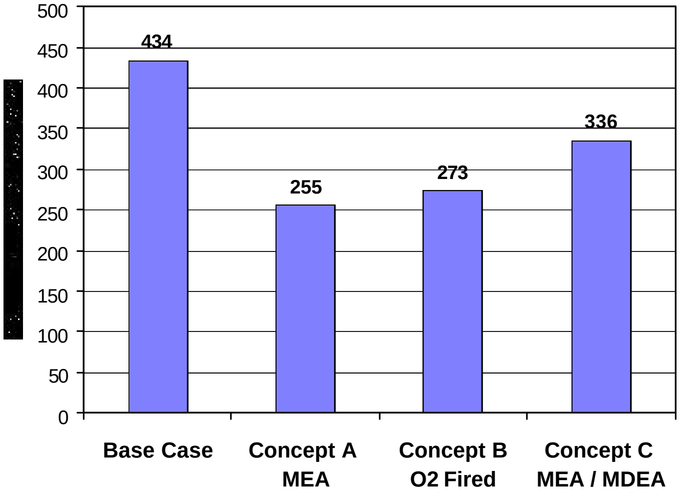


Figure 4.0. 1: Net Plant Electrical Output (Mw)

Net Plant Heat Rate and Thermal Efficiency

Because of the large energy requirements and/or auxiliary power demands for the new equipment required for the CO_2 capture systems, net plant heat rate and thermal efficiency are degraded substantially relative to the Base Case as shown in Figure 4.0.2. The Base Case plant thermal efficiency (HHV Basis) is about 35.0%, which equates to a net plant heat rate of 9,749 Btu/kwhr. Concept A, with large amounts of steam extracted for solvent regeneration and increased auxiliary power for the CO_2 compression and liquefaction system, has plant thermal efficiency degraded to about 20.5% (16,626 Btu/kwhr). Concept B, with large amounts of auxiliary power for the air separation unit and the CO_2 compression and liquefaction system, has plant thermal efficiency degraded to about 22.5% (15,187 Btu/kwhr). Concept C, with less steam extracted than Concept A for solvent regeneration and increased auxiliary power for the CO_2 compression and liquefaction system, has plant thermal efficiency degraded to about 22.4% (15,223 Btu/kwhr). These results correspond to energy penalties of 41%, 36%, and 36% for Concepts A, B, and C, respectively.

As shown in figure 4.0.1 plant output is reduced significantly with the addition of the CO_2 capture systems. Therefore replacement power is required to restore the original capacity of the unit. For cases with replacement power, it was assumed to be generated with a state-of-the-art natural gas fired combined cycle (NGCC) plant. The NGCC plant was analyzed without capturing the CO_2 from its flue gas. The NGCC plant was assumed to generate power with an efficiency of 57.1 percent (LHV basis). The additional CO_2 emitted from the NGCC was 0.762 lbm/kWh. The thermal efficiency of the three concepts including the replacement power systems is also shown in Figure 4.0.2. The efficiencies (HHV basis) range from about 26% to 28% with Concept B being the highest at 28.4%.

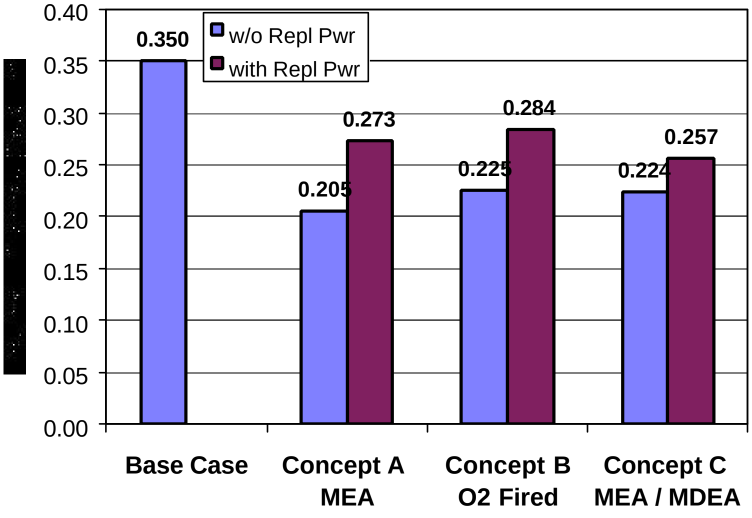


Figure 4.0. 2: Plant Thermal Efficiency (HHV Basis)

CO_2 Emissions

CO_2 emissions are summarized in Table 4.0.1 for the cases without replacement power. Specific carbon dioxide emissions were reduced from 1.997 lbm/kWh for the Base Case to 0.130, 0.194, and 0.274 lbm/kWh for Concepts A, B, and C, respectively. Figure 4.0.3 illustrates this comparison while also showing the CO_2 captured and the avoided emissions. Figure 4.0.4 compares specific CO_2 emissions both with and without replacement power. Recovery of CO_2 ranged from 91 to 96%. Normalized specific CO_2 emissions for Concepts A, B, and C respectively (without Replacement Power), were 6.5, 9.7, and 13.7 percent of the Base Case value.

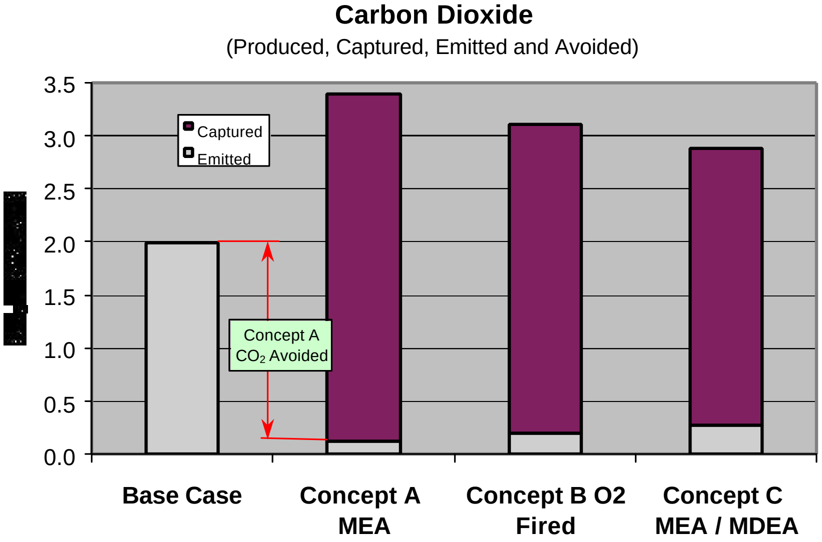


Figure 4.0. 3: Carbon Dioxide Distribution (without replacement power)

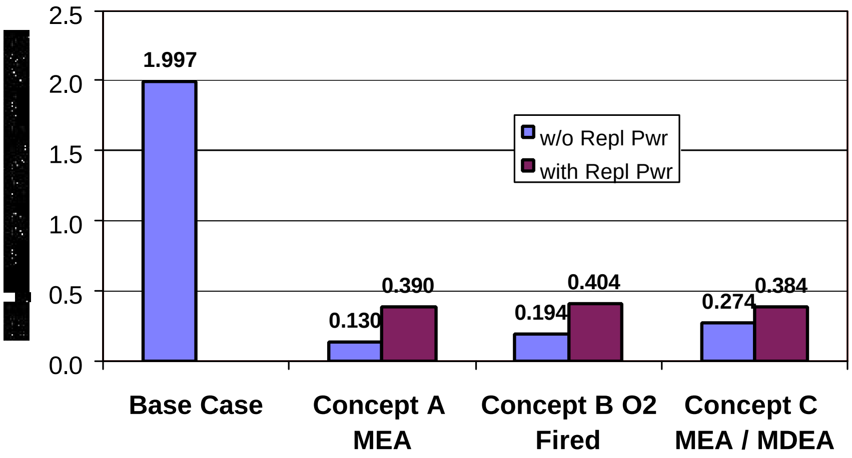


Figure 4.0. 4: **Specific Carbon Dioxide Emissions**

5.0 COST ANALYSIS

The project capital cost estimates, including engineering, procurement and construction (EPC basis), are shown in this section. These costs include all required equipment such as the amine-based CO_2 scrubbing systems of Concepts A and C, the modified FGD system of Concept A, the De-Oxy system of Concept C, the CO_2 compression and liquefaction systems for all three concepts, steam cycle modifications for all concepts, and the air separation unit and the boiler island modifications of Concept B. Boiler island modifications for Concept B are relatively small and include costs for such items as oxygen and gas ducts and dampers, booster fans, parallel feedwater heaters and piping, oxygen heater, and controls and instrumentation.

All these CO_2 capture options produce less net plant power output than the original plant (Base Case). Therefore, costs for replacement power to make up this difference were also calculated. Economic analyses discussed in Section 6 were done both with and without replacement power. For cases with replacement power, it was assumed to be generated with a state-of-the-art natural gas fired combined cycle (NGCC) plant. The NGCC plant cost and performance were developed without capturing the CO_2 from its flue gas.

Operating and maintenance (O&M) costs were calculated for all systems. The O&M costs for the Base Case were provided by American Electric Power (AEP). For the retrofit CO_2 capture system evaluations, additional O&M costs were calculated for the new equipment. The variable operating and maintenance (VOM) costs for the new equipment included such categories as chemicals and desiccants, waste handling, maintenance material and labor, and contracted services. The fixed operating and maintenance (FOM) costs for the new equipment includes operating labor only.

5.1 Cost Estimation Basis

The following assumptions were made in developing these cost estimates for each concept evaluated:

- Outdoor installation
- 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 heat exchangers designed to TEMA “C”
- All vessels are designed to ASME Section VIII, Div 1.
- Annual operating time is 5870 h/yr.
- The investment cost estimate was developed as a factored estimate based on in-house data for the major equipment. Such an estimate can be expected to have accuracy of +/-30%.
- No purchases of utilities or charges for shutdown time have been charged against the project.

Other exclusions from the cost estimate are as follows:

- 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

- Buildings except for Compressor building and electrical substation.
- Financing cost
- Owners cost
- 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 (Concept 3C)
- Excessive piling
- Contingency and risk

5.2 Boiler Modification Costs

Boiler modification costs for these three CO_2 capture options are relatively minor as compared to the other new equipment required. For this project the Boiler Scope is defined as everything on the gas side upstream of the FGD System. Therefore it includes equipment such as fans, ductwork, ESP, air heater, steam generator, pulverizers, coal and ash handling systems, etc. Not included in the boiler scope is the FGD system. The FGD system modification costs are shown separately in Section 5.4. For Concepts A and C, the Boiler Scope is not modified from the Base Case configuration and as such there are no costs in this category. Concept B Boiler Scope modifications were described in Section 3.3.4 and include such items as sealing the boiler for air leaks, new ductwork dampers and fans for the recirculated flue gas, oxygen heater, parallel low-pressure feedwater heater, and modified controls and instrumentation. The total cost required for the boiler scope modifications of Concept B is $7,000,000. This estimate includes material, engineering and construction. The expected level of accuracy for this budget level cost estimate is +/- 30%.

5.3 Carbon Dioxide Separation and Compression System Costs

5.3.1 Concept A

Investment Cost

Table 5.3.1 shows investment costs for the Concept A CO_2 Separation and Compression System. Included in this table (Acc’t. Code - 14200) are the steam cycle modification costs as well as the new letdown turbine and associated electric generator. The steam cycle modifications were described in Section 3.2.3. The Total Installed Cost (TIC) of this equipment is $393,325,000. The expected level of accuracy for this budget level cost estimate is +/- 30%.

Table 5.3.1: Concept A CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontract ($,000)	Total ($,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		44,970	697	28,481		29,178	7.4%
12000	Vessels / Filters		5,776	90	3,658		3,748	1.0%
12100	Towers / Internals		43,200	670	27,360		28,030	7.1%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		10,078	156	6,383		6,539	1.7%
14200	Compressors		100,925	1,564	63,919		65,483	16.6%
18000	Special Equipment		10,991	170	6,961		7,131	1.8%
	Sub-Total Equipment	**436**	**215,939**	**3,347**	**136,762**	**-**	**140,109**	**35.6%**
21000	Civil		287,919	4,463	9,573		14,036	3.6%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		75,579	1,171	5,607		6,779	1.7%
23000	Buildings		39,589	614	1,641		2,255	0.6%
30000	Piping		593,833	9,204	24,617		33,821	8.6%
40000	Electrical		305,914	4,742	10,941		15,683	4.0%
50000	Instruments		251,929	3,905	17,095		21,000	5.3%
61100	Insulation		215,939	3,347	7,112		10,459	2.7%
61200	Fireproofing		107,970	1,674	2,051		3,725	0.9%
61300	Painting		53,985	837	957		1,794	0.5%
	Sub-Total Commodities		**1,932,656**	**29,956**	**79,595**	**-**	**109,551**	**27.9%**
70000	**Construction Indirects**						**48,343**	**12.3%**
	Sub-Total Direct Cost		**2,148,595**	**33,303**	**216,357**	**-**	**298,003**	**75.8%**
71000	Constr. Management						2,000	0.5%
80000	Home Office Engineering						44,472	11.3%
80000	Basic Engineering						5,000	1.3%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						2,500	0.6%
19300	Spare parts						4,000	1.0%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						1,100	0.3%
97000	Freight						6,500	1.7%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**363,575**	**92.4%**
91400	Escalation						10,000	2.5%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**373,575**	**95.0%**
	Contracters Fee						19,750	5.0%
	Grand Total						**393,325**	**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.

Additional Operating Cost

Table 5.3.2 shows O&M costs for the Concept A CO_2 Separation and Compression System. They amount to $12,700,000/yr.

Table 5.3. 2: Concept A CO_2 Separation and Compression System Operating & Maintenance Costs

Operating Cost	Variable Costs Amount ($/ yr)	Fixed Costs Amount ($/ yr)
Chemical and Desiccant	2,958,000	
Waste Handling	120,000	
Utility Costs*	0*	
Maintenance (Material and Labor)	7,845,000	
Operating Labor**		1,380,000
Contracted Services	392,000	
Column Total	11,320,000	1,380,000
Grand Total (Fixed + Variable)	12,700,000	

*Included with heat rate reduction, operating expense included with power plant modifications operating cost.
** Operating labor is 365 days/ year, all other numbers are variable costs and are based on 245 days/ yr.

5.3.2 Concept B

Investment Cost

Table 5.3.3 shows investment costs for the Concept B CO_2 Separation and Compression System. Also included in this table is the cost of the Air Separation Unit ($138,544,000). The ASU System represents about 50% of the total cost for this concept whose Total Installed Cost (TIC) is $277,905,000. The expected level of accuracy for this budget level cost estimate is +/- 30%.

Table 5.3. 3: Concept B CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontract ($,000)	Total ($,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		15,060	233	9,538		9,771	3.5%
12000	Vessels / Filters		5,722	89	3,624		3,713	1.3%
12100	Towers / Internals		6,371	99	4,035		4,134	1.5%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		2,842	44	1,800		1,844	0.7%
14200	Compressors		33,098	513	20,962		21,475	7.7%
18000	Special Equipment		5,068	79	3,210		3,289	1.2%
	Sub-Total Equipment	**107**	**68,162**	**1,057**	**43,169**	**-**	**44,226**	**15.9%**
21000	Civil		102,242	1,585	3,885		5,470	2.0%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		23,857	370	1,943		2,312	0.8%
23000	Buildings		27,265	423	1,036		1,459	0.5%
30000	Piping		187,444	2,905	8,634		11,539	4.2%
40000	Electrical		96,562	1,497	3,454		4,950	1.8%
50000	Instruments		79,522	1,233	6,044		7,276	2.6%
61100	Insulation		51,121	792	1,295		2,087	0.8%
61200	Fireproofing		34,081	528	648		1,176	0.4%
61300	Painting		28,401	440	367		807	0.3%
	Sub-Total Commodities		**630,495**	**9,773**	**27,304**	**-**	**37,077**	**13.3%**
70000	**Construction Indirects**						**15,720**	**5.7%**
	Sub-Total Direct Cost		**698,656**	**10,829**	**70,473**	**-**	**97,022**	**34.9%**
	ASU TIC plant cost						138,544	49.9%
71000	Constr. Management						1,500	0.5%
80000	Home Office Engineering						10,914	3.9%
80000	Basic Engineering						1,500	0.5%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						1,750	0.6%
19300	Spare parts						2,900	1.0%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						161	0.1%
97000	Freight						2,114	0.8%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**256,405**	**92.3%**
91400	Escalation						7,500	2.7%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**263,905**	**95.0%**
	Contracters Fee						14,000	5.0%
	Grand Total						**277,905**	**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.

Additional Operating Cost

Table 5.3.4 shows O&M costs for the Concept B CO_2 Separation and Compression System. They amount to $10,643,000/yr.

Table 5.3. 4: Concept B CO_2 Separation and Compression System Operating & Maintenance Costs

Operating Cost	Variable Costs Amount ($/ yr)**	Fixed Costs Amount ($/ yr)
Chemical and Desiccant	376,000	
Waste Handling	0	
Utility Costs*	0*	
Maintenance (Material and Labor)	5,700,000	
Operating Labor**	0	261,000**
Contracted Services	1,660,000	
ASU Operating Labor**	0	902,000**
ASU Chemicals and Lubricants	237,400	
ASU Maintenance	1,506,200	
Column Totals	9,480,000	1,163,000
Grand Total (Fixed + Variable)	10,643,000	

*Included with heat rate reduction, operating expense included with power plant modifications operating cost.
** Operating labor is 365 days/ year, all other numbers are variable costs and are based on 245 days/ yr.

5.3.3 Concept C

Investment Cost

Table 5.3.5 shows investment costs for the Concept C CO_2 Separation and Compression System. Included in this table (Acc't. Code - 14200) are the steam cycle modification costs as well as the new letdown turbine and De-Oxy system turbine and associated electric generators. The steam cycle modifications were described in Section 3.4.3. The Total Installed Cost (TIC) of this equipment is $738,041,000. The expected level of accuracy for this budget level cost estimate is +/- 30%.

Table 5.3.5: Concept C CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($,000)	Material ($,000)	Subcontract ($,000)	Total ($,000)	%
11000	Heaters		3,947	61	2,500		2,561	0.3%
11200	Exchangers & Aircoolers		70,071	1,086	44,378		45,464	6.2%
12000	Vessels / Filters		6,611	102	4,187		4,289	0.6%
12100	Towers / Internals		81,947	1,270	51,900		53,170	7.2%
12200	Reactors		-	-			-	0.0%
13000	Tanks		14	0	9		9	0.0%
14100	Pumps		15,320	237	9,703		9,940	1.3%
14200	Compressors		98,073	1,520	62,113		63,633	8.6%
18000	Special Equipment		11,424	177	7,235		7,412	1.0%
	Sub-Total Equipment	**485**	**287,407**	**4,455**	**182,025**	**-**	**186,479**	**25.3%**
21000	Civil		383,209	5,940	12,742		18,681	2.5%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		100,592	1,559	7,463		9,022	1.2%
23000	Buildings		52,691	817	2,184		3,001	0.4%
30000	Piping		958,024	14,849	32,764		47,614	6.5%
40000	Electrical		407,160	6,311	14,562		20,873	2.8%
50000	Instruments		335,308	5,197	22,753		27,950	3.8%
61100	Insulation		287,407	4,455	9,465		13,920	1.9%
61200	Fireproofing		143,704	2,227	2,730		4,958	0.7%
61300	Painting		71,852	1,114	1,274		2,388	0.3%
	Sub-Total Commodities		**2,739,948**	**42,469**	**105,938**	**-**	**148,407**	**20.1%**
70000	**Construction Indirects**						**68,115**	**9.2%**
	Sub-Total Direct Cost		**3,027,355**	**46,924**	**287,963**	**-**	**403,002**	**54.6%**
71000	Constr. Management						2,500	0.3%
80000	Home Office Engineering						49,470	6.7%
80000	Basic Engineering						5,000	0.7%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						2,500	0.3%
19300	Spare parts						5,000	0.7%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals	Catalyst Excluded					222,430	30.1%
97000	Freight						8,639	1.2%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						**698,541**	**94.6%**
91400	Escalation						13,500	1.8%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						**712,041**	**96.5%**
	Contracters Fee						26,000	3.5%
	Grand Total						**738,041**	**100.0%**

Exclusions : Bonds,Taxes,Import duties , Hazerdous material handling & disposal, Capital spare parts, Reactor Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MC

Additional Operating Cost

Table 5.3.6 shows O&M costs for the Concept C CO_2 Separation and Compression System. They amount to $26,535,920/yr.

Table 5.3. 6: Concept C CO_2 Separation and Compression System Operating & Maintenance Costs

Operating Cost	Variable Costs Amount ($/ yr)**	Fixed Costs Amount ($/ yr)
Chemical and Desiccant	1,797,000	
Waste Handling	15,600	
Utility Costs*	0*	
Maintenance (Material and Labor)	22,141,320	
Operating Labor**	0	1,380,000**
Contracted Services	1,202,000	
Column Totals	25,155,920	1,380,000
Total (Fixed + Variable)	26,535,920	

*Included with heat rate reduction, operating expense included with power plant modifications operating cost.
** Operating labor is 365 days/ year, all other numbers are variable costs and are based on 245 days/ yr.

5.4 Flue Gas Desulfurization System Modification Costs

Flue Gas Desulfurization System modification costs for these three CO_2 capture options are relatively minor as compared to the other new equipment required. For Concepts B and C, the existing Flue Gas Desulfurization System is not modified from the Base Case configuration and as such there are no costs in this category. The Concept A Flue Gas Desulfurization System modifications, which include the addition of a secondary absorber island, building, booster fan, and ductwork, are described in Section 3.2.4. The total cost required for the Flue Gas Desulfurization System scope modifications of Concept A is $15,800,000. This estimate includes material, engineering and construction. The expected level of accuracy for this cost estimate is +/- 10%.

5.5 Charges for Loss of Power during Construction and Operation of the Modified Power Plant System

During the construction period it is assumed the existing Unit No. 5 power plant will be operated in its normal way. The new CO_2 capture equipment is being located some 1500 feet from the Unit No. 5/6 stack (see Appendix I for plant layout drawings) and as such will not impede operation of Unit No. 5 or any of the other units on site. Once construction is completed it has been assumed the final connections between the CO_2 capture systems and the existing power plant can be completed during the annual outage for the unit. Final shake down testing will be completed after the outage.

During plant operation, all these CO_2 capture options produce less net plant output at full load than the original plant (Base Case). Therefore, each concept was analyzed with replacement power to make up for this difference. For cases with replacement power, it was assumed to be generated with a state-of-the-art natural gas fired combined cycle (NGCC) plant. The NGCC plant performance and cost was calculated assuming the CO_2 from its flue gas was not captured. The NGCC plant was assumed to cost $450/kW, installed. Table 5.5.1 shows the NGCC outputs and costs for the three CO_2 capture concepts.

Table 5.5. 1: NGCC Outputs and Costs for Replacement Power

	Concept A	Concept B	Concept C
NGCC Power Output (kW)	178,363	160,430	97,805
NGCC Installed Cost ($ x 1000)	80,264	72,194	44,012

5.6 Summary of Total Plant Investment Costs

Table 5.6.1 summarizes the total retrofit investment costs required for each of the three Concepts. Two costs are shown for each concept, one with and one without replacement power. Figure 5.6.1 shows the specific investment costs ($/kW) for the three concepts without replacement power. Two costs are given for each of the three concepts in this figure. The bars on the left are relative to the new plant output and the bars on the right are relative to the original plant output. Figure 5.6.2 shows the specific investment costs ($/kW) for the three concepts with replacement power These costs were used in the economic evaluation (Section 6) to develop incremental Cost of Electricity values and comparisons.

Table 5.6. 1: Total Retrofit Investment Capital Costs

CO_2 Capture Concept	**Units**	w/o Repl Pwr	with Repl Pwr
Concept A	10^6 US$	409	489
	$/kW	1602	1128
Concept B	10^6 US$	285	357
	$/kW	1042	823
Concept C	10^6 US$	738	782
	$/kW	2197	1803

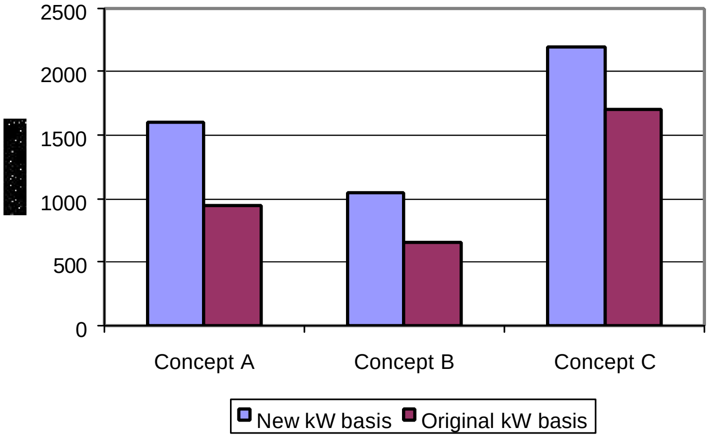


Figure 5.6. 1: New Equipment Specific Investment Costs (Without Replacement Power)

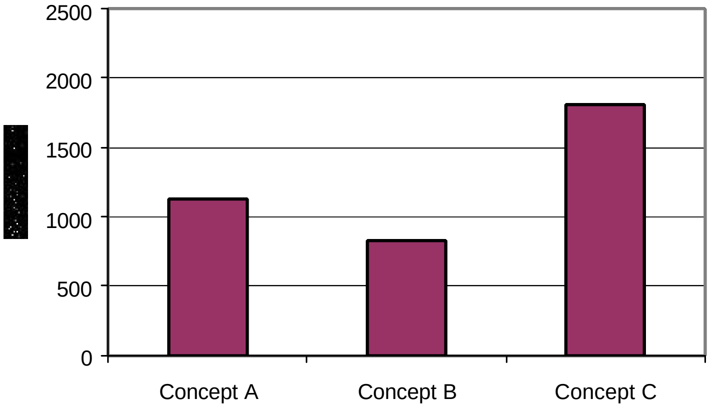


Figure 5.6. 2: New Equipment Specific Investment Costs (With Replacement Power)

6.0 ECONOMIC ANALYSIS

A comprehensive economic evaluation comparing the Base Case study unit and the three retrofit CO_2 capture concepts (**A** – MEA absorption, **B -** O_2 Fired with flue gas recycle, and **C -** MEA/MDEA absorption) was performed. The purpose of the evaluation was to quantify the impact of CO_2 capture on the Cost of Electricity (COE) of this existing coal fired unit. CO_2 mitigation costs were also determined in this analysis. The economic evaluation results are presented as incremental Costs of Electricity (first full year of operation basis). The first year basis is quite representative (at least for the input assumptions used in this study) as the post modification COE for year 10 is only about 7% greater than for year 1. The incremental costs of electricity are incremental relative to the Base Case (air fired without CO_2 capture, i.e., business as usual). Each CO_2 capture option was evaluated both with and without replacement power. Since all these CO_2 capture options produce less net plant output than the original plant (Base Case), the replacement power exactly represents this difference. For cases with replacement power, it was assumed to be generated with a state-of-the-art natural gas fired combined cycle (NGCC) plant. The NGCC plant was analyzed without CO_2 capture from its flue gases. Additionally, economic sensitivity studies were developed for the six primary cases (each of the CO_2 capture options with and without replacement power) to highlight which parameters affected the COE to the greatest extent. A total of 66 economic evaluation cases are reported.

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 capital 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.

In performing the economic evaluations, numerous assumptions were required. Table 6.0.1 lists the primary assumptions used in this analysis.

Table 6.0.1: Economic Evaluation Study Assumptions

Economic Assumptions			**Financing Terms**		
NPV Cash Discount Rate	percent	12.0	Equity	percent	50.0
Depreciable Life	years	15	Debt	percent	50.0
			Interest on Loan (APR)	percent	9.0
Project Schedule			Term of Loan	years	15
Project Start	date	9/1/01	Return on Equity	percent	12.0
Construction Period	months	30			
Evaluation Period	years	15	**Fuel Information**		
			Coal Cost	$\$/10^6$ Btu	1.32
Variable Operating Costs (Non-Fuel)			Natural Gas Cost	$\$/10^6$ Btu	4.00
Lime Cost (Delivered)	$/Ton	60			
Ash Disposal Cost	$/Ton	14			
			Plant Information		
Revenue Information			Capacity Factor	percent	67.0
Electricity Price	$/MWhr	32.00	Remaining Operating Life	years	20
CO_2 Market Price	$/Ton	0.0	Fixed O&M Costs	$/kw	16.0

The following list defines additional assumptions used in all Replacement Power Cases.

- Power Generation Technology: Natural Gas fired Combined Cycle (NGCC) without CO_2 capture
- NGCC Net Plant Heat Rate (HHV basis): 6640 Btu/kWh (GTW, 1998-1999)
- NGCC Net Plant Efficiency (LHV basis): 57.1 percent
- NGCC Investment Cost; Engineered, Procured, Constructed basis (EPC): 450 $/kW
- NGCC Fixed O&M Cost: 16.92 $/kW-yr
- NGCC Variable O&M Cost: 0.01 Cents/kWh

As is evident from the above list, the NGCC replacement power calculations were identical for all cases with the only difference between cases being the scaling of various items required for the economic evaluation as a function of output requirement. In other words a "rubber NGCC" was assumed with performance, O&M costs, and specific investment costs assumed constant and not a function of output. This was done such that all economic differences between the cases would be completely attributable to the CO_2 capture technology employed and not influenced by changes in NGCC specific costs, or performance, resulting from economy of scale.

The Economic Sensitivity Study was performed for each of six primary cases. The six primary cases include two cases each for Concepts A, B, and C (one case with and one case without replacement power). The sensitivity study shows the effect on incremental COE and on CO_2 mitigation cost of variations in five parameters of interest. CO_2 mitigation costs were calculated according to Equation (6.1) below. The five parameters varied in this study were investment cost (which included the new CO_2 capture equipment, replacement power equipment, and the book value of the existing plant), coal cost, natural gas cost, capacity factor, and CO_2 byproduct sell price. Eleven evaluations were done for each of the six cases providing a total of 66 economic evaluations from the overall sensitivity study. Table 6.0.2 shows the ranges for each of the five parameters varied in the sensitivity study. Three points were calculated for each parameter (Base Value, Minimum, and Maximum).

$$\text{Mitigation Cost} = (COE_{Cp} - COE_{Ref}) / (CO_{2\,Ref} - CO_{2\,Cp}) \qquad (6.1)$$

Where:

Mitigation Cost = \$/Ton of CO_2 Avoided
COE = Cost of Electricity (\$/kWh)
CO_2 = Carbon Dioxide Emitted (Ton/kWh)
Cp = Capture Plant
Ref = Reference Plant

Table 6.0. 2: Economic Sensitivity Study Parameters

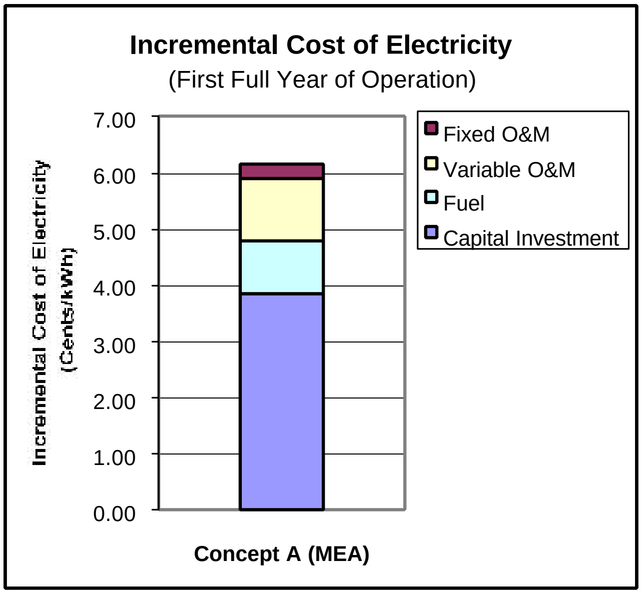
Parameter	Units	Base Value	Minimum	Maximum
Investment Cost	$	as estimated	Base - 25%	Base + 25%
Fuel Cost (Coal)	$/MMBtu	1.32	1.00	1.60
Fuel Cost (Natural Gas)	$/MMBtu	4.00	3.00	5.00
Capacity Factor	Fraction	0.67	0.60	0.75
CO_2 Byproduct Sell Price	$/ton	0.00	10.00	20.00

6.1 Concept A Results: CO_2 Separation with Monoethanolamine (MEA) Absorption

This section shows the economic analysis results for Concept A, CO_2 separation with Monoethanolamine Absorption. Results both with and without replacement power are reported including sensitivity results.

6.1.1 Concept A Results without Replacement Power

Figure 6.1.1 shows the economic analysis results using the base parameter values for Concept A (MEA) as compared with the Base Case (air fired without CO_2 capture). This analysis was done without replacement power. The first year incremental cost of electricity for this case was about 6.17 Cents/kWh greater than the Base Case COE.



Incremental Cost of Electricity Distribution	(Cents/ kWh)
Capital Investment	3.86
Fixed O&M	0.28
Variable O&M	1.10
Fuel	0.93
Total Incremental Cost of Electricity	**6.17**

Figure 6.1. 1: Concept A (MEA) Economic Results without Replacement Power

Results for the Concept A sensitivity study, without replacement power, are shown in Table 6.1.1 and Figure 6.1.2. This table and the associated figure show the sensitivity of incremental COE to coal cost, natural gas cost, capacity factor, CO_2 byproduct sell price, and new equipment installed capital cost. Results for the Base parameter values (i.e., Coal Cost = 1.32 $\$/10^6$ Btu, Natural Gas Cost = 4.0 $\$/10^6$ Btu, Capacity Factor = 67%, CO_2 Byproduct Sell Price = 0.0 \$/Ton, Investment Cost = As Estimated) are shown above in Figure 6.1.1. The base parameter values also represent the point in Figure 6.1.2 where all the sensitivity curves intersect (point 0.0, 0.0). Incremental COE ranges from a low of 2.90 to a high of 7.25 cents/kWh. The most sensitive parameters are CO_2 byproduct sell price, capital cost and capacity factor, in that order, with coal cost and natural gas cost showing significantly less impact on incremental COE. CO_2 mitigation costs ranged from 31-73 \$/Ton of CO_2 avoided with the baseline value at 66 \$/Ton of CO_2 avoided.

Table 6.1. 1: Economic Sensitivity Results for Concept A without Replacement Power

Parameter	Unit	CO_2 Capture with MEA, Without Replacement Power										
Power Output												
Net Power Output	kW	255414	255414	255414	255414	255414	255414	255414	255414	255414	255414	255414
Replacement Power	kW	0	0	0	0	0	0	0	0	0	0	0
Total Power Output	kW	255414	255414	255414	255414	255414	255414	255414	255414	255414	255414	255414
Plant Performance												
Net Plant heat Rate	Btu/kWh	16.626	16.626	16.626	16.626	16.626	16.626	16.626	16.626	16.626	16.626	16.626
Net Plant Efficiency	Fraction	0.2053	0.2053	0.2053	0.2053	0.2053	0.2053	0.2053	0.2053	0.2053	0.2053	0.2053
CO2 Emitted	lbm/h	33084	33084	33084	33084	33084	33084	33084	33084	33084	33084	33084
CO2 Emitted	lbm/kWh	0.130	0.130	0.130	0.130	0.130	0.130	0.130	0.130	0.130		
Total Capital Cost	$ (1000's)	**495881**	495881	495881	495881	495881	495881	495881	495881	495881	**371911**	**619851**
Capital Cost	$/kW	1941	1941	1941	1941	1941	1941	1941	1941	1941	1456	2427
Economic Analysis Basis												
Capital Charge Rate	%	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302
Fuel Cost (Coal)	$/MMBtu	**1.32**	**1.00**	**1.60**	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32
Fuel Cost (Natural Gas)	$/MMBtu	**4.00**	4.00	4.00	**3.00**	**5.00**	4.00	4.00	4.00	4.00	4.00	4.00
Capacity Factor	Fraction	**0.67**	0.67	0.67	0.67	0.67	**0.60**	**0.75**	0.67	0.67	0.67	0.67
CO2 Price	$/ton	**0.00**	0.00	0.00	0.00	0.00	0.00	0.00	**10.00**	**20.00**	0.00	0.00
Economic Analysis Results												
coe: Incremental	cents/kWh	6.17	5.96	6.36	6.17	6.18	6.66	5.73	4.54	2.90	5.10	7.25
CO2 Mitigation Cost	$/ton	66	64	68	66	66	71	61	49	31	51	73

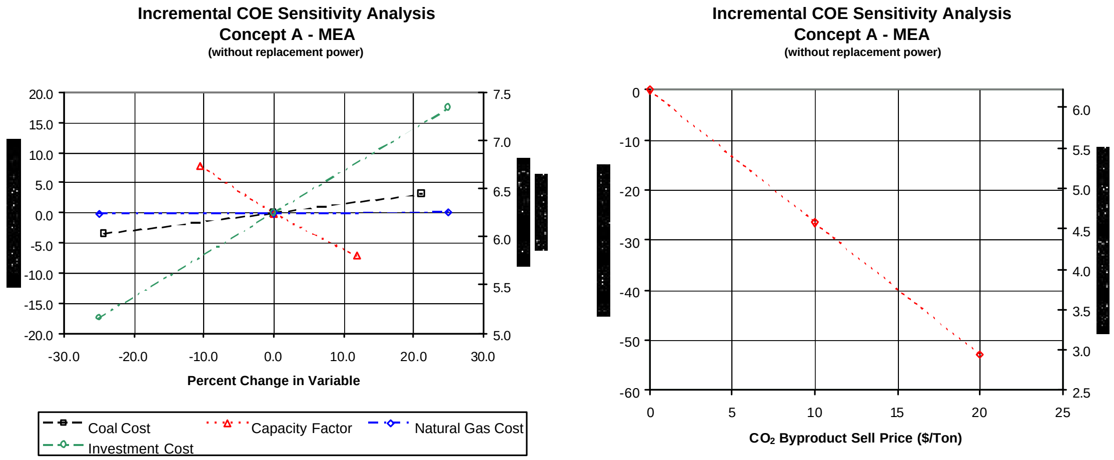
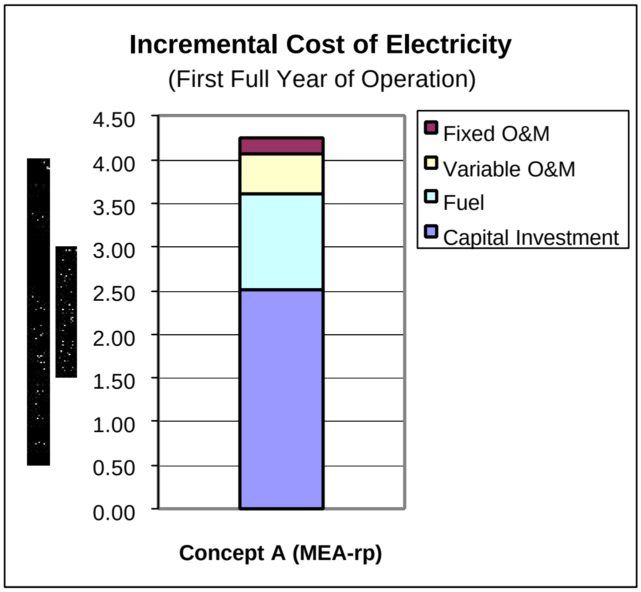


Figure 6.1. 2: Concept A (MEA) Economic Sensitivity Results without Replacement Power

6.1.2 Concept A Results with Replacement Power

Figure 6.1.3 shows the economic analysis results using the base parameter values for Concept A (MEA) with replacement power as compared with the Base Case. The first year incremental cost of electricity for this case was about 4.25 Cents/kWh greater than the Base Case COE.



Incremental Cost of Electricity Distribution	(Cents/ kWh)
Capital Investment	2.50
Fixed O&M	0.17
Variable O&M	0.47
Fuel	1.11
Total Incremental Cost of Electricity	**4.25**

Figure 6.1. 3: Concept A (MEA) Economic Results with Replacement Power

Results for the Concept A sensitivity study, with replacement power, are shown in Table 6.1.2 and Figure 6.1.4. Incremental COE ranges from a low of 2.33 to a high of 4.99 cents/kWh. The most sensitive parameters again are CO_2 byproduct sell price, investment cost and capacity factor, in that order. Natural gas cost shows significantly more impact on incremental COE when replacement power is included as would be expected with the NGCC being used to generate the replacement power. Coal cost, on the other hand, has no impact on incremental COE with replacement power since both the Base Case and Concept A use the same amount of coal and the net electrical output is also the same. CO_2 mitigation costs ranged from 29-73 \$/Ton of CO_2 avoided with the baseline value at 53 \$/Ton of CO_2 avoided.

Table 6.1. 2: Economic Sensitivity Results for Concept A with Replacement Power

Parameter	Unit	CO_2 Capture with MEA, Including Replacement Power										
Power Output												
Net Power Output	kW	255414	255414	255414	255414	255414	255414	255414	255414	255414	255414	255414
Replacement Power	kW	178363	178363	178363	178363	178363	178363	178363	178363	178363	178363	178363
Total Power Output	kW	433778	433778	433778	433778	433778	433778	433778	433778	433778	433778	433778
Plant Performance												
Net Plant heat Rate	Btu/kWh	12.520	12.520	12.520	12.520	12.520	12.520	12.520	12.520	12.520	12.520	12.520
Net Plant Efficiency	%	0.2726	0.2726	0.2726	0.2726	0.2726	0.2726	0.2726	0.2726	0.2726	0.2726	0.2726
CO2 Emitted	lbm/h	169053	169053	169053	169053	169053	169053	169053	169053	169053	169053	169053
CO2 Emitted	lbm/kWh	0.390	0.390	0.390	0.390	0.390	0.390	0.390	0.390	0.390	0.390	0.390
Total Capital Cost	$ (1000's)	**576144**	576144	576144	576144	576144	576144	576144	576144	576144	**432108**	**720180**
Capital Cost	$/kW	1328	1328	1328	1328	1328	1328	1328	1328	1328	996	1660
Economic Analysis Basis												
Capital Charge Rate	%	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302
Fuel Cost (Coal)	$/MMBtu	**1.32**	**1.00**	**1.60**	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32
Fuel Cost (Natural Gas)	$/MMBtu	**4.00**	4.00	4.00	**3.00**	**5.00**	4.00	4.00	4.00	4.00	4.00	4.00
Capacity Factor	Fraction	**0.67**	0.67	0.67	0.67	0.67	**0.60**	**0.75**	0.67	0.67	0.67	0.67
CO2 Price	$/ton	**0.00**	0.00	0.00	0.00	0.00	0.00	0.00	**10.00**	**20.00**	0.00	0.00
Economic Analysis Results												
coe: Incremental	cents/kWh	4.25	4.25	4.25	3.97	4.53	4.56	3.97	3.29	2.33	3.52	4.99
CO2 Mitigation Cost	$/ton	53	53	53	49	56	57	49	41	29	44	62

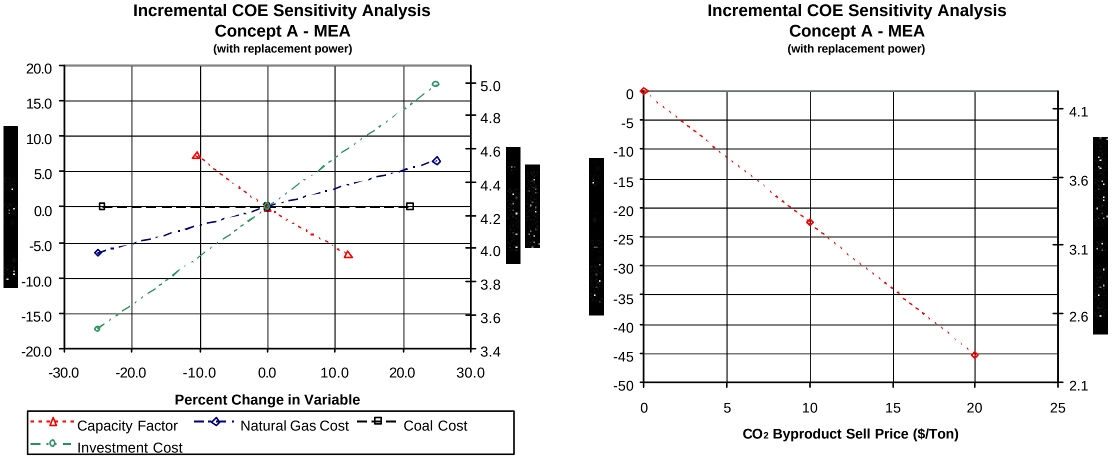


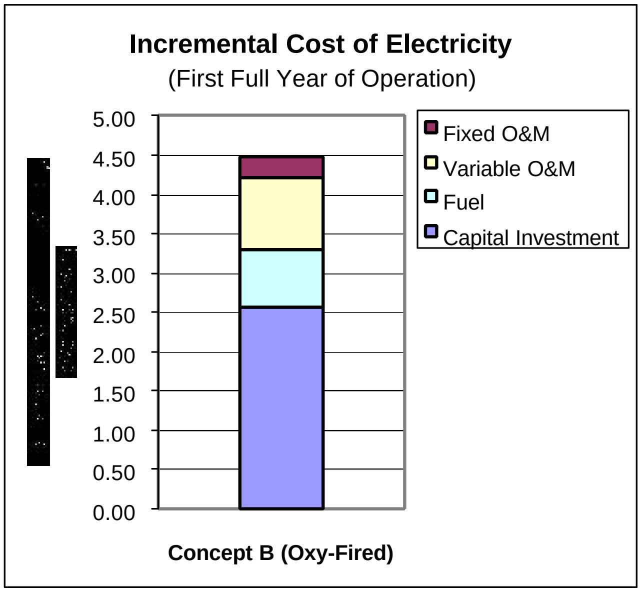
Figure 6.1. 4: Concept A (MEA) Economic Sensitivity Results with Replacement Power

6.2 Concept B Results: CO_2 Separation with Oxygen Firing and Flue Gas Re-circulation

This section shows the economic analysis results for Concept B, CO_2 Separation with Oxygen Firing and Flue Gas Re-circulation. Results both with and without replacement power are also reported along with sensitivity results.

6.2.1 Concept B Results without Replacement Power

Figure 6.2.1 shows the economic analysis results using the base parameter values for Concept B (Oxygen Fired) as compared with the Base Case (air fired without CO_2 capture). This analysis was done without replacement power. The first year incremental cost of electricity for this case was about 4.45 Cents/kWh greater than the Base Case COE.



Incremental Cost of Electricity Distribution	(Cents/ kWh)
Capital Investment	2.57
Fixed O&M	0.24
Variable O&M	0.91
Fuel	0.73
Total Incremental Cost of Electricity	**4.45**

Figure 6.2. 1: Concept B (Oxygen Fired) Economic Results without Replacement Power

Results for the Concept B sensitivity study, without replacement power, are shown in Table 6.2.1 and Figure 6.2.2. Incremental COE ranges from a low of 1.53 to a high of 5.20 cents/kWh. The most sensitive parameters are, again, CO_2 byproduct sell price, investment cost and capacity factor, in that order, with coal cost and natural gas cost showing significantly less impact on incremental COE. CO_2 mitigation costs ranged from 17-58 \$/Ton of CO_2 avoided with the baseline value at 49 \$/Ton of CO_2 avoided.

Table 6.2. 1: Economic Sensitivity Results for Concept B without Replacement Power

Parameter	Unit	CO_2 Capture with O2 Firing, Without Replacement Power										
Power Output												
Net Power Output	kW	273347	273347	273347	273347	273347	273347	273347	273347	273347	273347	273347
Replacement Power	kW	0	0	0	0	0	0	0	0	0	0	0
Total Power Output	kW	273347	273347	273347	273347	273347	273347	273347	273347	273347	273347	273347
Plant Performance												
Net Plant heat Rate	Btu/kWh	15.188	15.188	15.188	15.188	15.188	15.188	15.188	15.188	15.188	15.188	15.188
Net Plant Efficiency	Fraction	0.2247	0.2247	0.2247	0.2247	0.2247	0.2247	0.2247	0.2247	0.2247	0.2247	0.2247
CO2 Emitted	lbm/h	53016	53016	53016	53016	53016	53016	53016	53016	53016	53016	53016
CO2 Emitted	lbm/kWh	0.194	0.194	0.194	0.194	0.194	0.194	0.194	0.194	0.194	0.194	0.194
Total Capital Cost	$ (1000's)	**371661**	371661	371661	371661	371661	371661	371661	371661	371661	**278746**	**464576**
Capital Cost	$/kW	1360	1360	1360	1360	1360	1360	1360	1360	1360	1020	1700
Economic Analysis Basis												
Capital Charge Rate	%	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302
Fuel Cost (Coal)	$/MMBtu	**1.32**	**1.00**	**1.60**	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32
Fuel Cost (Natural Gas)	$/MMBtu	**4.00**	4.00	4.00	**3.00**	**5.00**	4.00	4.00	4.00	4.00	4.00	4.00
Capacity Factor	Fraction	**0.67**	0.67	0.67	0.67	0.67	**0.60**	**0.75**	0.67	0.67	0.67	0.67
CO2 Price	$/ton	**0.00**	0.00	0.00	0.00	0.00	0.00	0.00	**10.00**	**20.00**	0.00	0.00
Economic Analysis Results												
coe: Incremental	cents/kWh	4.45	4.27	4.60	4.44	4.45	4.77	4.15	2.99	1.53	3.69	5.20
CO2 Mitigation Cost	$/ton	49	47	51	49	49	53	46	33	17	41	58

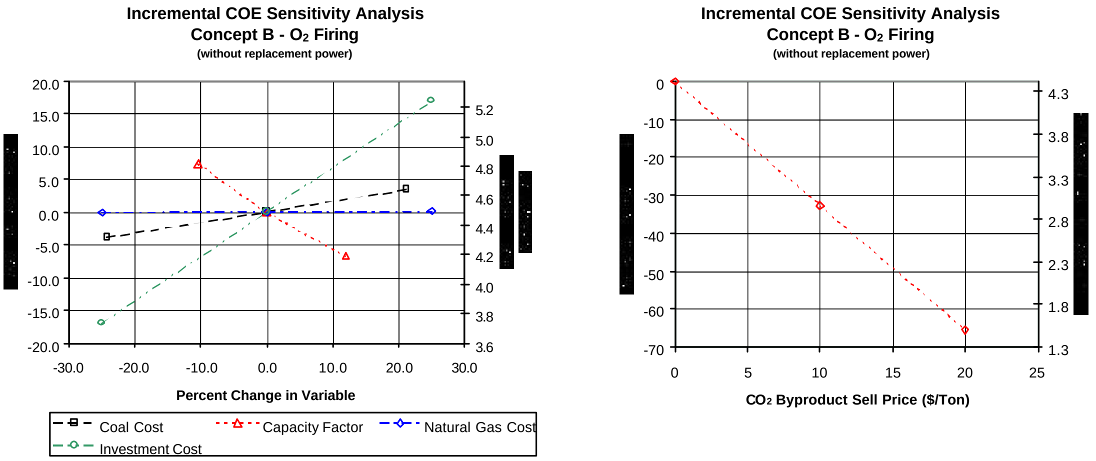
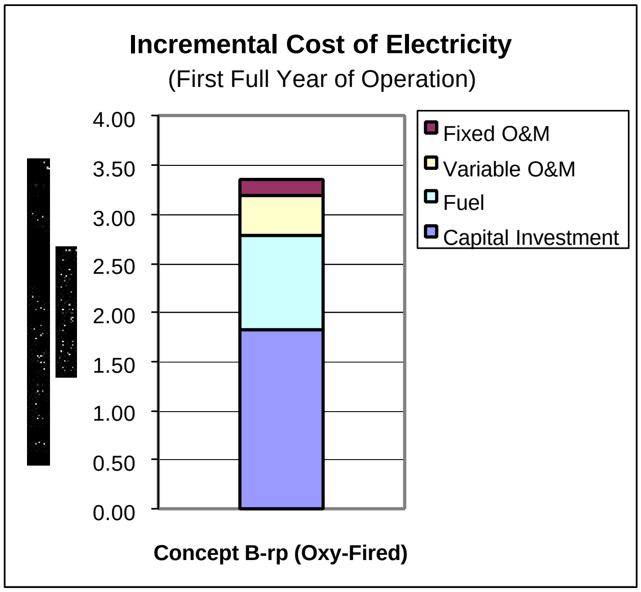


Figure 6.2. 2: Concept B (Oxygen Firing) Economic Sensitivity Results without Replacement Power

6.2.2 Concept B Results with Replacement Power

Figure 6.2.3 shows the economic analysis results using the base parameter values for Concept B (Oxygen Firing) with replacement power as compared with the Base Case. The first year incremental cost of electricity for this case was about 3.35 Cents/kWh greater than the Base Case COE.



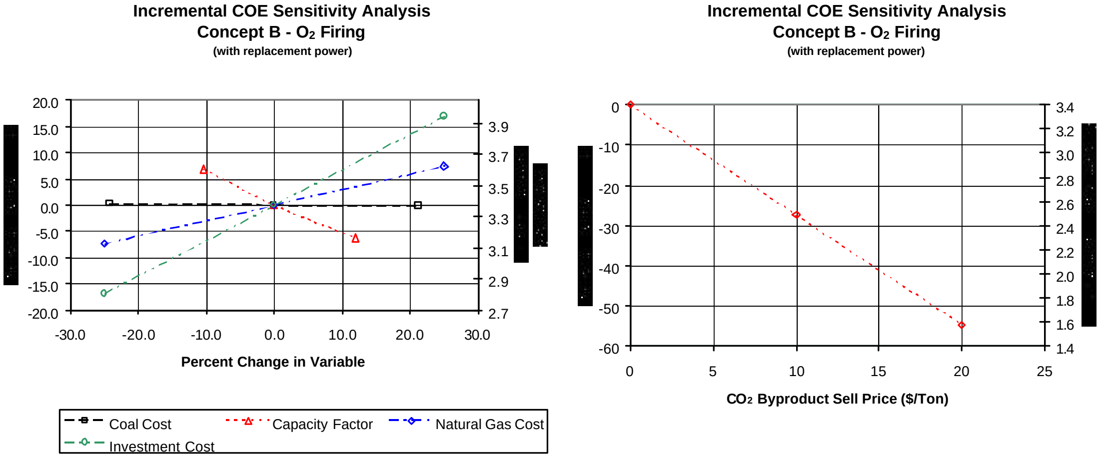
Incremental Cost of Electricity Distribution	(Cents/ kWh)
Capital Investment	1.83
Fixed O&M	0.15
Variable O&M	0.41
Fuel	0.97
Total Incremental Cost of Electricity	**3.35**

Figure 6.2. 3: Concept B (Oxygen Firing) Economic Results with Replacement Power

Results for the Concept B sensitivity study, with replacement power, are shown in Table 6.2.2 and Figure 6.2.4. Incremental COE ranges from a low of 1.52 to a high of 3.92 cents/kWh. The most sensitive parameters, again, are CO_2 byproduct sell price, investment cost and capacity factor, in that order. Natural gas cost shows significantly more impact on incremental COE when replacement power is included as would be expected with the NGCC being used to generate the replacement power. Coal cost, on the other hand, shows minimal impact on incremental COE with replacement power since both the Base Case and Case A use nearly the same amount of coal and the net electrical output is exactly the same. CO_2 mitigation costs ranged from 19-49 \$/Ton of CO_2 avoided with the baseline value at 42 \$/Ton of CO_2 avoided.

Table 6.2. 2: Economic Sensitivity Results for Concept B with Replacement Power

Parameter	Unit	CO_2 Capture with O2 Firing, Including Replacement Power										
Power Output												
Net Power Output	kW	273347	273347	273347	273347	273347	273347	273347	273347	273347	273347	273347
Replacement Power	kW	160430	160430	160430	160430	160430	160430	160430	160430	160430	160430	160430
Total Power Output	kW	433778	433778	433778	433778	433778	433778	433778	433778	433778	433778	433778
Plant Performance												
Net Plant heat Rate	Btu/kWh	12.026	12.026	12.026	12.026	12.026	12.026	12.026	12.026	12.026	12.026	12.026
Net Plant Efficiency	Fraction	0.2838	0.2838	0.2838	0.2838	0.2838	0.2838	0.2838	0.2838	0.2838	0.2838	0.2838
CO2 Emitted	lbm/h	175315	175315	175315	175315	175315	175315	175315	175315	175315	175315	175315
CO2 Emitted	lbm/kWh	0.404	0.404	0.404	0.404	0.404	0.404	0.404	0.404	0.404	0.404	0.404
Total Capital Cost	$ (1000's)	**443855**	443855	443855	443855	443855	443855	443855	443855	443855	**332891**	**554818**
Capital Cost	$/kW	1023	1023	1023	1023	1023	1023	1023	1023	1023	767	1279
Economic Analysis Basis												
Capital Charge Rate	%	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302
Fuel Cost (Coal)	$/MMBtu	**1.32**	**1.00**	**1.60**	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32
Fuel Cost (Natural Gas)	$/MMBtu	**4.00**	4.00	4.00	**3.00**	**5.00**	4.00	4.00	4.00	4.00	4.00	4.00
Capacity Factor	Fraction	**0.67**	0.67	0.67	0.67	0.67	**0.60**	**0.75**	0.67	0.67	0.67	0.67
CO2 Price	$/ton	**0.00**	0.00	0.00	0.00	0.00	0.00	0.00	**10.00**	**20.00**	0.00	0.00
Economic Analysis Results												
coe: Incremental	cents/kWh	3.35	3.36	3.35	3.10	3.60	3.58	3.14	2.44	1.52	2.79	3.92
CO2 Mitigation Cost	$/ton	42	42	42	39	45	45	39	31	19	35	49


Incremental COE Sensitivity Analysis
Concept B - O_2 Firing
(with replacement power)

Incremental COE Sensitivity Analysis
Concept B - O_2 Firing
(with replacement power)

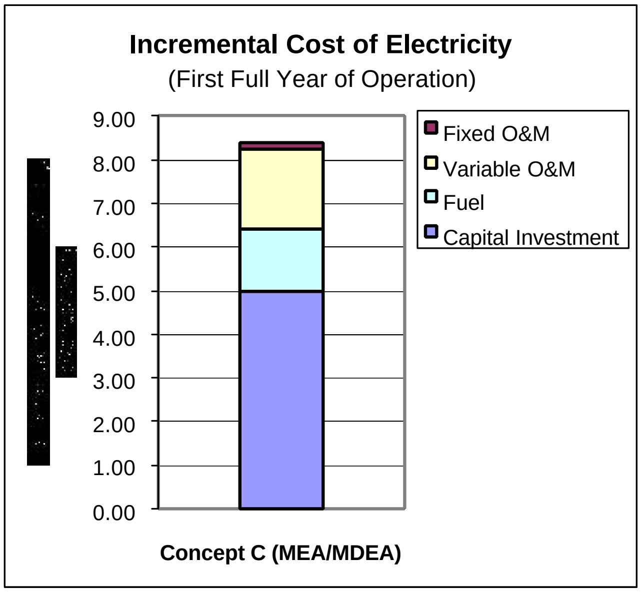
Figure 6.2. 4: Concept B (Oxygen Firing) Economic Sensitivity Results with Replacement Power

6.3 Concept C Results: CO_2 Separation with MEA / MDEA

This section shows the economic analysis results for Concept C, CO_2 Separation with MEA / MDEA absorption. Results both with and without replacement power are reported along with sensitivity results.

6.3.1 Concept C Results without Replacement Power

Figure 6.3.1 shows the economic results using the base parameter values for Concept C (MEA /MDEA) as compared with the Base Case (air fired without CO_2 capture). This analysis was done without replacement power. The first year incremental cost of electricity for this case was about 8.41 Cents/kWh greater than the Base Case COE.



Incremental Cost of Electricity Distribution	(Cents/ kWh)
Capital Investment	5.00
Fixed O&M	0.17
Variable O&M	1.81
Fuel	1.43
Total Incremental Cost of Electricity	**8.41**

Figure 6.3. 1: Concept C (MEA/MDEA) Economic Results without Replacement Power

Results for the Concept C sensitivity study, without replacement power, are shown in Table 6.3.1 and Figure 6.3.2. Incremental COE ranges from a low of 5.93 to a high of 9.77 cents/kWh. The most sensitive parameters are CO_2 byproduct sell price, investment cost and capacity factor with coal cost and natural gas cost showing significantly less impact on incremental COE.

CO_2 mitigation costs ranged from 69-113 $/Ton of CO_2 avoided with the baseline value at 98 $/Ton of CO_2 avoided.

Table 6.3. 1: Economic Sensitivity Results for Concept C without Replacement Power

Parameter	Unit	CO_2 Capture with MEA / MDEA, Without Replacement Power										
Power Output												
Net Power Output	kW	335973	335973	335973	335973	335973	335973	335973	335973	335973	335973	335973
Replacement Power	kW	0	0	0	0	0	0	0	0	0	0	0
Total Power Output	kW	335973	335973	335973	335973	335973	335973	335973	335973	335973	335973	335973
Plant Performance												
Net Plant heat Rate	Btu/kWh	15.223	15.223	15.223	15.223	15.223	15.223	15.223	15.223	15.223	15.223	15.223
Net Plant Efficiency	Fraction	0.2242	0.2242	0.2242	0.2242	0.2242	0.2242	0.2242	0.2242	0.2242	0.2242	0.2242
CO2 Emitted	lbm/h	92153	92153	92153	92153	92153	92153	92153	92153	92153	92153	92153
CO2 Emitted	lbm/kWh	0.274	0.274	0.274	0.274	0.274	0.274	0.274	0.274	0.274	0.274	0.274
Total Capital Cost	$ (1000's)	**824797**	824797	824797	824797	824797	824797	824797	824797	824797	**618598**	**1030996**
Capital Cost	$/kW	2455	2455	2455	2455	2455	2455	2455	2455	2455	1841	3069
Economic Analysis Basis												
Capital Charge Rate	%	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302
Fuel Cost (Coal)	$/MMBtu	**1.32**	**1.00**	**1.60**	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32
Fuel Cost (Natural Gas)	$/MMBtu	**4.00**	4.00	4.00	**3.00**	**5.00**	4.00	4.00	4.00	4.00	4.00	4.00
Capacity Factor	Fraction	**0.67**	0.67	0.67	0.67	0.67	**0.60**	**0.75**	0.67	0.67	0.67	0.67
CO2 Price	$/ton	**0.00**	0.00	0.00	0.00	0.00	0.00	0.00	**10.00**	**20.00**	0.00	0.00
Economic Analysis Results												
coe: Incremental	cents/kWh	8.41	8.32	8.49	8.15	8.67	9.01	7.86	7.17	5.93	7.05	9.77
CO2 Mitigation Cost	$/ton	98	97	99	95	101	105	91	83	69	82	113

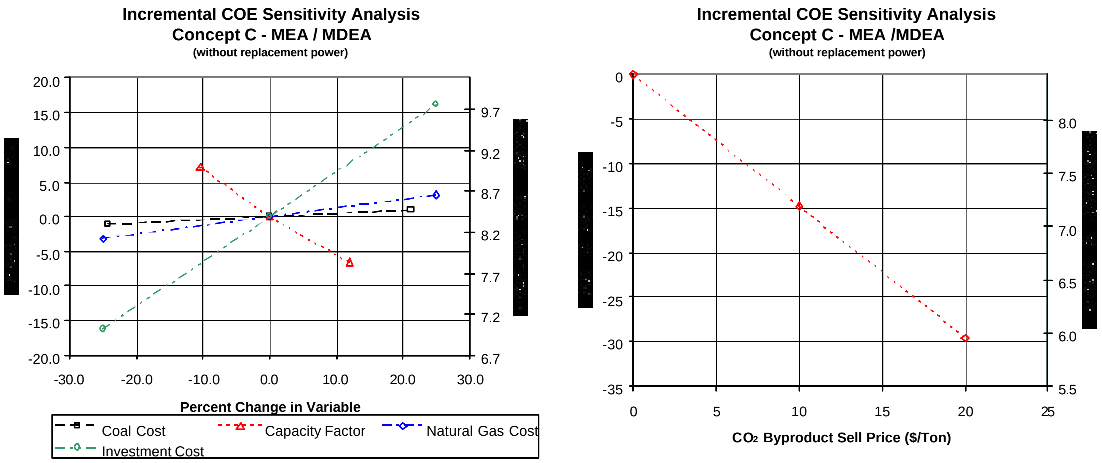
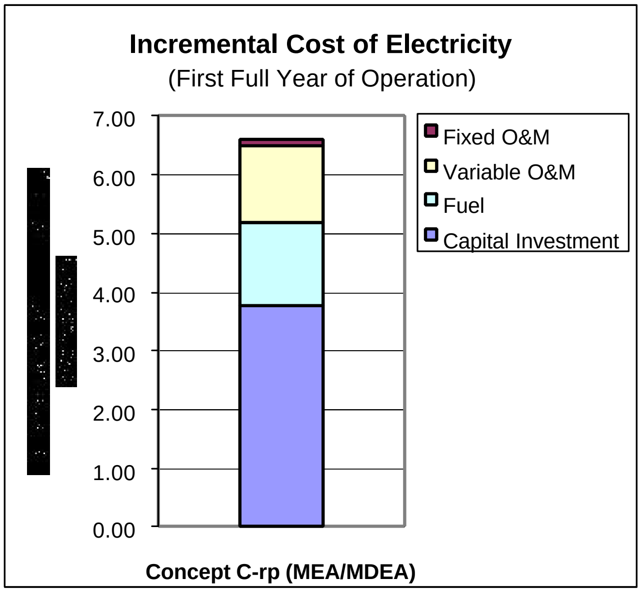


Figure 6.3. 2: Concept C (MEA/MDEA) Economic Sensitivity Results without Replacement Power

6.3.2 Concept C Results with Replacement Power

Figure 6.3.3 shows the economic analysis results using the base parameter values for Concept C (MEA / MDEA) with replacement power as compared with the Base Case. The first year incremental cost of electricity for this case was about 6.61 Cents/kWh greater than the Base Case COE.



Incremental Cost of Electricity Distribution	(Cents/ kWh)
Capital Investment	3.77
Fixed O&M	0.12
Variable O&M	1.30
Fuel	1.42
Total Incremental Cost of Electricity	**6.61**

Figure 6.3. 3: Concept C (MEA/MDEA) Economic Results with Replacement Power

Results for the Concept C sensitivity study with replacement power are shown in Table 6.3.2 and Figure 6.3.4. Incremental COE ranges from a low of 4.69 to a high of 7.66 cents/kWh. The most sensitive parameters again are CO_2 byproduct sell price, investment cost and capacity factor. Natural gas cost shows significantly more impact on incremental COE when replacement power is included as would be expected with the NGCC being used to generate the replacement power. Coal cost, on the other hand, shows minimal impact on incremental COE with replacement power since both the Base Case and Case A use the same amount of coal and the net electrical output is exactly the same.

CO_2 mitigation costs ranged from 58-95 \$/Ton of CO_2 avoided with the baseline value at 82 \$/Ton of CO_2 avoided.

Table 6.3. 2: Economic Sensitivity Results for Concept C with Replacement Power

Parameter	Unit	CO_2 Capture with MEA / MDEA, Without Replacement Power										
Power Output												
Net Power Output	kW	335973	335973	335973	335973	335973	335973	335973	335973	335973	335973	335973
Replacement Power	kW	97805	97805	97805	97805	97805	97805	97805	97805	97805	97805	97805
Total Power Output	kW	433778	433778	433778	433778	433778	433778	433778	433778	433778	433778	433778
Plant Performance												
Net Plant heat Rate	Btu/kWh	13.288	13.288	13.288	13.288	13.288	13.288	13.288	13.288	13.288	13.288	13.288
Net Plant Efficiency	Fraction	0.2568	0.2568	0.2568	0.2568	0.2568	0.2568	0.2568	0.2568	0.2568	0.2568	0.2568
CO2 Emitted	lbm/h	166711.546	166711.546	166711.546	166711.546	166711.546	166711.546	166711.546	166711.546	166711.546	166711.546	166711.546
CO2 Emitted	lbm/kWh	0.384	0.384	0.384	0.384	0.384	0.384	0.384	0.384	0.384	0.384	0.384
Total Capital Cost	$ (1000's)	**824797**	824797	824797	824797	824797	824797	824797	824797	824797	**618598**	**1030996**
Capital Cost	$/kW	1901	1901	1901	1901	1901	1901	1901	1901	1901	1426	2377
Economic Analysis Basis												
Capital Charge Rate	%	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302	0.1302
Fuel Cost (Coal)	$/MMBtu	**1.32**	**1.00**	**1.60**	1.32	1.32	1.32	1.32	1.32	1.32	1.32	1.32
Fuel Cost (Natural Gas)	$/MMBtu	**4.00**	4.00	4.00	**3.00**	**5.00**	4.00	4.00	4.00	4.00	4.00	4.00
Capacity Factor	Fraction	**0.67**	0.67	0.67	0.67	0.67	**0.60**	**0.75**	0.67	0.67	0.67	0.67
CO2 Price	$/ton	**0.00**	0.00	0.00	0.00	0.00	0.00	0.00	**10.00**	**20.00**	0.00	0.00
Economic Analysis Results												
coe: Incremental	cents/kWh	6.61	6.61	6.61	6.26	6.96	7.06	6.20	5.65	4.69	5.56	7.66
CO2 Mitigation Cost	$/ton	82	82	82	78	86	88	77	70	58	69	95

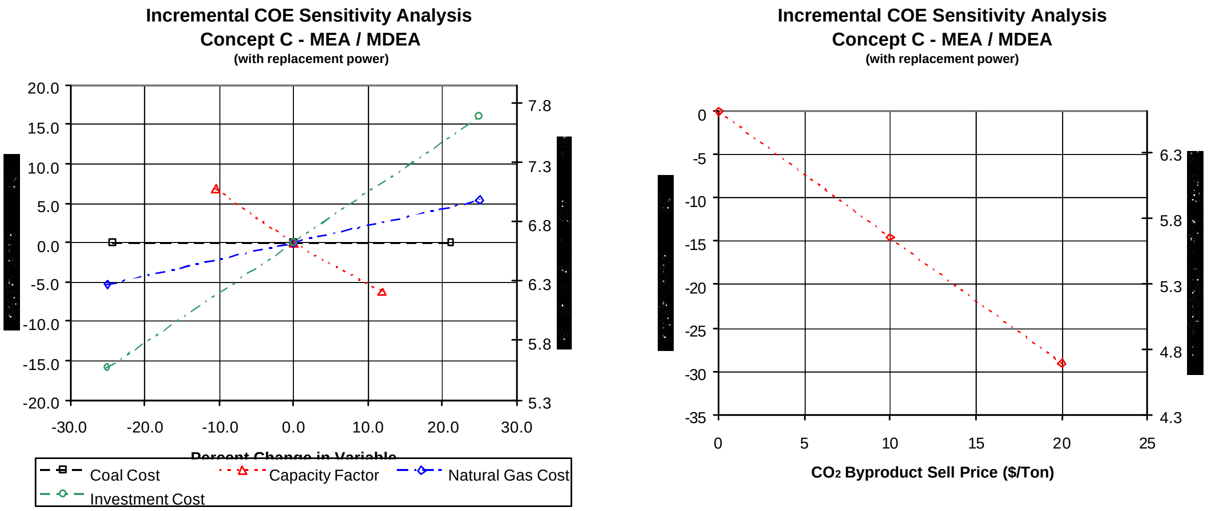


Figure 6.3. 4: Concept C (MEA/MDEA) Economic Sensitivity Results with Replacement Power

6.4 Economic Study Summary and Conclusions

A total of 66 CO_2 capture cases were compared in the reported sensitivity studies. All cases studied indicate significant increases to the COE as a result of CO_2 capture. The incremental COE as compared to the Base Case (air firing without CO_2 capture) ranges from 1.52 to 9.77 Cents/kWh. Similarly CO_2 mitigation costs range from about 17-113 \$/Ton of CO_2 avoided for the range of cases studied. Table 6.4.1 summarizes the economic analysis results for the six primary cases using base parameter values (i.e.; Coal Cost = 1.32 \$/10^6 Btu, Natural Gas Cost = 4.0 \$/10^6 Btu, Capacity Factor = 67%, CO_2 Byproduct Sell Price = 0.0 \$/Ton, Investment Cost = As Estimated). Figure 6.4.1 shows the incremental COE values for these cases and similarly, Figure 6.4.2 shows the associated CO_2 mitigation costs.

Table 6.4. 1: Summary of Economic Analysis Results (for 6 primary cases)

Parameter	Units	Concept A		Concept B		Concept C	
		w/o Repl Pwr	with Repl Pwr	w/o Repl Pwr	with Repl Pwr	w/o Repl Pwr	with Repl Pwr
Incremental COE	Cents/kWh	6.2	4.3	4.4	3.4	8.4	6.6
CO2 Mitigation Cost	$/Ton	68	53	49	42	98	82

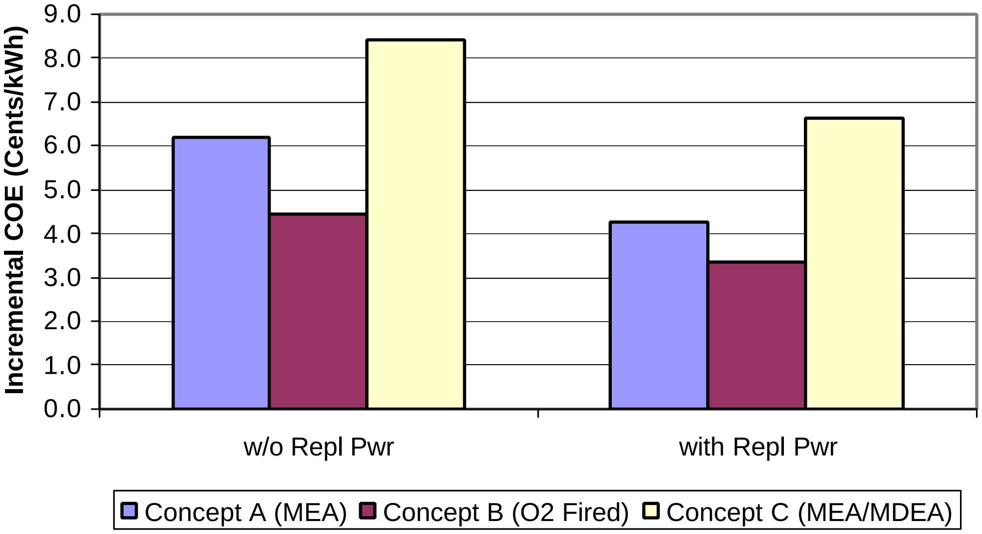


Figure 6.4. 1: Incremental Cost of Electricity (Using Base Parameter Values)

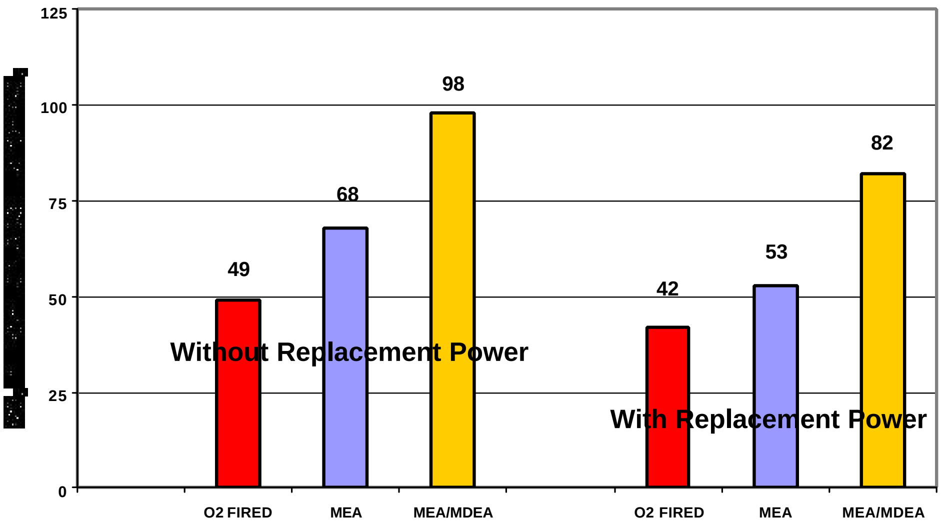


Figure 6.4. 2: Co$_2$ Mitigation Costs (Using Base Parameter Values)

Concept B (oxygen firing with flue gas recycle) appears to clearly be the best alternative of the three concepts studied based on incremental COE evaluation criteria. Using the base parameters of the sensitivity study for Concept B yields incremental COE values of 3.35 and 4.45 Cents/kWh for cases with and without replacement power, respectively. The actual COE values for Concept B are 20% and 13% lower than the corresponding values for Concept A and 47% and 51% lower than the corresponding values for Concept C for cases with and without replacement power respectively. If CO$_2$ byproduct can be sold at \$20/ton, incremental COE values are reduced to 1.53 and 1.52 Cents/kWh, respectively. Additional reductions would be possible with capacity factor increases, investment cost decreases or system efficiency increases. Significant investment cost decreases and efficiency increases may be possible as oxygen transport membrane technology develops. Previous internal studies (Liljedahl, et al., 1999) have shown membrane technology to provide specific investment cost (\$/kW) reductions of about 30% and net plant heat rate improvements greater than 20%.

Barring governmental mandates, it is clear that none of these three retrofit concepts will be acceptable to the electric utility industry from the standpoint of cost competitiveness.

7.0 COMPARISONS WITH PRIOR WORK

This section provides a comparison of the basic performance, CO_2 emissions, investment costs, and cost of electricity results of this study with selected results from the literature (David and Herzog, 2000; Palkes, et al., 1999; and Liljedahl, et al., 1999). Table 7.0.1 summarizes the pertinent technical results determined in this study (Marion, et al., 2001; Griffin, et al., 2001; and Liljedahl, et al., 2001).

Figures 7.0.1 and 7.0.2 compare net plant heat rates and CO_2 emissions for this study with selected results from the literature (David and Herzog, 2000; Palkes, et al., 1999; and Liljedahl, et al., 1999). This study shows a significantly greater impact on net plant heat rate, for the MEA process, than David and Herzog show. David and Herzog show about a 33% increase in net plant heat rate (e.g. 8,277 vs. 11,037 Btu/kWh; LHV basis) corresponding to an energy penalty of about 25%. This study shows about a 70% increase in net plant heat rate for the MEA system (e.g. 9,309 vs. 15,872 Btu/kWh; LHV basis) corresponding to an energy penalty of about 41%. A partial explanation for this difference can be seen in Figure 7.0.2, which shows specific CO_2 emissions. The present work shows significantly higher CO_2 removal (e.g., 0.906 to 0.059 kg/kWh) than David and Herzog show (e.g. 0.789 to 0.105 kg/kWh). Using the CO_2 emission and heat rate values from these figures, it can be determined that the CO_2 captured for the David and Herzog study is about 0.71 kg per original kWh, and for this study it is 0.87 kg per original kWh (e.g. this study is capturing about 23% more CO_2 per unit of original net electrical output). Additionally, this study produced a liquid CO_2 product stream at 2000 psig with product purity meeting the Dakota Gasification Project specification for EOR (Dakota Gasification WebPages). The David and Herzog product stream was reported at 1470 psia and the details regarding purity are not given although purity is probably similar in both cases. Both these differences would help explain at least partially, the efficiency penalty deviation between the two studies.

Table 7.0. 1: Summary of Performance for Original Plant and CO_2 Capture Study Cases

	(units)	**Original Plant (Base)**	**Concept A MEA**	**Concept B O2 Fired**	**Concept C MEA-MDEA**
Fuel Paramaters					
Coal Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4228.7	4140.0	4228.7
Natural Gas Heat Input (HHV)	(10^6 Btu/hr)	---	17.7	11.4	885.9
Total Fuel Heat Input (HHV)	(10^6 Btu/hr)	4228.7	4246.4	4151.5	5114.6
Steam Cycle Paramaters					
Existing Steam Turbine Generator Output	(kW)	463478	269341	463056	357196
CO2 Removal System Turbine Generator Output	(kW)	0	62081	0	36343
Total Turbine Generator Output	(kW)	463478	331422	463056	431290
Total Auxiliary Power	(kW)	29700	76007	189709	95317
Net Plant Output	(kW)	433778	255414	273347	335973
Overall Plant Performance Paramaters					
Net Plant Efficiency (HHV)	(fraction)	0.3501	0.2053	0.2247	0.2242
Net Plant Efficiency (LHV)	(fraction)	0.3666	0.2150	0.2354	0.2371
Normalized Efficiency (HHV; Relative to Base Case)	(fraction)	1.0000	0.5864	0.6419	0.6404
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9749	16626	15188	15223
Net Plant Heat Rate (LHV)	(Btu/kwhr)	9309	15872	14500	14395
Overall Plant CO_2 Emissions					
Carbon Dioxide Emissions	(lbm/hr)	866102	33084	53016	92153
Specific Carbon Dioxide Emissions	(lbm/kwhr)	1.997	0.130	0.194	0.274
Normalized Specific CO2 Emissions (Relative to Base Case)	(fraction)	1.000	0.065	0.097	0.137
Avoided Carbon Dioxide Emissions (as compared to Base)	(lbm/kwhr)	---	1.867	1.803	1.722
Specific Carbon Dioxide Emissions	(kg/kwhr)	0.906	0.059	0.088	0.125
Avoided Carbon Dioxide Emissions (as compared to Base)	(kg/kwhr)	---	0.848	0.818	0.782

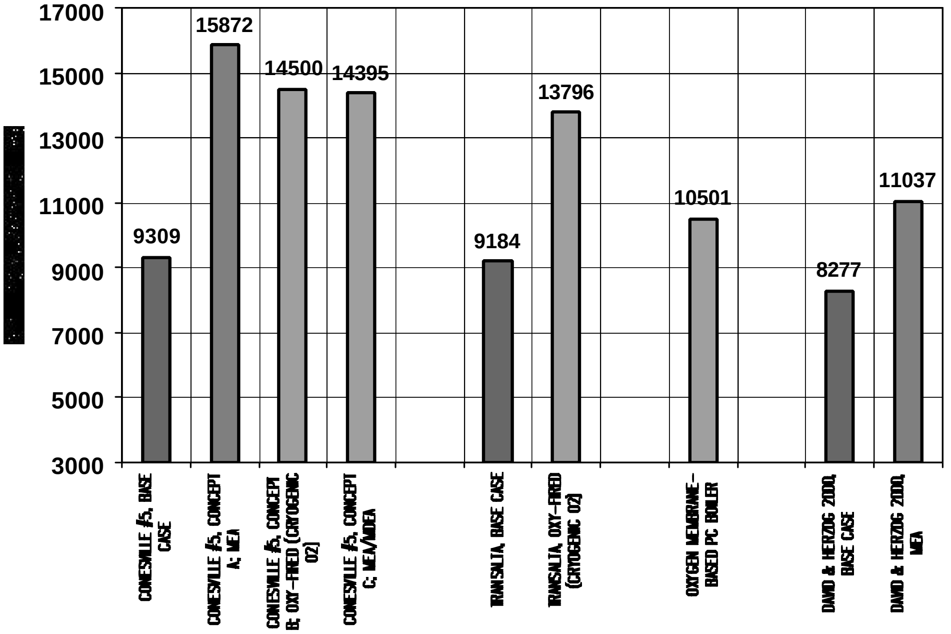


Figure 7.0. 1: Comparative Net Plant Heat Rate Results for Coal Fired Power Plants

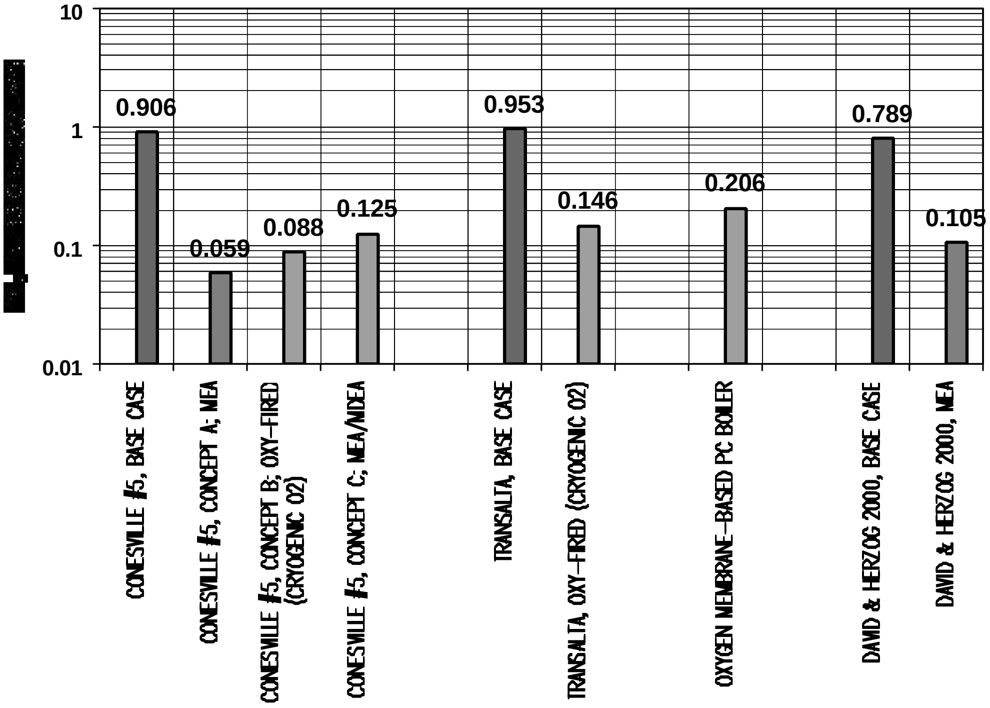


Figure 7.0. 2: Comparative CO_2 Emissions for Coal Fired Power Plants

Investment costs are also significantly different between the studies. David and Herzog (2000) show the cost of the CO_2 processing plant to be about 529 \$/kg/hr of CO_2. This study shows 1,037 \$/kg/hr of CO_2 for this equipment. The differences in the CO_2 product conditions and purity, discussed above, may represent part of this deviation. Additionally, for this retrofit study, the steam extracted from the existing steam turbine at the IP/LP crossover pipe had to be further expanded in a new letdown turbine before being exhausted to the reboilers for solvent regeneration. The David and Herzog study was for a new plant and therefore may not have required this letdown turbine/generator and the investment cost associated with it. Not enough information is given in their paper to determine if a letdown turbine is used in their particular system. Other savings may also possible for a new plant designed specifically for CO_2 removal. The steam turbine and generator may be significantly less expensive than for a reference plant designed for the same steam flow as a result of the large quantity of extracted low-pressure steam. The low-pressure feedwater heaters and condenser would also be significantly smaller and less expensive. David and Herzog show the incremental capital cost for a new plant, as compared to the reference plant without CO_2 capture, to be about 940 \$/kW as compared to this study, for a retrofit situation, which shows about 1,600 \$/kW.

Ultimately David and Herzog show the incremental cost of electricity to be about 3.3 Cents/kWh as compared to Concept A (MEA) of this study, which shows about 6.2 Cents/kWh. Financial assumptions, while probably not identical for both cases, would likely not cause a large impact on this difference.

With respect to oxy-fuel firing (Palkes, et al., 1999; Liljedahl, et al., 1999), it is seen that producing the oxygen in a ceramic membrane system leads to an improvement in net plant heat rate of more than 20% over the cases whereby the cryogenic method is used to produce oxygen (e.g., 10,501 vs. 13,796 Btu/kWh). CO_2 emissions are somewhat higher with the membrane system (e.g. 0.206 vs. 0.146 kg/kWh) due to natural gas firing in the membrane system air heater. Investment costs were shown to be more than 30% lower. Cost of electricity was not reported in this study.

8.0 CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE WORK

Conclusions:

No major technical barriers exist for retrofitting AEP's Conesville Unit #5 to capture CO_2 for any of the three concepts considered under this study. Nominally, 5-8 acres of new equipment space is needed and is located approximately 1500 feet north of the Unit #5 stack on the existing 200-acre power plant site which accommodates a total of 6 units. If all 6 units on this site were converted to CO_2 capture, it would be difficult if not impossible to accommodate all the new CO_2 capture equipment on the existing site. Energy requirements and power consumption are high, resulting in significant decrease in overall power plant efficiencies, ranging from about 20.5 to 23% as compared to 35% (all HHV basis) for the Base Case. Specific carbon dioxide emissions were reduced from about 2 lbm/kWh for the Base Case to 0.12 – 0.26 lbm/kWh for the study cases. Recovery of CO_2 ranged from 91 to 96%.

Specific investment costs are also high ranging from about 1000 to 2200 \$/kW without replacement power, and from about 800 to 1800 \$/kW with replacement power. All cases studied indicate significant increases to the COE as a result of CO_2 capture. The incremental COE as compared to the Base case (air firing without CO_2 capture) ranges from 3.4 to 8.4 ¢/kWh. Similarly CO_2 mitigation costs range from about 42-98 \$/ton of CO_2 avoided for the range of cases studied.

Concept B (oxygen firing with flue gas recycle) appears to clearly be the best alternative of the three concepts studied from both an efficiency and incremental COE viewpoint for systems designed for very high CO_2 capture (i.e. > 90%). If lower CO_2 capture fractions are considered, it appears that Concept A would likely be the best alternative with capture fractions below some as yet undetermined value. Concept C would also improve considerably with lower capture fractions. This study has confirmed two important issues related to firing coal in a CO_2 -rich flue gas /O_2 mixture:

- Modifications to the existing steam generator unit pressure parts are not required, and as such will also allow the unit to continue to operate in the conventional air fired mode. This is an important consideration as it is unlikely that much new coal-fired capacity can be built in the short term.
- CO_2-rich flue gas can be cleaned and compressed with a relatively simple system to provide high purity CO_2 for usage or sequestration.

While overall plant performance penalty and investment costs are high, an optimum integration of power generation, air separation, and CO_2 separation and compression systems may significantly improve the overall system efficiency and economics. The major cost item for Concept B is the air separation plant, both from a capital and an operating cost point of view. While in recent years advances have been made in air separation technology which have steadily improved the costs, new membrane based technologies are being researched which promise to provide greatly reduced auxiliary power requirements and investment costs which will significantly improve the overall system efficiency and economics. Previous studies (Liljedahl, et al., 1999) have shown membrane technology, applied to coal-fired boilers, to provide specific investment cost (\$/kW) reductions of about 30% and net plant heat rate improvements greater than 20% as compared to cryogenic based air separation.

Recommendations for Future Work:

- A sensitivity study, for Concepts A and C, showing the impact of reducing CO_2 capture percentage on plant efficiency, investment cost, emissions, and cost of electricity. The current work utilized 90% CO_2 capture (nominal). The sensitivity study would investigate nominal CO_2 capture percentages of 70% and 50%.
- Detailed analysis of the existing steam turbine for Concept A: In Concept A about 79% of the steam leaving the intermediate-pressure (IP) turbine is extracted from the IP/LP crossover pipe for solvent regeneration. The capability of the existing low-pressure (LP) turbine to operate under these conditions of very low steam flow should be investigated in detail, preferably by the original equipment manufacturer (OEM).

- Optimization of the amine system reboiler steam pressure for Concepts A and C.

- Concept B detailed boiler system analysis: A startup/shutdown procedure and system design, particularly the transition from air to oxygen firing (including transient conditions) should be developed. Detailed metal temperature analysis for all heat exchanger sections, including operation at part loads should be analyzed. This should also include furnace wall metal temperatures and analysis of the circulation system. The existing fans should be checked (preferably by the OEM's) for operation under the new conditions. The feasibility of operating the boiler under a slight positive pressure to eliminate air infiltration should also be investigated.

- Investigation of Improved oxygen production systems for Concept B, in line with the membrane-based air separation research being conducted by various research groups (e.g., Praxair, Air Products, Norsk Hydro). Also optimization of an integrated boiler and oxygen production system.

- Measurement of furnace heat transfer in CO_2/O_2 environments in a proof of concept boiler simulation facility.

- Improved solvents, which require lower regeneration energy requirements and/or can be regenerated at a lower temperature level.

- Hybrid process using oxygen-enriched combustion and amine based CO_2 absorption, to accrue, simultaneously, both CO_2 capture and drastic NOx emissions reduction.

- Investigation of a new novel high risk CO_2 capture process that would reduce efficiency penalty and retrofit investment cost. This would likely not utilize the existing boiler.

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APPENDIX I - DRAWINGS

This appendix contains all drawings developed for this project for Concepts A, B, and C and also includes a plot plan of the existing site.

Existing Plant:
66-530.00 Plot Plan – Existing Overall Site before CO_2 Unit Addition

Concept A:
U01-D-0208 Plot Plan – Concept A: Flue Gas Cooling & CO_2 Absorption
U01-D-0214 Plot Plan – Concept A: Solvent Stripping
U01-D-0204 Plot Plan – Concept A: CO_2 Compression & Liquefaction
U01-D-0211 Plot Plan – Concept A: Overall Layout Conceptual Plan
U01-D-0200R Plot Plan – Concept A: Modified Overall Site Plan

Concept B:
U01-D-0203 Plot Plan – Concept B: Air Separation Plants
U01-D-0209 Plot Plan – Concept B: Flue Gas Cooling
U01-D-0205 Plot Plan – Concept B: CO_2 Compression & Liquefaction
U01-D-0212 Plot Plan – Concept B: Overall Layout Conceptual Plan
U01-D-0201R Plot Plan – Concept B: Modified Overall Site Plan

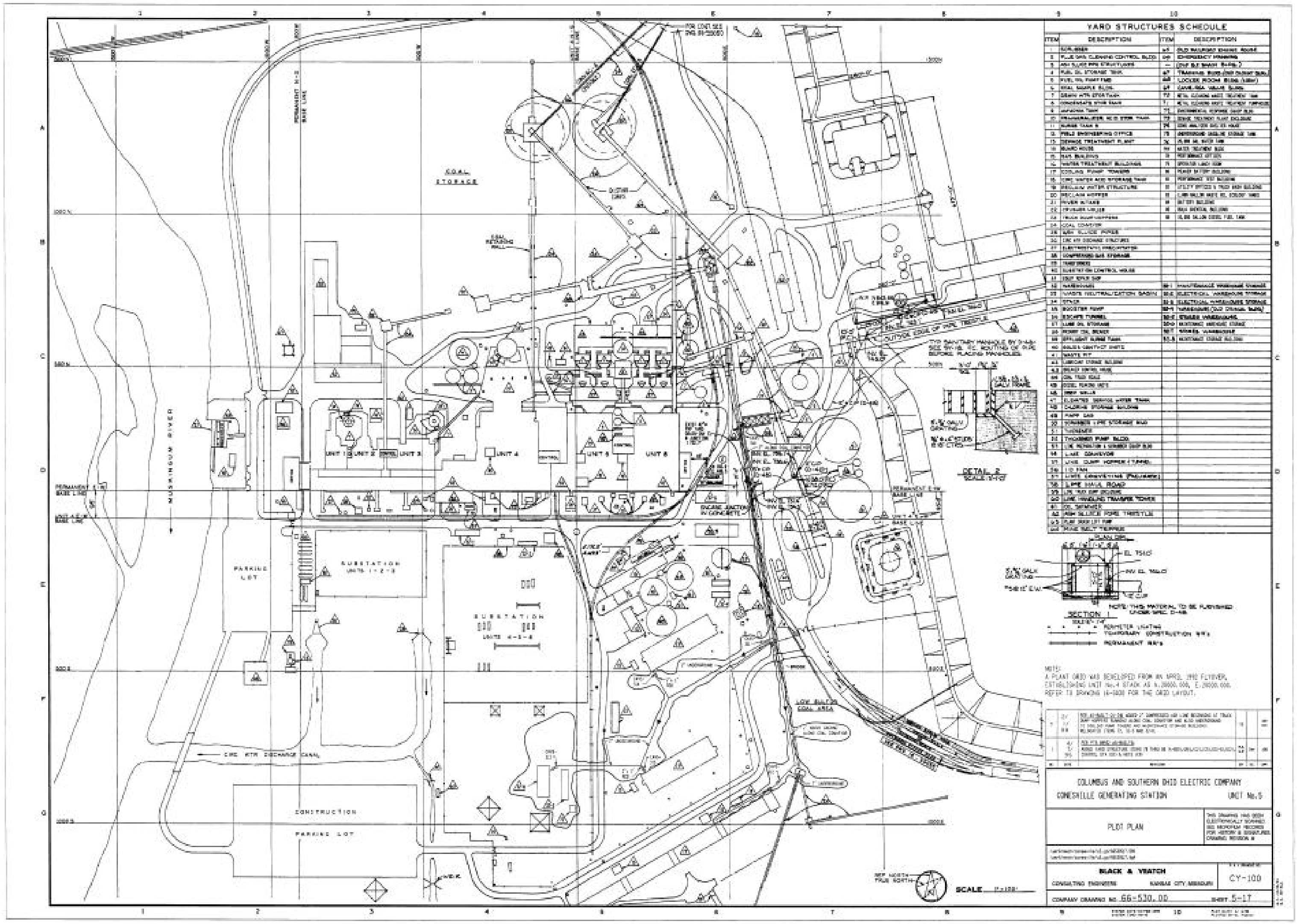
Boiler Modifications:
U00-E-0551R General Arrangement - Side Elevation for Concept B
U00-E-0552R General Arrangement - Plan View "B-B" for Concept B
U00-E-0585R General Arrangement - Plan View "A-A" for Concept B

Concept C:
U01-D-0207 Plot Plan – Concept C: De-Oxy Catalyst & Cooling
U01-D-0210 Plot Plan – Concept C: Flue Gas Cooling & CO_2 Absorption
U01-D-0215 Plot Plan – Concept C: Solvent Stripping
U01-D-0206 Plot Plan – Concept C: CO_2 Compression & Liquefaction
U01-D-0213 Plot Plan – Concept C: Overall Layout Conceptual Plan
U01-D-0202R Plot Plan – Concept C: Modified Overall Site Plan

Existing Plant:

The existing plant site drawing is shown below:

66-530.00 Plot Plan – Existing Overall Site before CO_2 Unit Addition


YARD STRUCTURES SCHEDULE
ITEM
DESCRIPTION
COAL STORAGE
MUSKINGUM RIVER
UNIT 1
UNIT 2
UNIT 3
UNIT 4
UNIT 5
UNIT 6
PARKING LOT
SUBSTATION UNITS 1-2-3
SUBSTATION UNITS 4-5-6
CIRC WTR DISCHARGE CANAL
CONSTRUCTION PARKING LOT
LOW SULFUR COAL AREA
DETAIL 2
SECTION 1
EL 754.0
PERIMETER LIGHTING
TEMPORARY CONSTRUCTION RR's
PERMANENT RR's
NOTE:
A PLANT GRID WAS DEVELOPED FROM AN APRIL 1992 FLYOVER, ESTABLISHING UNIT No.4 STACK AS N.20000.000, E.20000.000. REFER TO DRAWING 16-1000 FOR THE GRID LAYOUT.
COLUMBUS AND SOUTHERN OHIO ELECTRIC COMPANY
CONESVILLE GENERATING STATION
UNIT No.5
PLOT PLAN
BLACK & VEATCH
CONSULTING ENGINEERS
KANSAS CITY, MISSOURI
CY-100
COMPANY DRAWING NO. 66-530.00
SHEET 5-17
SCALE
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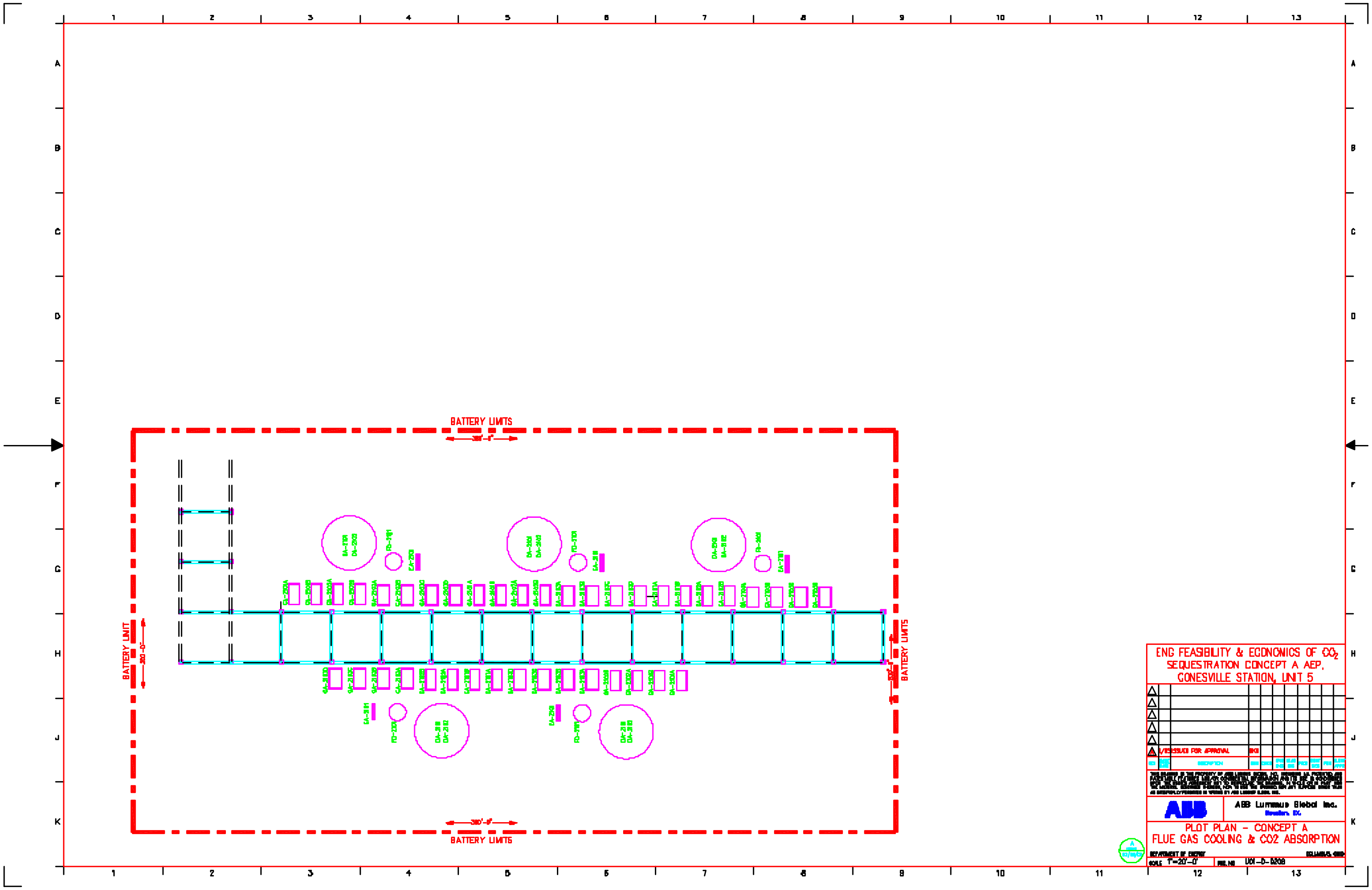
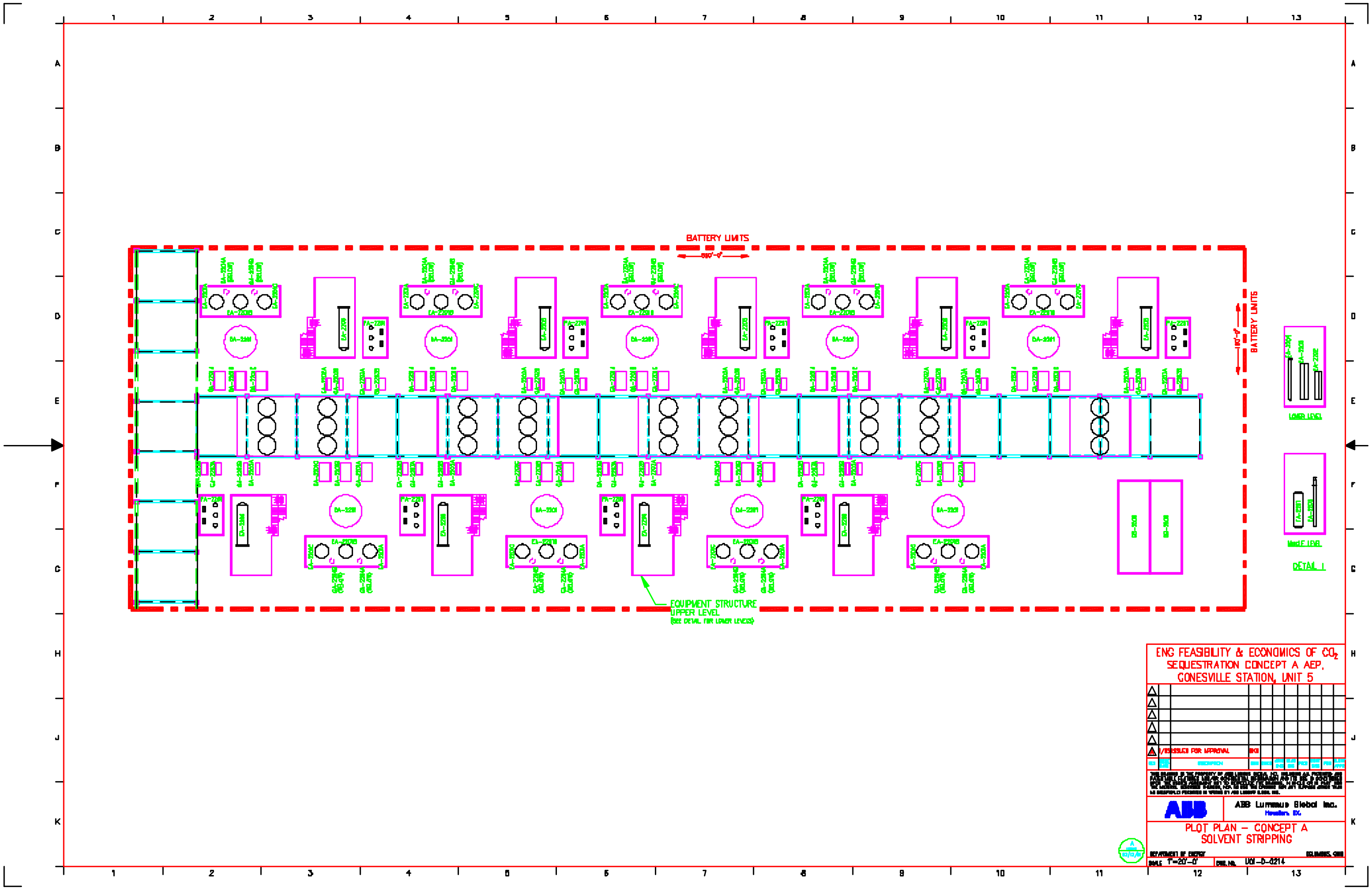
A
B
C
D
E
F
G

Concept A:

The plant layout drawings prepared for the Concept A CO_2 Recovery System are as follows:

U01-D-0208 Plot Plan – Concept A: Flue Gas Cooling & CO_2 Absorption
U01-D-0214 Plot Plan – Concept A: Solvent Stripping
U01-D-0204 Plot Plan – Concept A: CO_2 Compression & Liquefaction
U01-D-0211 Plot Plan – Concept A: Overall Layout Conceptual Plan
U01-D-0200 Plot Plan – Concept A: Modified Overall Site Plan


BATTERY LIMITS
BATTERY LIMIT
BATTERY LIMITS
BATTERY LIMITS
ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT A AEP,
CONESVILLE STATION, UNIT 5
ISSUED FOR APPROVAL
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN - CONCEPT A
FLUE GAS COOLING & CO2 ABSORPTION
DEPARTMENT OF ENERGY
COLUMBUS, OHIO
SCALE 1"=20'-0"
DWG. NO. UOI-D-D208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550'-0"
BATTERY LIMITS
180'-0"
EA-2201B
EA-2204A (BELOW)
EA-2204B (BELOW)
EQUIPMENT STRUCTURE
UPPER LEVEL
(SEE DETAIL FOR LOWER LEVELS)
LOWER LEVEL
MIDDLE LEVEL
DETAIL 1
ENG FEASIBILITY & ECONOMICS OF CO_2
SEQUESTRATION CONCEPT A AEP,
CONESVILLE STATION, UNIT 5
ISSUED FOR APPROVAL
REV
DESCRIPTION
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN – CONCEPT A
SOLVENT STRIPPING
DEPARTMENT OF ENERGY
COLUMBUS, OHIO
SCALE 1"=20'-0"
DWG. NO. U01-D-0214
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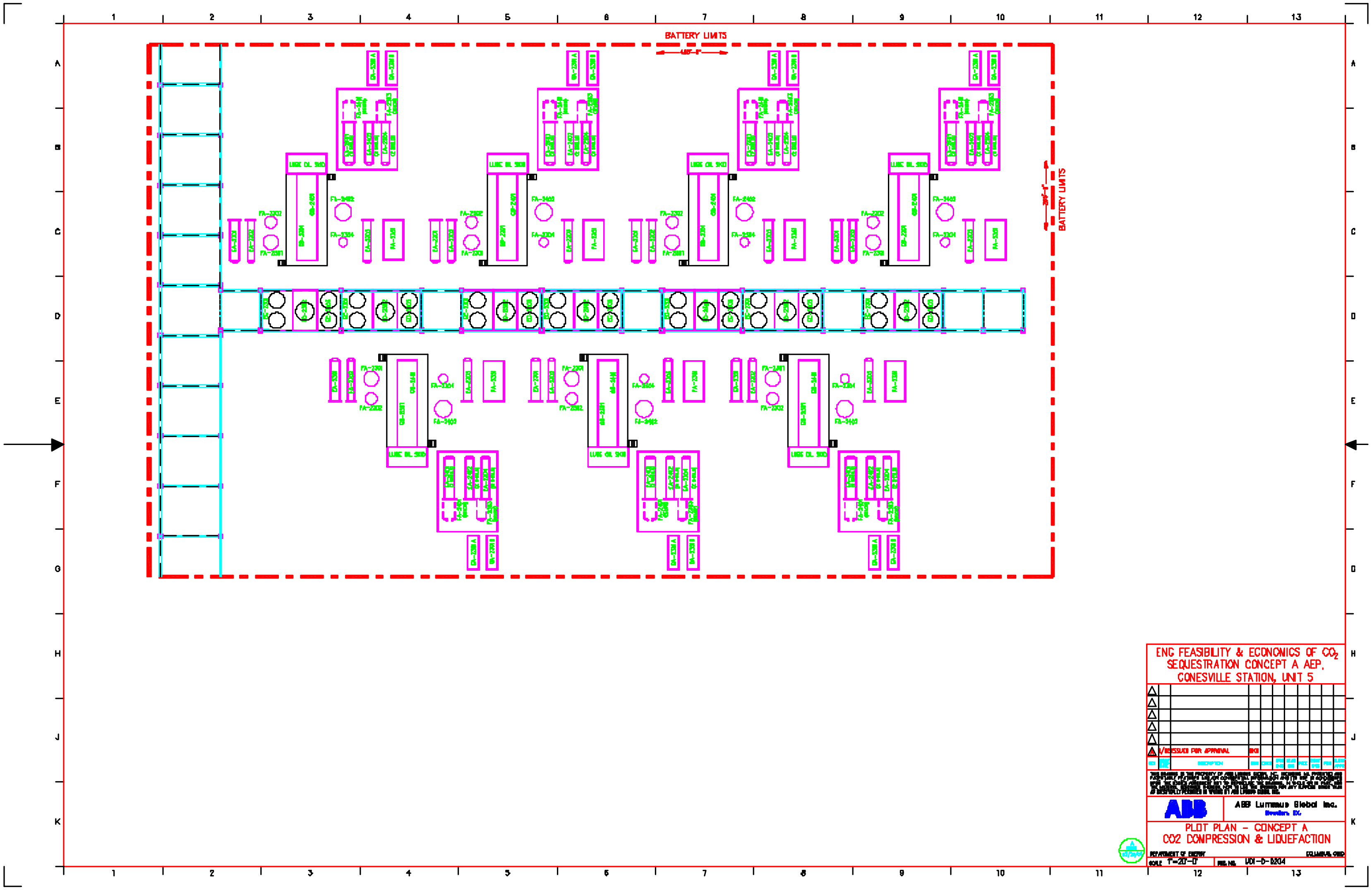
K


BATTERY LIMITS
BATTERY LIMITS
ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT A AEP,
CONESVILLE STATION, UNIT 5
A 1/05 ISSUED FOR APPROVAL
REV
DESCRIPTION
THIS DRAWING IS THE PROPERTY OF ABB LUMMUS GLOBAL INC. INCLUDING ALL PATENTED AND PATENTABLE FEATURES AND/OR CONFIDENTIAL INFORMATION AND ITS USE IS CONDITIONED UPON THE USER'S AGREEMENT NOT TO REPRODUCE THE DRAWING, IN WHOLE OR IN PART, NOR THE MATERIAL DESCRIBED THEREON, NOR TO USE THE DRAWING FOR ANY PURPOSE OTHER THAN AS SPECIFICALLY PERMITTED IN WRITING BY ABB LUMMUS GLOBAL INC.
ABB
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN - CONCEPT A
CO2 COMPRESSION & LIQUEFACTION
DEPARTMENT OF ENERGY
COLUMBUS, OHIO
SCALE 1"=20'-0"
DWG. NO. UDI-D-0204
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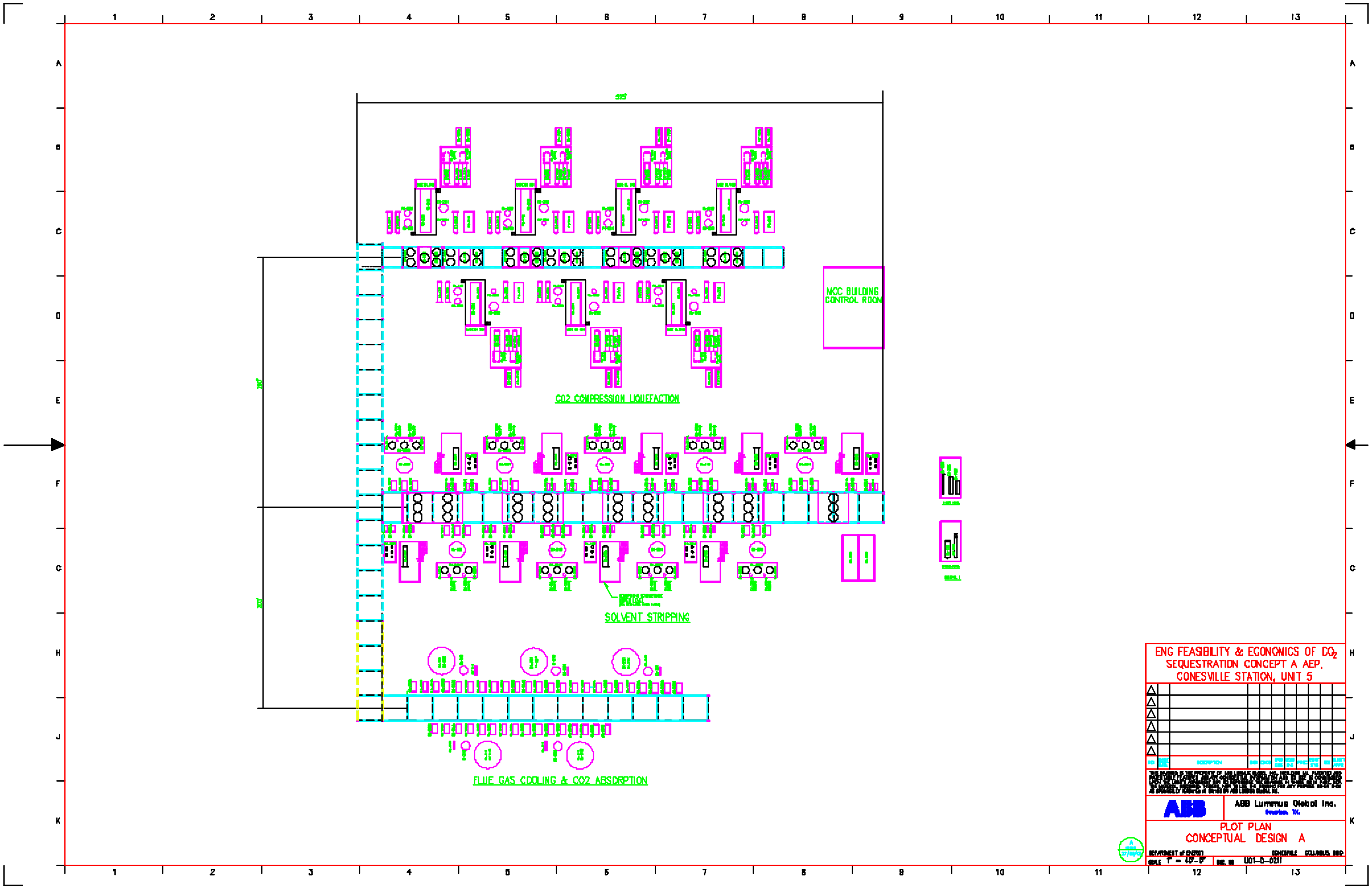
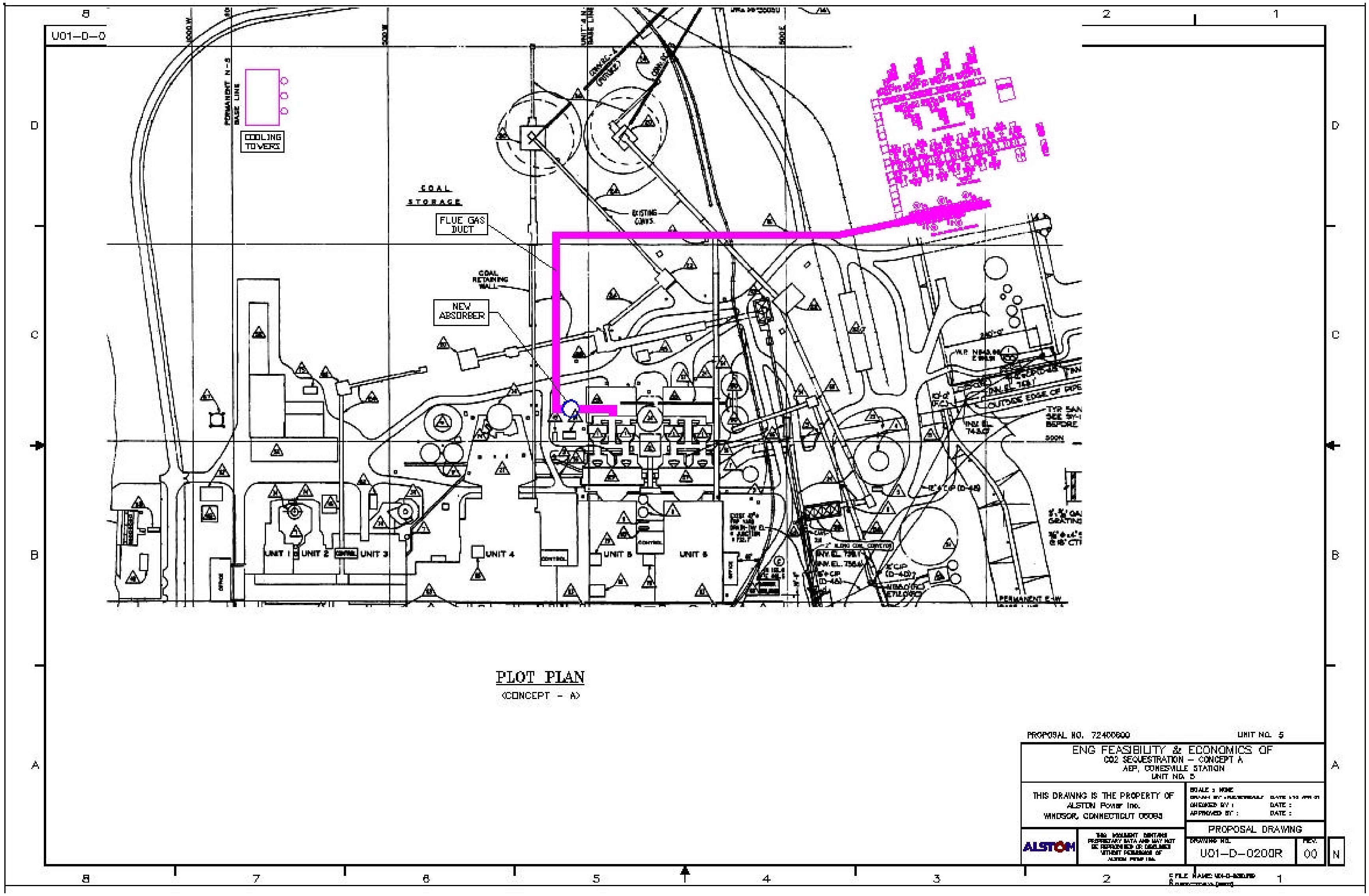

CO2 COMPRESSION LIQUEFACTION
MCC BUILDING
CONTROL ROOM
SOLVENT STRIPPING
FLUE GAS COOLING & CO2 ABSORPTION
ENG FEASIBILITY & ECONOMICS OF CO_2
SEQUESTRATION CONCEPT A AEP,
CONESVILLE STATION, UNIT 5
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN
CONCEPTUAL DESIGN A
DEPARTMENT of ENERGY
CONESVILLE COLUMBUS, OHIO
SCALE 1" = 40'-0"
DWG. NO. U01-D-0211


U01-D-0
COOLING TOWERS
PERMANENT N-S BASE LINE
COAL STORAGE
FLUE GAS DUCT
COAL RETAINING WALL
NEW ABSORBER
EXISTING CONVS.
UNIT 1
UNIT 2
UNIT 3
UNIT 4
UNIT 5
UNIT 6
CONTROL
OFFICE
OUTSIDE EDGE OF PIPE
PERMANENT E-W
PLOT PLAN
(CONCEPT - A)
PROPOSAL NO. 72400600
UNIT NO. 5
ENG FEASIBILITY & ECONOMICS OF
CO2 SEQUESTRATION - CONCEPT A
AEP, CONESVILLE STATION
UNIT NO. 5
THIS DRAWING IS THE PROPERTY OF
ALSTOM Power Inc.
WINDSOR, CONNECTICUT 06095
SCALE : NONE
CHECKED BY :
DATE :
APPROVED BY :
DATE :
PROPOSAL DRAWING
DRAWING NO.
U01-D-0200R
REV.
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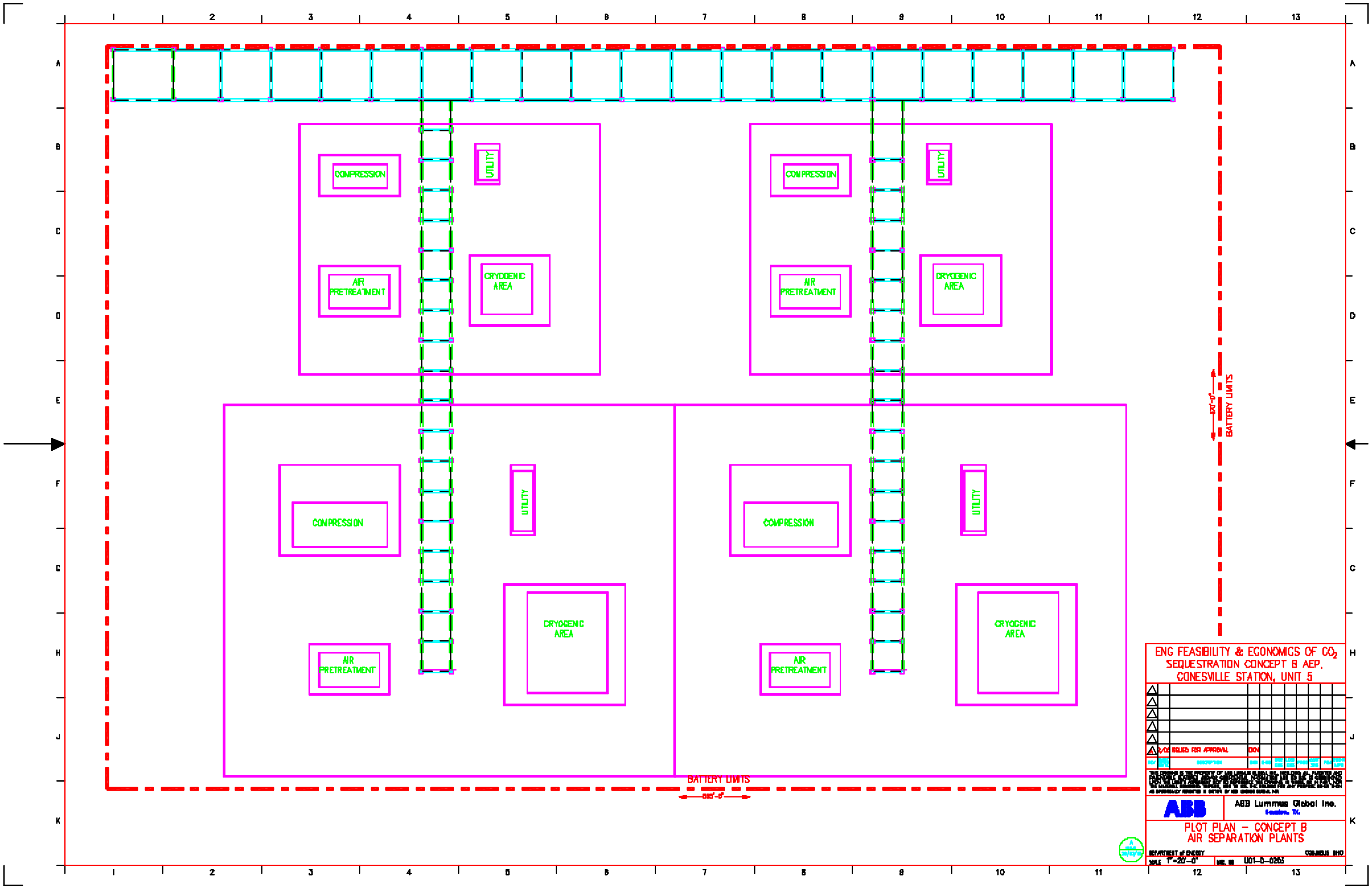
Concept B:

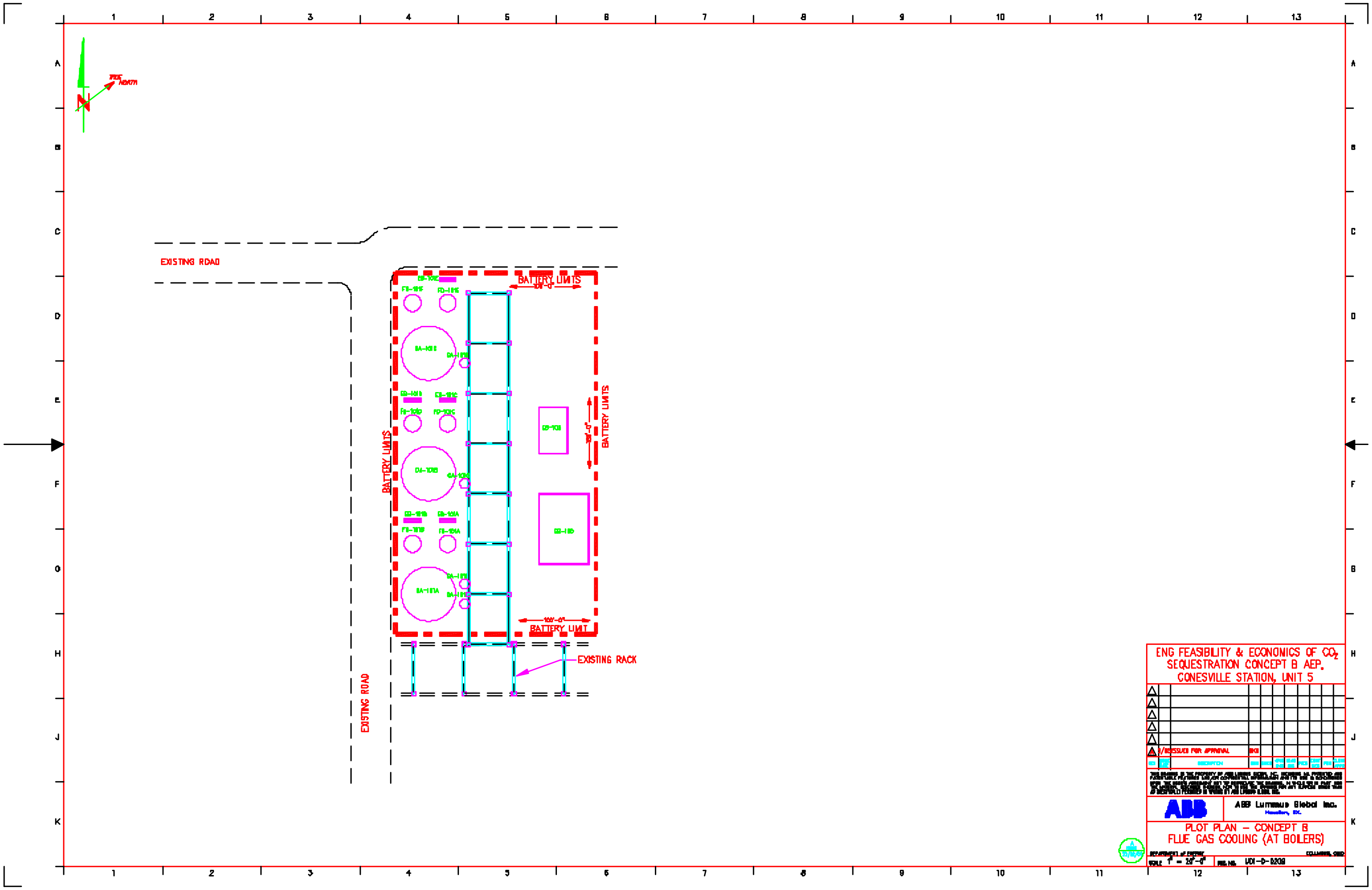
The plant layout drawings prepared for the Concept B Air Separation Unit & CO_2 Recovery System are as follows:

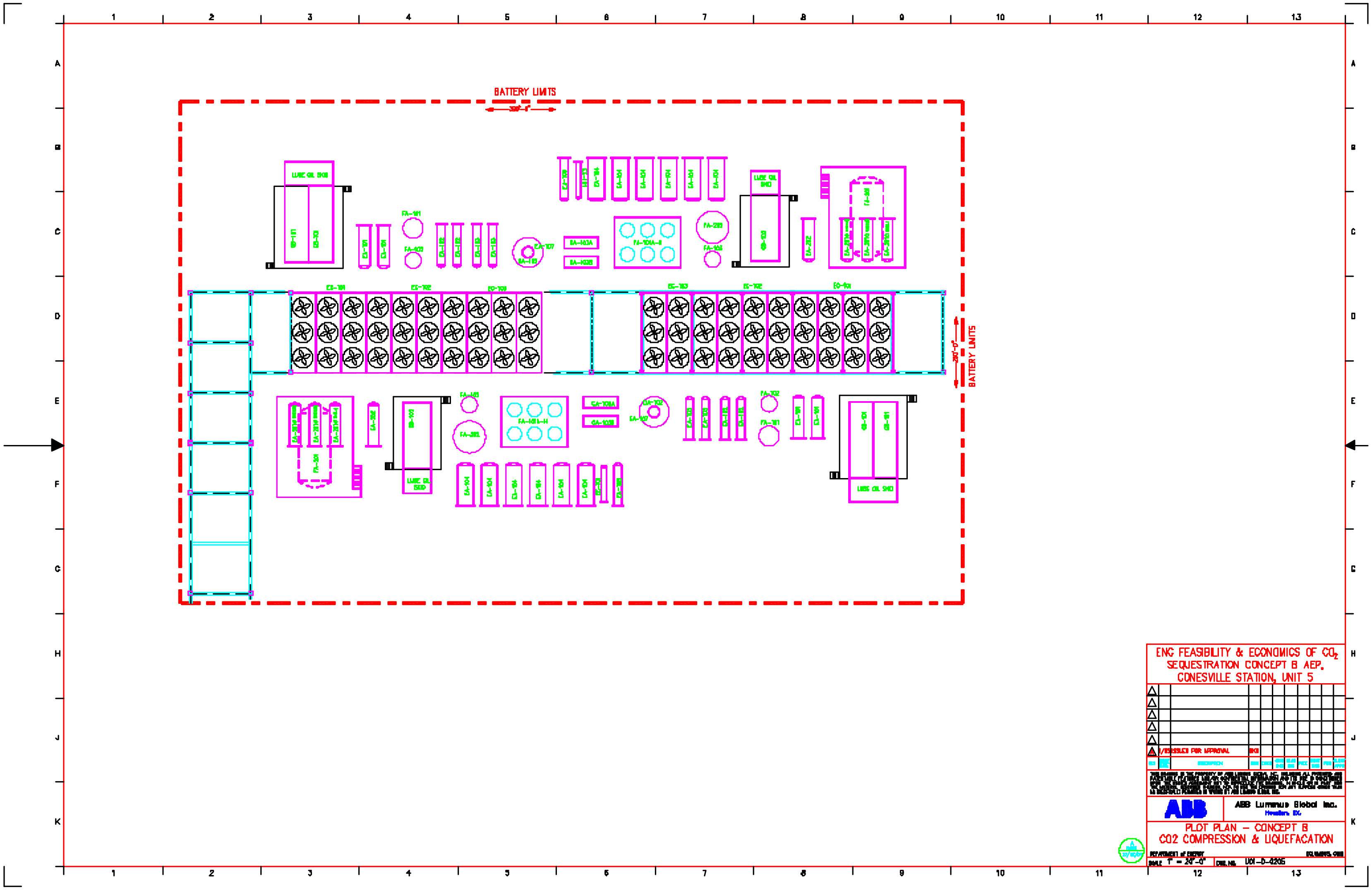
U01-D-0203 Plot Plan – Concept B: Air Separation Plants
U01-D-0209 Plot Plan – Concept B: Flue Gas Cooling
U01-D-0205 Plot Plan – Concept B: CO_2 Compression & Liquefaction
U01-D-0212 Plot Plan – Concept B: Overall Layout Conceptual Plan
U01-D-0201 Plot Plan – Concept B: Modified Overall Site Plan

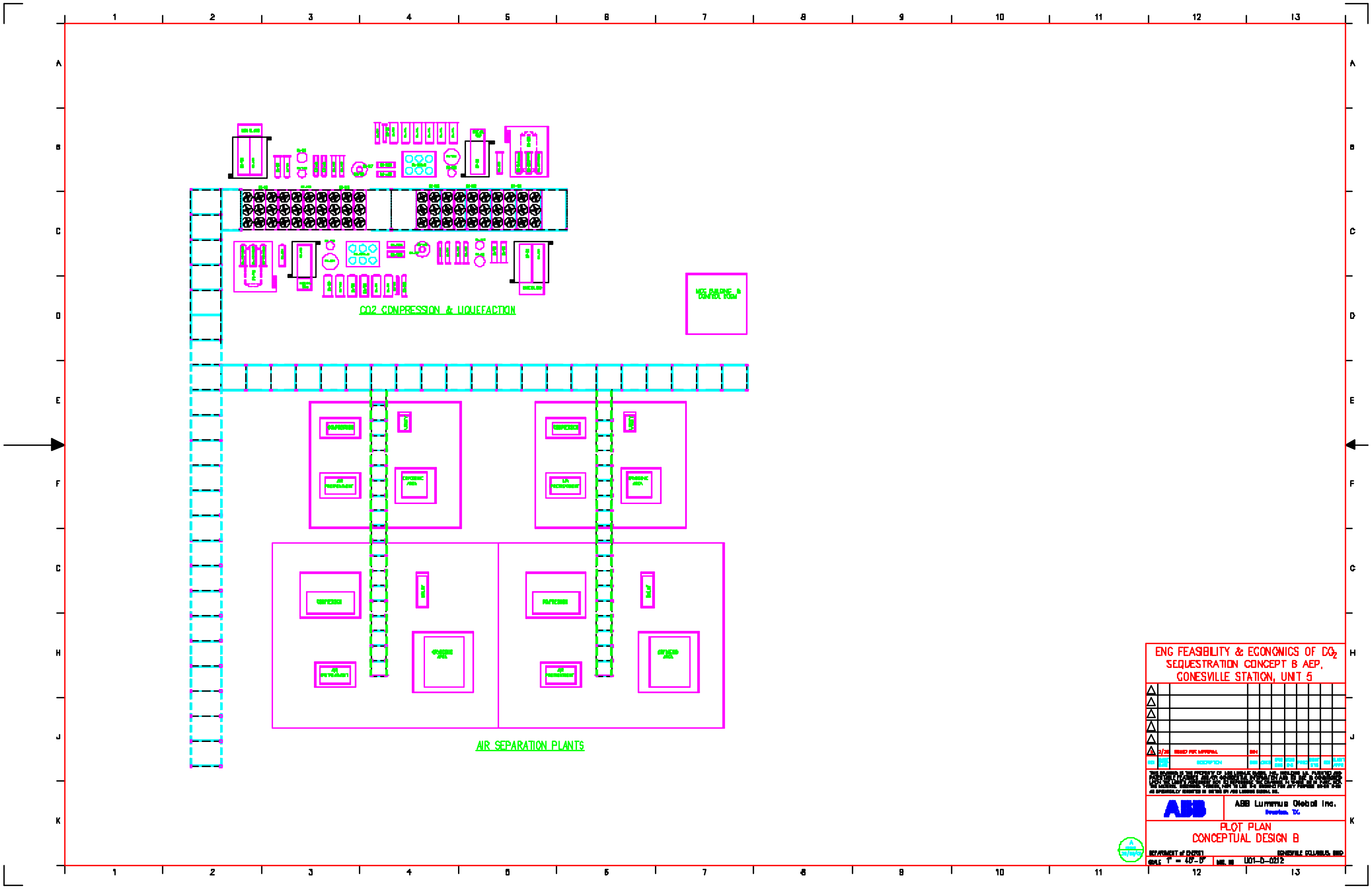
The boiler modification drawings prepared for the Concept B are as follows:

U00-E-0551- General Arrangement - Side Elevation for "Concept B"
U00-E-0552 - General Arrangement - Plan View "B-B" for "Concept B"
U00-E-0585 - General Arrangement - Plan View "A-A" for "Concept


COMPRESSION
UTILITY
AIR PRETREATMENT
CRYOGENIC AREA
COMPRESSION
UTILITY
AIR PRETREATMENT
CRYOGENIC AREA
COMPRESSION
UTILITY
CRYOGENIC AREA
AIR PRETREATMENT
COMPRESSION
UTILITY
CRYOGENIC AREA
AIR PRETREATMENT
BATTERY LIMITS
575'-0"
BATTERY LIMITS
585'-0"
ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT B AEP,
CONESVILLE STATION, UNIT 5
A 2/02 ISSUED FOR APPROVAL DKN
REV
DESCRIPTION
ABB
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN – CONCEPT B
AIR SEPARATION PLANTS
DEPARTMENT of ENERGY
COLUMBUS OHIO
SCALE 1"=20'-0"
DWG. NO U01-D-0203
1 2 3 4 5 6 7 8 9 10 11 12 13
A B C D E F G H J K


TRUE NORTH
EXISTING ROAD
EXISTING ROAD
BATTERY LIMITS
BATTERY LIMITS
BATTERY LIMITS
100'-0"
100'-0"
BATTERY LIMIT
EXISTING RACK
ENG FEASIBILITY & ECONOMICS OF CO2 SEQUESTRATION CONCEPT B AEP. CONESVILLE STATION, UNIT 5
ISSUED FOR APPROVAL
REV
DESCRIPTION
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN – CONCEPT B
FLUE GAS COOLING (AT BOILERS)
DEPARTMENT of ENERGY
COLUMBUS, OHIO
SCALE 1" = 20'-0"
DWG. NO. UDI-D-0209


BATTERY LIMITS
BATTERY LIMITS
ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT B AEP.
CONESVILLE STATION, UNIT 5
A /05 ISSUED FOR APPROVAL
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN – CONCEPT B
CO2 COMPRESSION & LIQUEFACATION
DEPARTMENT of ENERGY
COLUMBUS, OHIO
SCALE 1" = 24'-0"
DWG. NO. U01-D-0205


CO2 COMPRESSION & LIQUEFACTION
MCC BUILDING & CONTROL ROOM
AIR SEPARATION PLANTS
ENG FEASIBILITY & ECONOMICS OF CO2 SEQUESTRATION CONCEPT B AEP, CONESVILLE STATION, UNIT 5
ISSUED FOR MATERIAL
DESCRIPTION
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN
CONCEPTUAL DESIGN B
DEPARTMENT of ENERGY
CONESVILLE COLUMBUS, OHIO
SCALE 1" = 40'-0"
DWG. NO U01-D-0212
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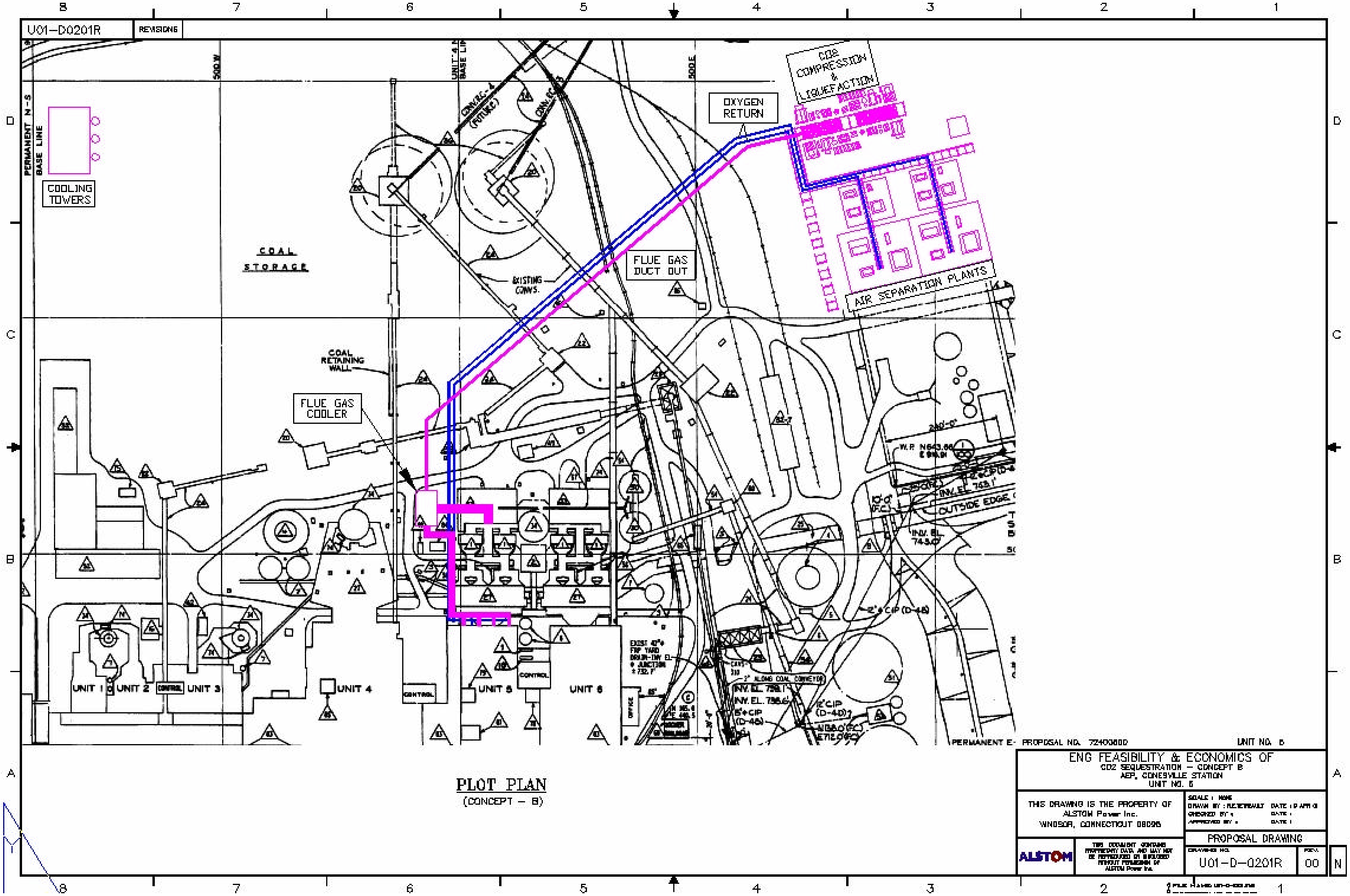
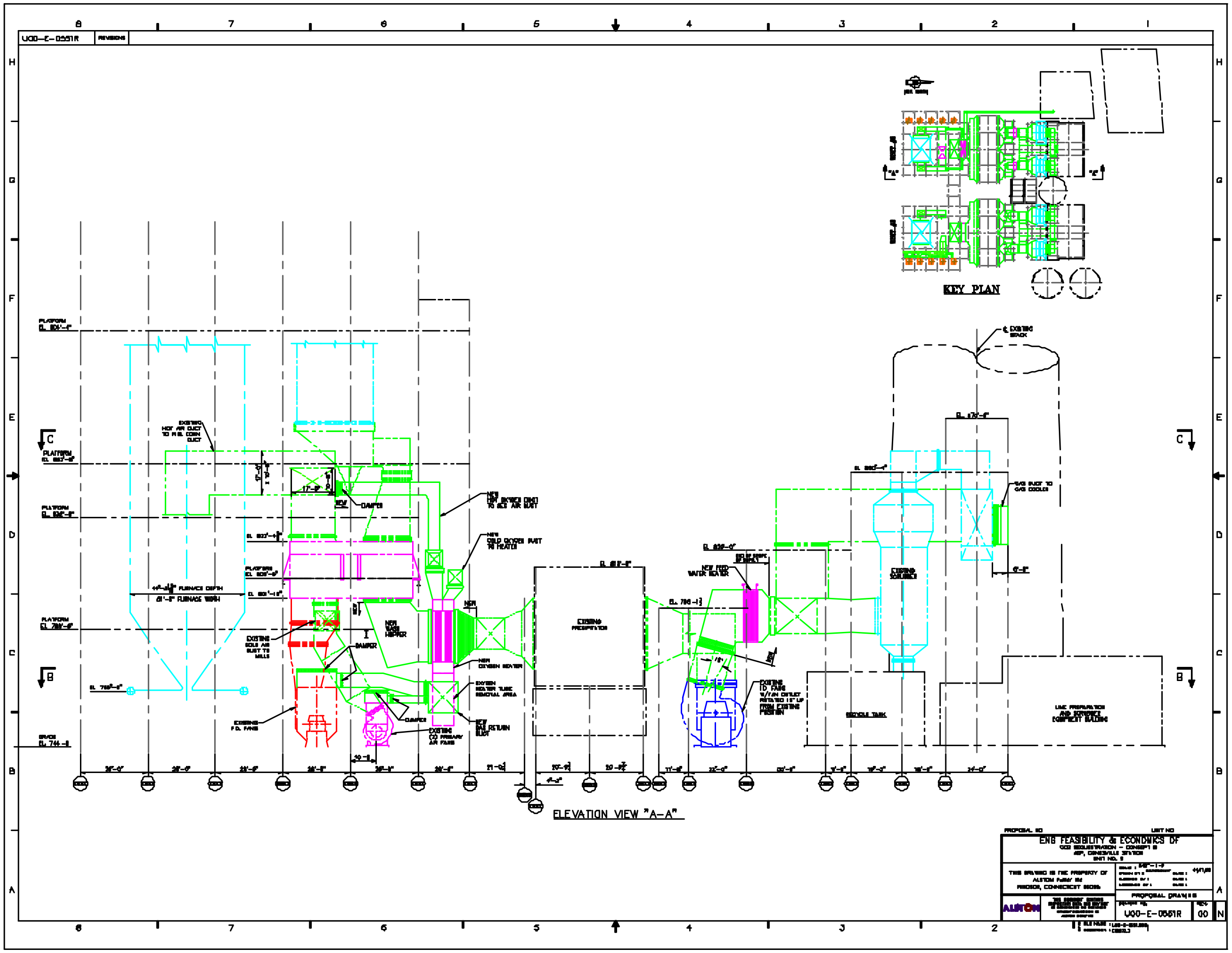


PLOT PLAN
(CONCEPT – B)


U00-E-0551R
KEY PLAN
ELEVATION VIEW "A-A"
PLATFORM EL 864'-4"
PLATFORM EL 853'-6"
PLATFORM EL 832'-6"
PLATFORM EL 789'-6"
GRADE EL 744'-0
EXISTING HOT AIR DUCT TO MILL COMN DUCT
17'-6"
NEW
DAMPER
NEW HOT OXYGEN DUCT TO SEC AIR DUCT
NEW COLD OXYGEN DUCT TO HEATER
44'-3" FURNACE DEPTH
61'-8" FURNACE WIDTH
NEW WASH HOPPER
EXISTING COLD AIR DUCT TO MILLS
NEW OXYGEN HEATER
OXYGEN HEATER TUBE REMOVAL AREA
EXISTING F.D. FANS
EXISTING (2) PRIMARY AIR FANS
NEW GAS RETURN DUCT
EXISTING PRECIPITATOR
NEW FEED WATER HEATER
END OF SCOPE OF SUPPLY
EL 796'-1 1/2"
EXISTING ID FANS W/FAN OUTLET ROTATED 15° UP FROM EXISTING POSITION
EXISTING SCRUBBER
RECYCLE TANK
LIME PREPARATION AND SCRUBBER EQUIPMENT BUILDING
℄ EXISTING STACK
GAS DUCT TO GAS COOLER
6'-8"
24'-0"
PROPOSAL NO
UNIT NO
ENG FEASIBILITY & ECONOMICS OF
CO2 SEQUESTRATION – CONCEPT B
AEP, CONESVILLE STATION
UNIT NO. 5
THIS DRAWING IS THE PROPERTY OF
ALSTOM Power Inc
WINDSOR, CONNECTICUT 06095
ALSTOM
PROPOSAL DRAWING
U00-E-0551R
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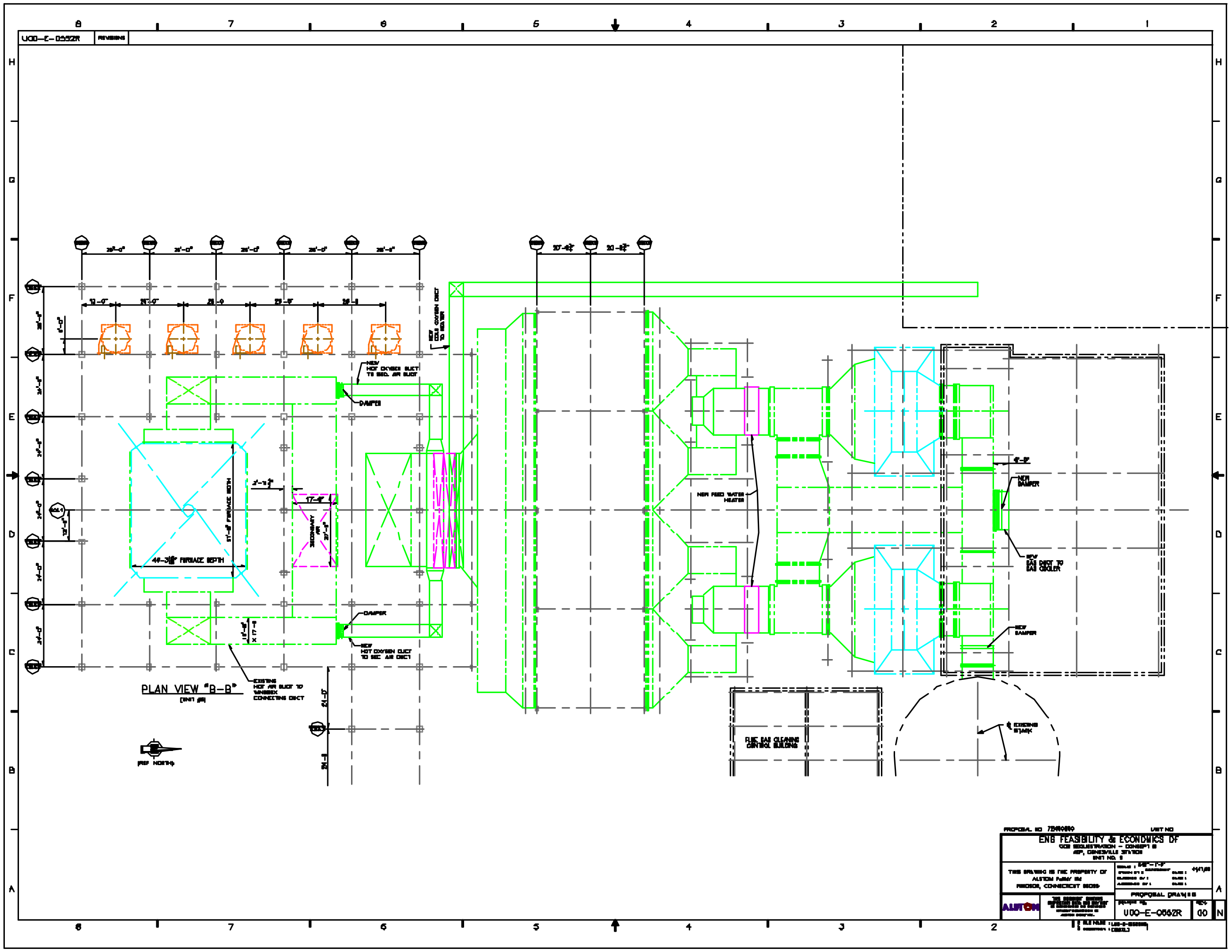
U00-E-0552R
PLAN VIEW "B-B"
(UNIT #5)
(REF NORTH)
NEW COLD OXYGEN DUCT TO HEATER
NEW HOT OXYGEN DUCT TO SEC. AIR DUCT
DAMPER
51'-8" FURNACE WIDTH
44'-3" FURNACE DEPTH
2'-11"
17'-8"
SECONDARY AIR
27'-8"
16'-8" x 17'-8"
DAMPER
NEW HOT OXYGEN DUCT TO SEC. AIR DUCT
EXISTING HOT AIR DUCT TO WINDBOX CONNECTING DUCT
24'-0"
24'-8"
NEW FEED WATER HEATER
8'-0"
NEW DAMPER
NEW GAS DUCT TO GAS COOLER
NEW DAMPER
FLUE GAS CLEANING CONTROL BUILDING
℄ EXISTING STACK
PROPOSAL NO 72410819
UNIT NO
ENG FEASIBILITY & ECONOMICS OF
CO2 SEQUESTRATION - CONCEPT B
AEP, CONESVILLE STATION
UNIT NO. 5
THIS DRAWING IS THE PROPERTY OF
ALSTOM POWER INC
WINDSOR, CONNECTICUT 06095
ALSTOM
SCALE : 3/32" = 1'-0"
PROPOSAL DRAWING
U00-E-0552R
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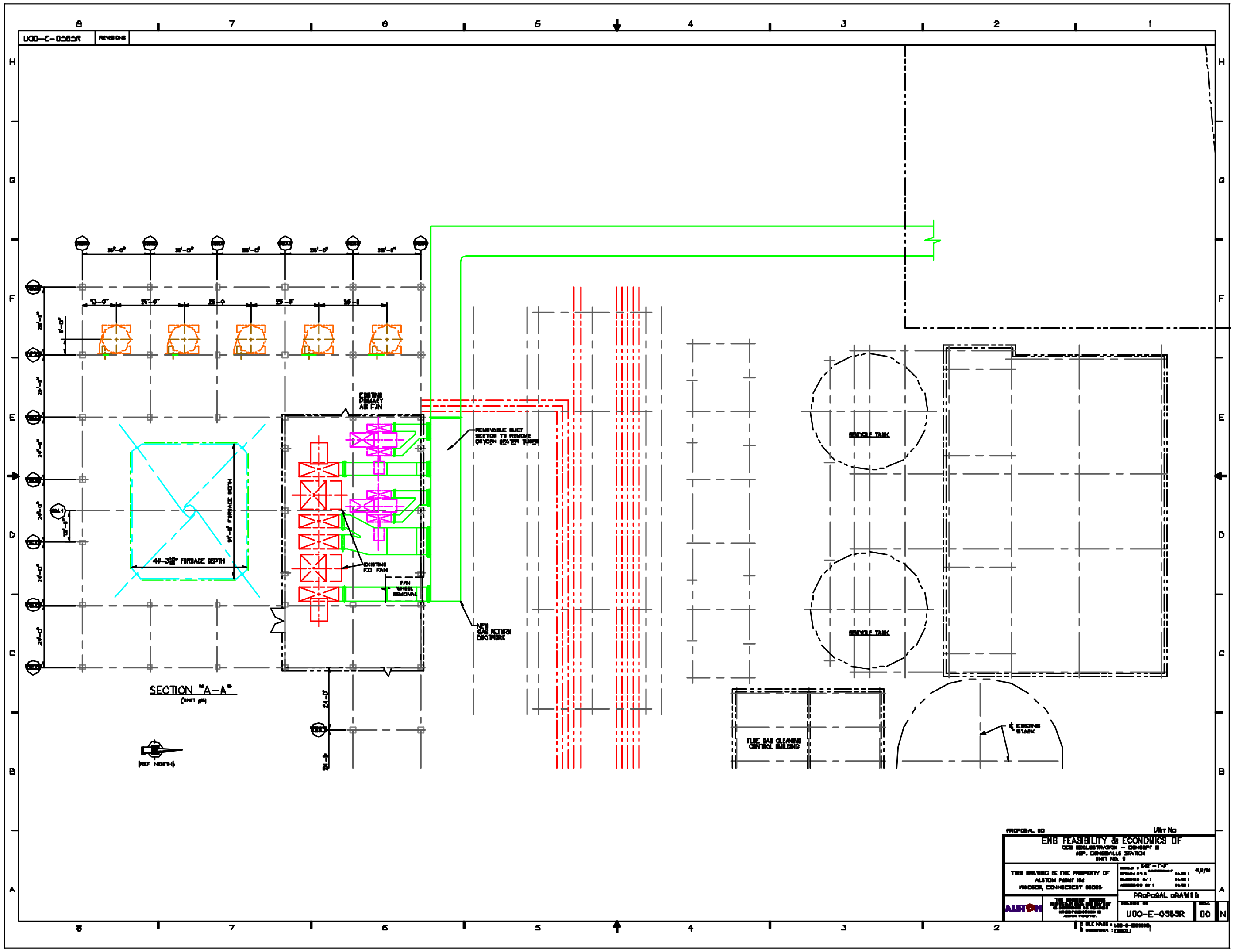
N


U00-E-0585R
REVISIONS
EXISTING PRIMARY AIR FAN
REMOVABLE DUCT SECTION TO REMOVE OXYGEN HEATER TUBES
EXISTING F.D. FAN
FAN WHEEL REMOVAL
NEW GAS RETURN DUCTWORK
44'-3⅜" FURNACE DEPTH
51'-6" FURNACE WIDTH
SECTION "A-A"
(UNIT #5)
(REF NORTH)
RECYCLE TANK
RECYCLE TANK
FLUE GAS CLEANING CONTROL BUILDING
℄ EXISTING STACK
PROPOSAL NO
Unit No
ENG FEASIBILITY & ECONOMICS OF
CO2 SEQUESTRATION - CONCEPT B
AEP, CONESVILLE STATION
UNIT NO. 5
THIS DRAWING IS THE PROPERTY OF
ALSTOM POWER INC
WINDSOR, CONNECTICUT 06095
ALSTOM
SCALE : 3/32" = 1'-0"
DRAWN BY :
DATE :
PROPOSAL DRAWING
U00-E-0585R
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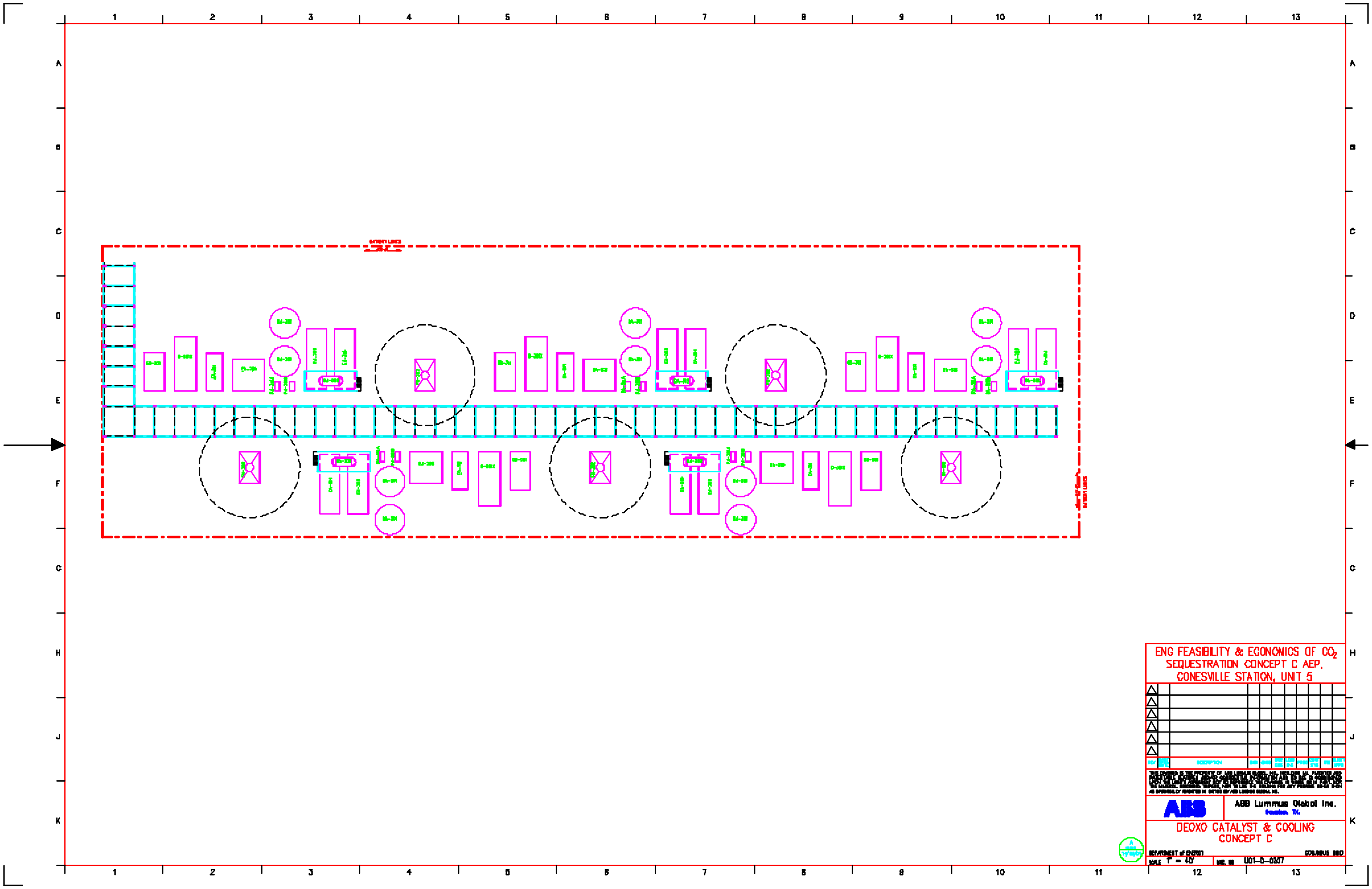
N
FILE NAME :

Concept C:

The plant layout drawings prepared for the Concept C CO_2 Recovery System are as follows:

U01-D-0207 Plot Plan – Concept C: De-Oxy Catalyst & Cooling
U01-D-0210 Plot Plan – Concept C: Flue Gas Cooling & CO_2 Absorption
U01-D-0215 Plot Plan – Concept C: Solvent Stripping
U01-D-0206 Plot Plan – Concept C: CO_2 Compression & Liquefaction
U01-D-0213 Plot Plan – Concept C: Overall Layout Conceptual Plan
U01-D-0202 Plot Plan – Concept C: Modified Overall Site Plan


ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT C AEP,
CONESVILLE STATION, UNIT 5
REV
DESCRIPTION
ABB
ABB Lummus Global Inc.
Houston, TX.
DEOXO CATALYST & COOLING
CONCEPT C
DEPARTMENT of ENERGY
COLUMBUS OHIO
SCALE 1" = 40'
DWG. NO U01-D-0207
A
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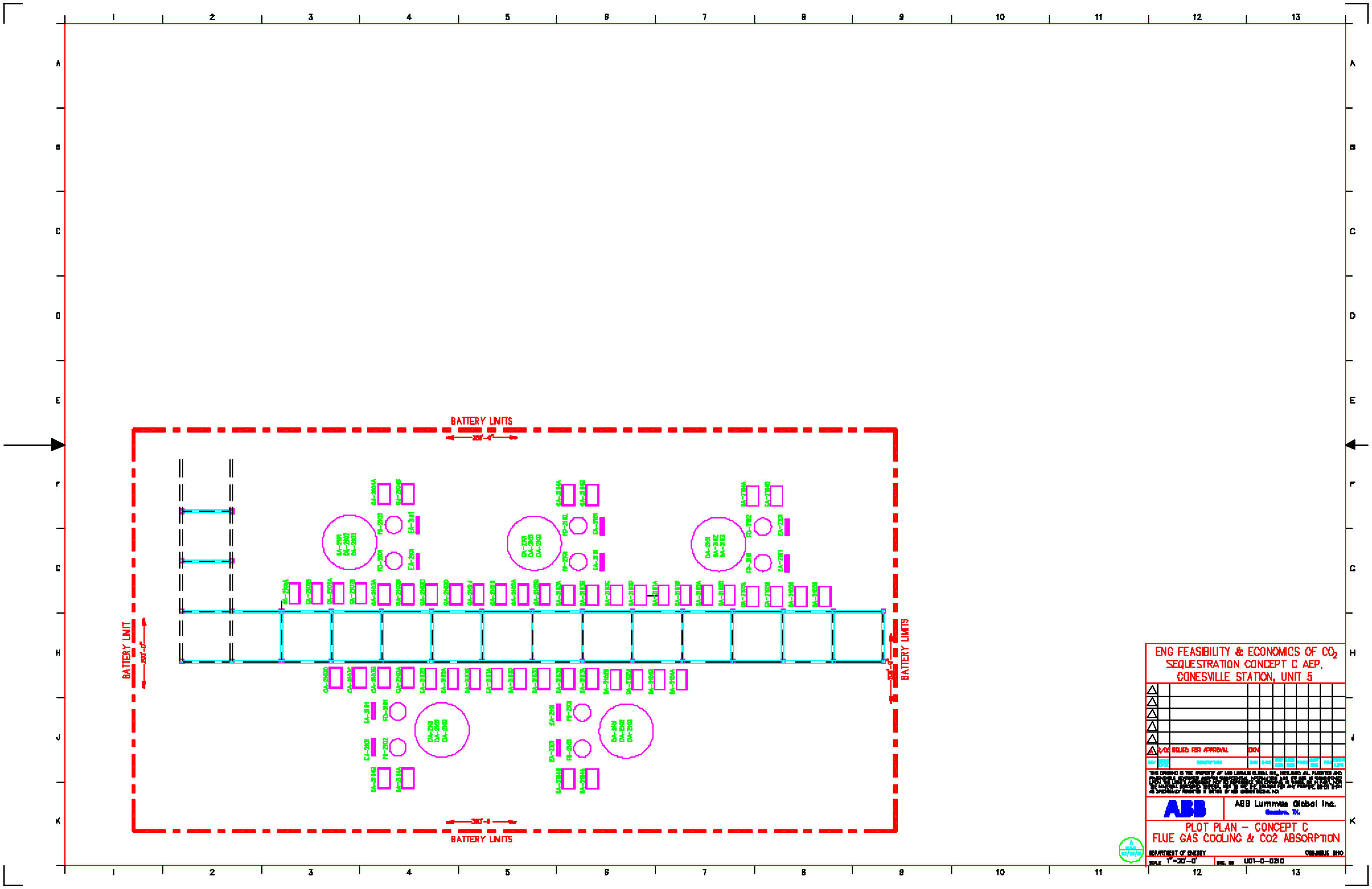
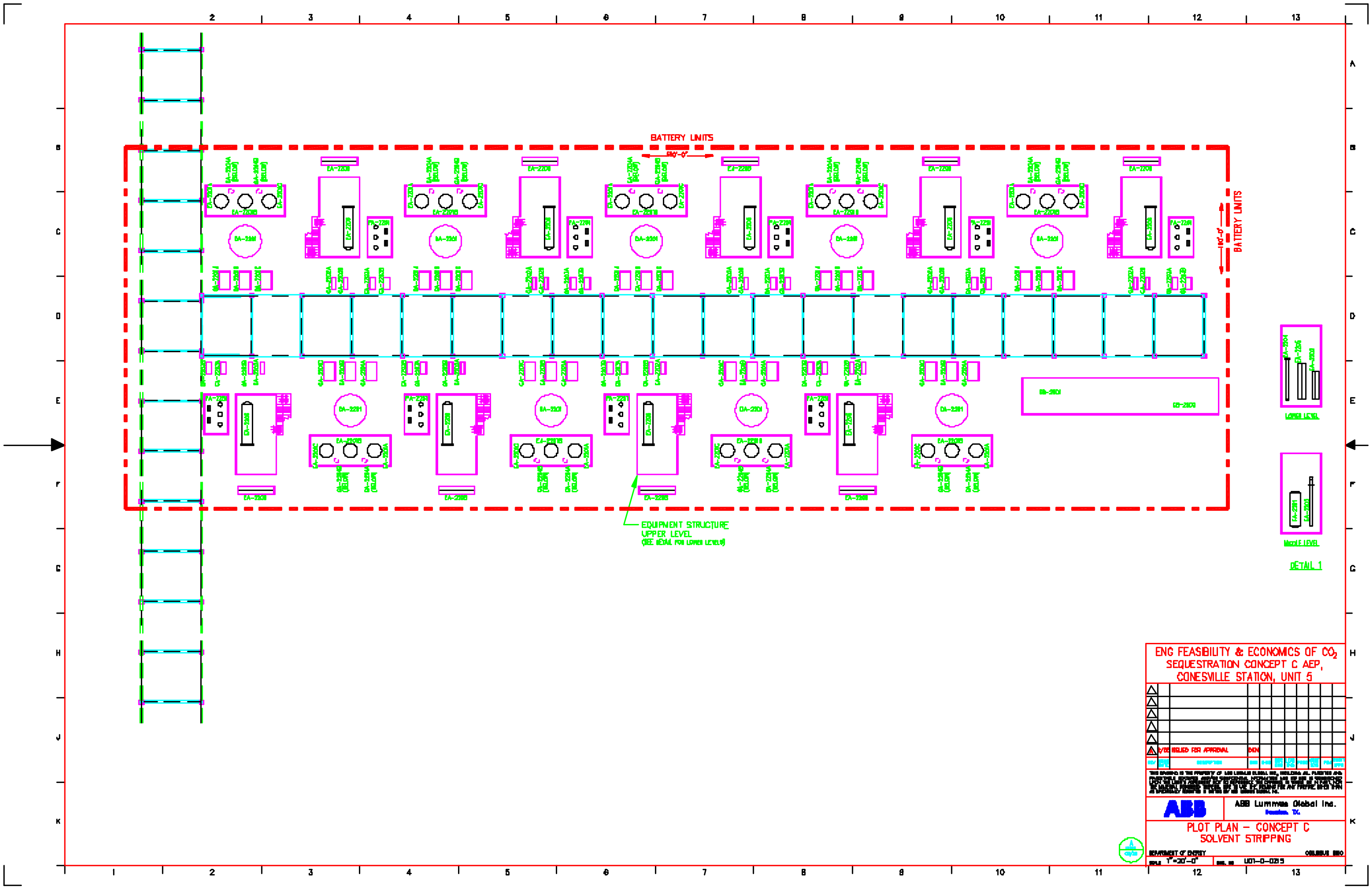

BATTERY LIMITS
BATTERY LIMITS
BATTERY LIMIT
BATTERY LIMITS
ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT C AEP,
CONESVILLE STATION, UNIT 5
A 2/01 ISSUED FOR APPROVAL DKN
ABB
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN – CONCEPT C
FLUE GAS COOLING & CO2 ABSORPTION
DEPARTMENT OF ENERGY
COLUMBUS OHIO
SCALE 1"=20'-0"
DWG. NO U01-0-0210


BATTERY LIMITS
250'-0"
BATTERY LIMITS
110'-0"
EQUIPMENT STRUCTURE
UPPER LEVEL
(SEE DETAIL FOR LOWER LEVELS)
LOWER LEVEL
MIDDLE LEVEL
DETAIL 1
ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT C AEP,
CONESVILLE STATION, UNIT 5
ISSUED FOR APPROVAL
DKN
REV
DESCRIPTION
ABB
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN - CONCEPT C
SOLVENT STRIPPING
DEPARTMENT OF ENERGY
COLUMBUS OHIO
SCALE 1"=20'-0"
DWG. NO. U01-0-0215
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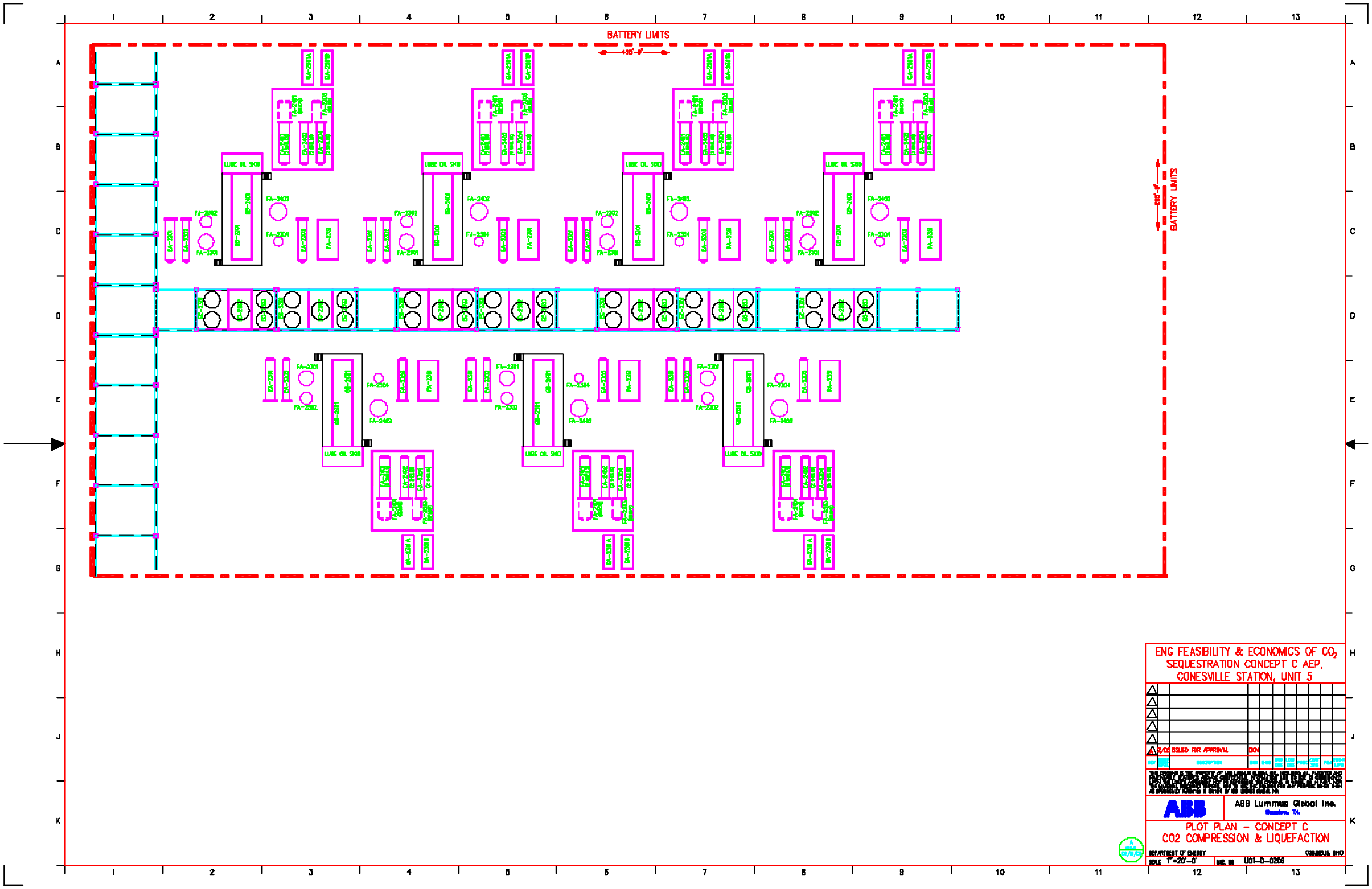
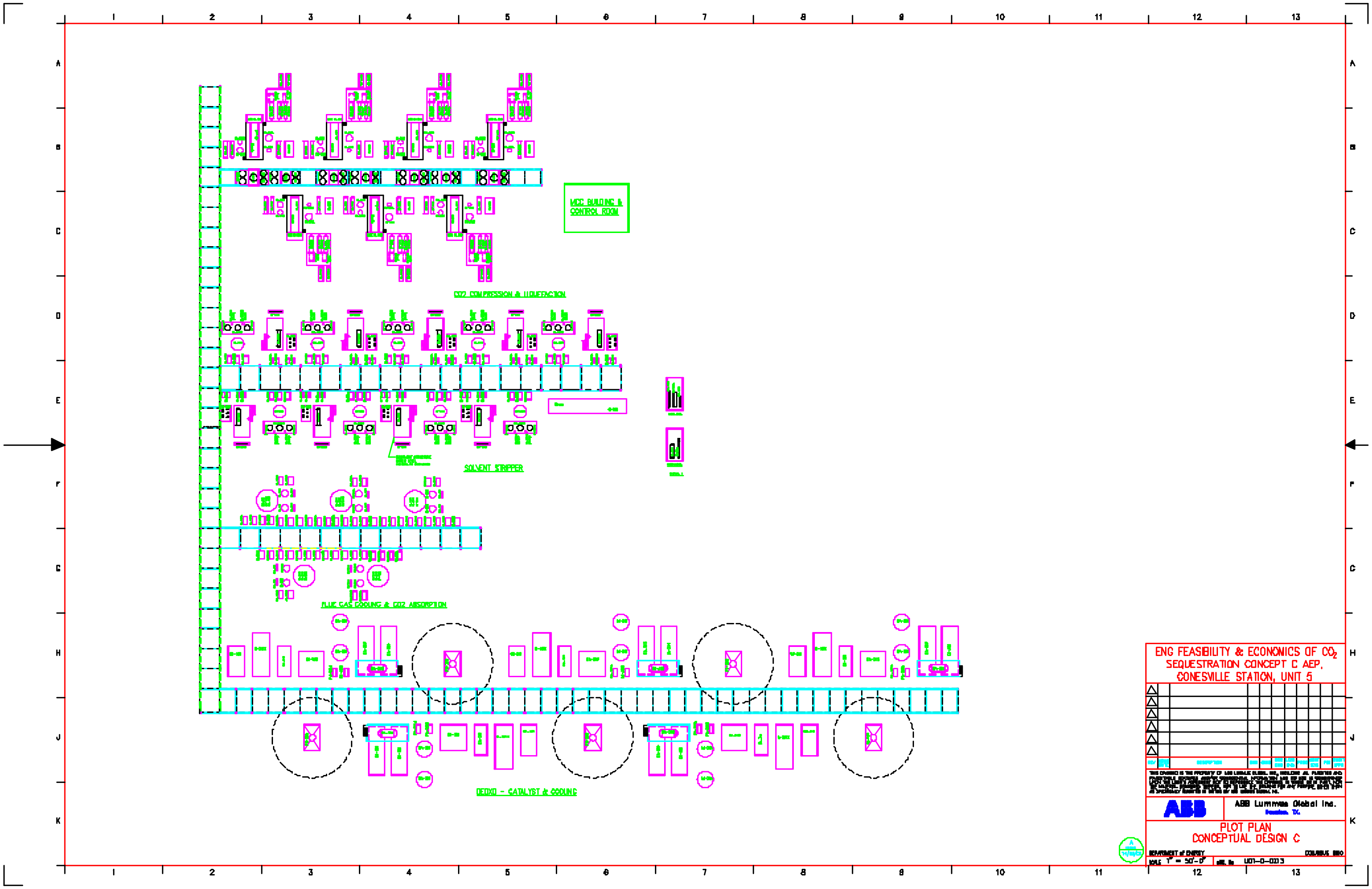

BATTERY LIMITS
BATTERY LIMITS
LUBE OIL SKID
ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT C AEP,
CONESVILLE STATION, UNIT 5
A 2/01 ISSUED FOR APPROVAL DKN
REV
DESCRIPTION
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN – CONCEPT C
CO2 COMPRESSION & LIQUEFACTION
DEPARTMENT OF ENERGY
COLUMBUS, OHIO
SCALE 1"=20'-0"
DWG. NO. U01-D-0206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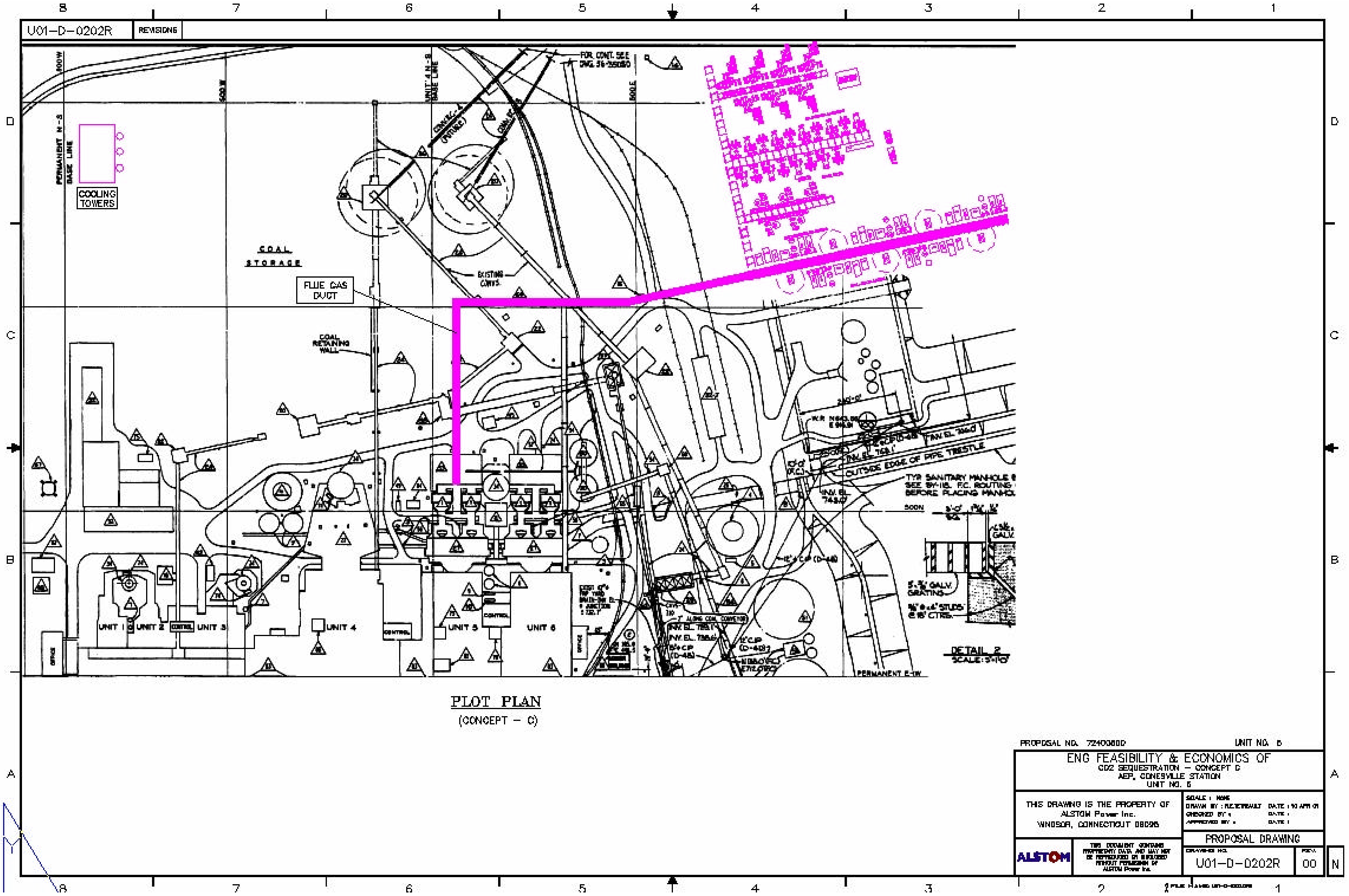
MCC BUILDING &
CONTROL ROOM
CO2 COMPRESSION & LIQUEFACTION
SOLVENT STRIPPER
FLUE GAS COOLING & CO2 ABSORPTION
DEOXO - CATALYST & COOLING
ENG FEASIBILITY & ECONOMICS OF CO2
SEQUESTRATION CONCEPT C AEP,
CONESVILLE STATION, UNIT 5
REV
DESCRIPTION
ABB Lummus Global Inc.
Houston, TX.
PLOT PLAN
CONCEPTUAL DESIGN C
DEPARTMENT of ENERGY
COLUMBUS OHIO
SCALE 1" = 50'-0"
U01-0-0213
A



PLOT PLAN
(CONCEPT – C)

APPENDIX II - EQUIPMENT LISTS

This appendix contains equipment lists for the CO$_2$ Capture Systems of Concepts A, B, and C.

Concept A Equipment List and Data:

Number of Trains	Tag no.	Description	Size Parameters	Material
5	DA-2101	Direct Contact Flue Gas Cooler	27' ID x 34' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
5	DA-2102	CO2 Absorber	27' ID x 92' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
9	DA-2201	Solvent Stripper	16' ID x 100' S/S, DP 35 psig/ FV	CS/SS
9	EA-2201	Solvent Stripper Reboiler	217 MMBTU/HR DP S/T, 50 psig/ 60 psig	CS/SS
9	EA-2203	Solvent Stripper Reclaimer	5.6 MMBTU/HR, DP S/T, 120 psig/ 190 psig	CS/TI
9	EA-2204	Solvent Reclaimer Effluent Cooler	5 MMBTU/HR, DP S/T, 125 psig, 100 psig	CS/TI
9	EA-2206	Solvent Stripper CW Condenser	41.6 MMBTU/HR, DP S/T, 35 psig/ 100 psig	SS/TI
7	EA-2301	CO2 Compr. 1st Stage Aftercooler	1.9 MMBTU/HR, DP S/T, 75 psig/ 100 psig	SS/TI
7	EA-2302	CO2 Compr. 2nd Stage Aftercooler	1.3 MMBTU/HR, DP S/T, 125 psig/ 100 psig	SS/TI
7	EA-2303	CO2 Compr. 3rd Stage Aftercooler	1 MMBTU/HR, DP S/T, 235 psig/ 100 psig	CS/TI
7	EA-2304	CO2 Condenser	19 MMBTU/hr DP S/T, 235 psig/ 300 psig	CS/TI
5	EA-2101	Direct Contact Flue Gas Water Clr	4.8 MMBTU/HR, DP P/U, 50 psig/ 100 psig	TI
9	EA-2205	Rich / Lean Solvent Exchanger	210 MMBTU/HR, DP P/P, 135 psig/ 155 psig	SS316
9	EA-2202	Lean Solvent Cooler	101.8 MMBTU/HR, DP P/U 135 psig/ 100 psig	TI
7	EA-2401	Propane Refrig Condenser	20.45 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
7	EA-2402	Propane Refrig Subcooler	5.9 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
7	EC-2301	CO2 compressor 1st stage air cooler	2.94 MMBTU/HR, DP 75 psig	SS
7	EC-2302	CO2 compressor 2nd stage air cooler	3.1 MMBTU/HR, DP 125 psig	SS
7	EC-2303	CO2 compressor 3rd stage air cooler	4.6 MMBTU/HR, DP 235 psig	SS
9	EC-2201	Solvent stripper bottoms cooler	80.3 MMBTU/HR, DP 135 psig	SS
9	FA-2201	Solvent Stripper Reflux Drum	5' ID x 16' S/S, DP 35 psig/ FV	304L
7	FA-2301	CO2 Compressor 2st Stage Suction Drum	7'- 6" ID x 10' S/S, DP 75 psig	CS/SS
7	FA-2302	CO2 Compressor 3rd Stage	6' ID x 10' S/S, DP 125 psig	CS/SS

		Suction Drum		
7	FA-2303	Liquid CO2 Surge Drum	4'- 6" ID x 14' S/S, DP 235 psig	KCS
7	FA-2304	CO2 Compressor 3rd stage Discharge KO Drum	4' 6" ID x 10' S/S, DP 235 psig	CS/SS
7	FA-2401	Propane Refrig Surge Drum	10' ID x 30' S/S, DP 300 psig	CS
7	FA-2402	Propane Refrig Suction Scubber	8' 6" ID x 12' S/S, DP 300 psig	LTCS
3	FB-2503	Caustic day tank	2' ID x 4' S/S, DP atm	CS
5	FD-2101	DCC Water Filter	205 gpm, DP 35 psig	SS
5	GA-2101 A/B	Wash Water Pump	1425 gpm, DP 29 psi	DI/SS
5	GA-2102 A/B	Direct Contact Cooler Water Pump	205 gpm, DP 36 psi	SS/SS
5	GA-2103 A/B/C/D	Rich Solvent Pump	3450 gpm, DP 92 psi	SS/SS
9	GA-2201A/B/C	Lean Solvent Pump	3000 gpm, DP 85 psi	SS/SS
9	GA-2202 A/B	Solvent Stripper Reflux Pump	310 gpm, DP 75 psi	DI/SS
9	GA-2203 A/B	Filter Circ. Pump	290 gpm, DP 91 psi	SS/SS
9	GA-2204 A/B	LP Condensate Booster Pump	512 gpm, DP 237 psi	CI/ SS
7	GA-2301 A/B	CO2 Pipeline Pump	217 gpm, DP 1815 psi	CS/CS
3	GA-2501	Caustic metering pump	.45 gpm, DP 50 psi	SS
7	GB-2301	CO2 Compressor (Motor driven)	4480 hp	SS wheels
7	GB-2401	Propane Refrig Compressor	3075 hp	LTCS
1	GB-2500	LP steam turbine/ generator	83365 hp	
9	PA-2551	Corrosion Inhibitor Package	Metering, 25 lb/ hr	
9	PA-2251	Solvent Filter Package	140 gpm	
7	PA-2351	CO2 Dryer Package	4 driers, 200 hp compressor, elec heater, cooler	
1		Crane for Compr. Bldg. Flue gas ducting		
1	PA-2551	Cooling Tower	22000 gpm, includes basin, pumps, chlorine injection	
1	PA-2552	Cooling tower blowdown treatment package	100 gpm sand filters and dechlorinator, hypochlorite storage tank	

Concept B Equipment List and Data:

Number of Trains	Tag No.	Service	Sizing Parameters	MOC
	DA	Columns and Towers		
1	DA-101 A/B/C	Direct Contact Flue Gas Cooler	27' ID x 30' S/S, DP 10 psig, 3 psi vacuum	CS w/ SS liner
2	DA-102	CO2 Rectifier	6'/ 13' ID x 30'/ 10' S/S, DP 425 psig	LTCS
	E	Heat Transfer Equipment		
	EA	Shell & Tube Exchangers		
2	EA-101	Flue Gas Compressor 1 Stage Trim Cooler	7.25 MMBTU/h, DP S/T, 85 psig/ 85 psig	CS/SS
2	EA-102	Flue Gas Compressor 2 Stage Trim Cooler	4.2 MMBTU/HR, DP S/T, 175 psig/ 125 psig	CS/SS
2	EA-103	Flue Gas Compressor 3 Stage Trim Cooler	4.1 MMBTU/HR, DP S/T, 425 psgi/ 300 psig	CS/SS
2	EA-104	CO2 Condenser	64.3 MMBTU/HR, DP S/T, 300 psig/ 425 psig	LTCS/ LTCS
2	EA-201	Refrig condenser	68.8 MMBTU/HR, DP S/T, 300 psig/ 200 psig	CS/CS
2	EA-202	Refrig Subcooler	20 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/ LTCS
2	EA-107	CO2 Rectifier Condenser	5.2 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		
1	EB-101	Water Cooler	Total 313 MMBTU/HR, DP P/U, 65 psig/ 100 psig	CS
	EC	Air Coolers		
2	EC-101	Flue Gas Compressor 1 Stage Aftercooler	25.5 MMBTU/HR, DP 85 psig	SS
2	EC-102	Flue Gas Compressor 2 Stage Aftercooler	21.6 MMBTU/HR, DP 175 psig	SS
2	EC-103	Flue Gas Compressor 3 Stage Aftercooler	21.4 MMBTU/HR, DP 425 psig	SS
	FH	Heaters		
2	FH-101	Dryer Regeneration Gas Heater	Gas fired, 8.5 MBTU/HR	
	FA	Drums and Vessels		
2	FA-101	Flue Gas Compressor 2nd Stage Suction Drum	10' ID x 26' S/S, DP 85 psig	CS w/ SS liner
2	FA-102	Flue Gas Compressor 3rd Stage Suction Drum	8' ID x 20' S/S DP 175 psig	CS w/ SS liner
2	FA-103	Flue Gas Compressor Third Stage Discharge K/O Drum	8' ID x 12' S/S, DP 425 psig	CS w/ SS liner
2	FA-201	Refrig Surge Drum	16' ID x 34' S/S DP 300 psig	CS
2	FA-202	Refrig Suction Srubber	16' ID x 16' S/S, DP 300 psig	ITCS

	FD	Filters and Dryers		
1	FD-101	Water Filter	6 units, 2422 gpm each, DP 50 psig	SS
2	FD-102	Flue Gas Filter	2718 ACFM, DP 425 psig	CS
	FF	Dryers (Desiccant Type)		
2	FF-101 A/H	Flue Gas Dryer	Six Vessels, 5 drying, 1 regeneration	CS
	GA	Pumps Centrifugal		
1	GA-101 A/B/C/D	Water Pump	4900 gpm, DP 35 psi	CI w/ SS impeller
2	GA-103A/B	CO2 Pipeline pump	800 gpm, DP 1655 psi	ITCS
	GB	Compressors & Blowers		
2	GB-101	Flue Gas Compressor	3 stages Includes Lube/Seal Oil Systems, 18818 kW	SS
2	GB-102	Propane Refrig Compressor	2 Stage, Includes lube oil/ seal oil system, 7188 kW	ITCS
1	GB-100	Flue Gas Blower	2750 HP	SS
1	GB-103	Flue Gas Booster	500 HP	SS
	PA	Packaged Equipment		
1	PA-101A/B	Large Air Separation Unit	3925 tons/ day 99% pure O2	
1	PA-102 A/B	Small Air Separation Unit	550 tons/ day 99% pure O2	
1	PA-103	Cooling Tower	93200 gpm Includes pumps, basin, and chlorine injection, hypochlorite storage tank	
1	PA-104	Cooling Tower Make-up water Purification unit		
		Ducting		

Concept C Equipment List and Data:

Number of Trains	Tag no.	Description	Size Parameters	Material
5	D-301X	Flue gas scrubber	192500 ACFM, 213 lb/hr SO2	316L
5	DA-301	De-oxy Catalyst Reactor	2 shells 32' dia x 50' S/S DP 14 psig	CS w/ refractory
5	DA-302	Reactor Effluent/ Steam Drum	8' 6" ID x 16' S/S, DP 2000 psig	CS
5	DA-303	De-oxy Catalyst Reactor 2		
5	DA-2101	Direct Contact Flue Gas Cooler	27' ID x 34' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
5	DA-2102	CO2 Absorber	27' ID x 92' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
5	DA-2103	Bottom Direct Contact Flue Gas Cooler	27' ID x 20' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
9	DA-2201	Solvent Stripper	16' ID x 100' S/S, DP 35 psig/ FV	CS/SS
5	EA-301	Feed/ Effluent Exchanger	167.8 MMBTU/HR, DP 5 psig, 5 psig	CS
5	EA-303	Reactor effluent/ steam superheater	24 MMBTU/HR, DP 2100 psig	CS
5	EA-304	Reactor effluent/ steam generator	70 MMBTU/HR, DP 2100 psig	CS
5	EA-305	Dry Flue Gas Cooler	42.5 MMBTU/HR, DP 350 psig	CS
9	EA-2201	Solvent Stripper Reboiler	167 MMBTU/HR DP S/T, 50 psig/ 60 psig	CS/SS
9	EA-2203	Solvent Stripper Reclaimer	5.6 MMBTU/HR, DP S/T, 120 psig/ 190 psig	CS/TI
9	EA-2204	Solvent Reclaimer Effluent Cooler	5 MMBTU/HR, DP S/T, 125 psig, 100 psig	CS/TI
9	EA-2206	Solvent Stripper CW Condenser	45.7 MMBTU/HR, DP S/T, 35 psig/ 100 psig	SS/TI
9	EA-2208	Solvent stripper bottoms cooler	66.7 MMBTU/HR, DP P/U 135 psig/ 100 psig	SS/TI
7	EA-2301	CO2 Compr. 1st Stage Aftercooler	2 MMBTU/HR, DP S/T, 75 psig/ 100 psig	SS/TI
7	EA-2302	CO2 Compr. 2nd Stage Aftercooler	1.35 MMBTU/HR, DP S/T, 125 psig/ 100 psig	SS/TI
7	EA-2303	CO2 Compr. 3rd Stage Aftercooler	1.1 MMBTU/HR, DP S/T, 235 psig/ 100 psig	CS/TI
7	EA-2304	CO2 Condenser	20 MMBTU/hr DP S/T, 235 psig/ 300 psig	CS/TI
5	EA-2101	Direct Contact Flue Gas Water Clr	77.3 MMBTU/HR, DP P/U, 50 psig/ 100 psig	TI
9	EA-2205	Rich / Lean Solvent Exchanger	133 MMBTU/HR, DP P/P, 135 psig/ 155 psig	SS316
9	EA-2202	Lean Solvent Cooler	60 MMBTU/HR, DP P/U 135 psig/ 100 psig	TI
7	EA-2401	Propane Refrig Condenser	22 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
7	EA-2402	Propane Refrig Subcooler	6.2 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS

7	EC-2301	CO2 compressor 1st stage air cooler	3.2 MMBTU/HR, DP 75 psig	SS
7	EC-2302	CO2 compressor 2nd stage air cooler	3.3 MMBTU/HR, DP 125 psig	SS
7	EC-2303	CO2 compressor 3rd stage air cooler	4.9 MMBTU/HR, DP 235 psig	SS
9	FA-2201	Solvent Stripper Reflux Drum	5' ID x 16' S/S, DP 35 psig/ FV	304L
7	FA-2301	CO2 Compressor 2st Stage Suction Drum	7'- 6" ID x 10' S/S, DP 75 psig	CS/SS
7	FA-2302	CO2 Compressor 3rd Stage Suction Drum	6' ID x 10' S/S, DP 125 psig	CS/SS
7	FA-2303	Liquid CO2 Surge Drum	4'- 6" ID x 14' S/S, DP 235 psig	KCS
7	FA-2304	CO2 Compressor 3rd stage Discharge KO Drum	4' 6" ID x 10' S/S, DP 235 psig	CS/SS
7	FA-2401	Propane Refrig Surge Drum	10' ID x 30' S/S, DP 300 psig	CS
7	FA-2402	Propane Refrig Suction Scubber	8' 6" ID x 12' S/S, DP 300 psig	LTCS
3	FB-2503	Caustic day tank	2' ID x 4' S/S, DP atm	CS
5	FD-2101A/B	DCC Water Filter	1645 gpm each, DP 35 psig	SS
5	FH- 302	Start-up heater	54.6 MMBTU/HR	CS
5	GA-301 A/B	HP Condensate Booster pumps	163 gpm, DP 1500 psi	DI/SS
5	GA-2101 A/B	Wash Water Pump	1425 gpm, DP 29 psi	DI/SS
5	GA-2102 A/B	Direct Contact Cooler Water Pump	3289 gpm, DP 36 psi	SS/SS
5	GA-2103 A/B/C/D	Rich Solvent Pump	2435 gpm, DP 92 psi	SS/SS
5	GA-2104 A/B	Bottom Direct Contact Cooler Water Pump	325 gpm, DP 36 psi	SS/SS
9	GA-2201A/B/C	Lean Solvent Pump	2095 gpm, DP 85 psi	SS/SS
9	GA-2202 A/B	Solvent Stripper Reflux Pump	310 gpm, DP 75 psi	DI/SS
9	GA-2203 A/B	Filter Circ. Pump	290 gpm, DP 91 psi	SS/SS
9	GA-2204 A/B	LP Condensate Booster Pump	395 gpm, DP 237 psi	DI/SS
7	GA-2301 A/B	CO2 Pipeline Pump	228 gpm, DP 1815 psi	CS/CS
3	GA-2501	Caustic metering pump	.45 gpm, DP 50 psi	SS
5	GB-301	Flue Gas Blower	3350 hp	SS
7	GB-2301	CO2 Compressor (Motor driven)	4750 hp	SS wheels
7	GB-2401	Propane Refrig Compressor	3225 hp	LTCS
1	GB-2500	LP steam turbine/ generator	48809 hp	
1	GB-2501	HP steam turbine/ generator	52421 hp	

9	PA-2251	Solvent Filter Package	140 gpm
7	PA-2351	CO2 Dryer Package	4 dryers, 200 hp compressor, fired heater, cooler
1		Crane for Compr.	
		Flue gas ducting	
1	PA-2551	Cooling Tower	67500 gpm, includes basin, pumps, chlorine injection
1	PA-2552	Make-up water purification package	1570 gpm clarifier w/ b/d sump and pump, and CT b/d dechlorinator, hypochlorite storage tank

VOLUME II

ENGINEERING FEASIBILITY AND ECONOMICS OF CO_2 CAPTURE ON AN EXISTING COAL-FIRED POWER PLANT

FINAL REPORT

VOLUME II
BENCH-SCALE TESTING AND COMPUTATIONAL FLUID DYNAMICS (CFD) EVALUATION

SUBMITTED BY

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AND

US DEPARTMENT OF ENERGY
NATIONAL ENERGY TECHNOLOGY LABORATORY
PITTSBURGH, PENNSYLVANIA 15236
(CONTRACT NO. DE-FC26-99FT40576)

REPORT SUBMITTAL DATE: JUNE 29, 2001

PROJECT PERFORMANCE PERIOD: NOVEMBER 11, 1999 – JUNE 30, 2001

PPL REPORT NO. PPL-01-CT-10

Disclamer

This report was prepared by ALSTOM Power Plant Laboratories with support in part by a grant from both the Ohio Coal Development Office/Ohio Department of Development and the United States Government. Neither the State of Ohio, the United States Government, nor any of their agencies, nor any person acting on their behalf:

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Public Abstract

ALSTOM Power Inc.'s Power Plant Laboratories (ALSTOM) has teamed with American Electric Power (AEP), ABB Lummus Global Inc. (ABB), the US Department of Energy National Energy Technology Laboratory (DOE NETL), and the Ohio Coal Development Office (OCDO) to conduct a comprehensive study evaluating the technical feasibility and economics of alternate CO_2 capture and sequestration technologies applied to an existing US coal-fired electric generation power plant.

The motivation for this study was to provide input to potential US electric utility actions concerning GHG emissions reduction. If the US decides to reduce CO_2 emissions, action would need to be taken to address existing power plants. Although fuel switching from coal to natural gas may be one scenario, it will not necessarily be a sufficient measure and some form of CO_2 capture for use or disposal may also be required. The output of this CO_2 capture study will enhance the public's understanding of control options and influence decisions and actions by government, regulators, and power plant owners in considering the costs of reducing greenhouse gas CO_2 emissions.

The total work breakdown structure is encompassed within three major reports, namely: (1) Literature Survey, (2) AEP's Conesville Unit #5 Retrofit Study, and (3) Bench-Scale Testing and CFD Evaluation. The report on the literature survey results was issued earlier by Bozzuto, et al. (2000). Reports entitled "AEP's Conesville Unit #5 Retrofit Study" and "Bench-Scale Testing and CFD Evaluation" are provided as companion volumes, denoted Volumes I and II, respectively, of the final report. The work performed, results obtained, and conclusions and recommendations derived therefrom are summarized below.

Volume I

The Volume I report discusses three retrofit technology concepts, which were evaluated in conjunction with AEP's Conesville Unit #5, namely:

- **Concept A:** Coal combustion in air, followed by CO_2 separation with Kerr-McGee/ABB Lummus Global's commercial MEA-based absorption/stripping process
- **Concept B:** Coal combustion with O_2 firing and flue gas recycle (oxy-fuel firing)
- **Concept C:** Coal combustion in air with oxygen removal and CO_2 separation by a mixture of primary and tertiary amines, i.e., monoethanolamine (MEA)/methyldiethanolamine (MDEA).

Each of these technologies was evaluated against a baseline case, without CO_2 capture, from the standpoints of performance, impacts on power generating cost, and CO_2 emissions. AEP's 450 MW Conesville Unit No. 5, located in Conesville, Ohio, was used for the power plant case study. All technical performance and cost results associated with the available options were evaluated in a comparative manner.

Major conclusions:

- No major technical barriers exist for retrofitting AEP's Conesville Unit #5 to capture CO_2 for any of the three concepts considered under this study.

- Concept B (oxygen firing with flue gas recycle) appears clearly to be **the best** alternative of the three concepts studied from both an efficiency and incremental COE viewpoint. This study has confirmed two important issues related to firing coal in a CO_2-rich flue gas / O_2 mixture:

    - ➢ **Modifications to the existing steam generator unit pressure arts are not required, and as such will also allow the unit to continue to operate in the conventional air- fired mode.**

- ➢ **CO_2-rich flue gas can be cleaned and compressed with a relatively simple system to provide high purity CO_2 for usage or sequestration.**

- Energy requirements and power consumption are high, resulting in significant decreases in overall power plant efficiencies (HHV basis), ranging from efficiencies of 20.5 to 22.5%, as compared to 35.0% for the Base case (air

firing without CO_2 capture), which is equivalent to an energy penalty ranging from 36% to 41%. That is, the net power plant output was reduced to 77% - 59%.

- Specific investment costs are high, ranging from about 800 to1800 $/kW and from 1000 to 2200 $/kW, corresponding to scenarios with and without replacement power, respectively.

- All cases studied indicate significant increases to the COE as a result of CO_2 capture. The incremental COE as compared to the Base case (air firing without CO_2 capture) ranges from 3.4 to 8.4 ¢/kWh. Similarly CO_2 mitigation costs range from about 42-98 $/ton of CO_2 avoided for the range of cases studied.

- Specific carbon dioxide emissions were reduced from about 2 lb_m/kWh for the Base case to 0.13 – 0.27 lb_m/kWh for the study cases. Recovery or capture of CO_2 ranged from 91 to 96%.

- Nominally, 5-8 acres of new equipment space is needed for Unit #5 alone on the existing 200-acre power plant site, which accommodates a total of 6 units with a total power generating capacity of 2,080 MW.

Major recommendations:

- **A sensitivity study, for Concepts A and C, showing the impact of reducing CO_2 capture percentage on plant efficiency, investment cost, emissions, and cost of electricity. The current work utilized 90% CO_2 capture (nominal). The sensitivity study would investigate nominal CO_2 capture percentages of 70% and 50%.**

- **Detailed analysis of the existing steam turbine for Concept A: In Concept A about 79% of the steam leaving the intermediate-pressure (IP) turbine is extracted from the IP/LP crossover pipe for solvent regeneration. The capability of the existing low-pressure (LP) turbine to operate under these conditions of very low steam flow over the load range should be investigated in detail, preferably by the original equipment manufacturer (OEM).**

- **Optimization of the amine system reboiler steam pressure for Concepts A and C.**

- **Concept B detailed boiler system analysis: A startup/shutdown procedure and system design, particularly the transition from air to oxygen firing (including transient conditions) should be developed. Detailed metal temperature analysis for all heat exchanger sections, including operation at part loads should be analyzed. This should also include furnace wall metal temperatures and analysis of the circulation system. The existing fans should be checked (preferably by the OEM's) for operation under the new conditions. The feasibility of operating the boiler under a slight positive pressure to eliminate air infiltration should also be investigated.**

- **Investigation of Improved oxygen production systems for Concept B, in line with the membrane-based air separation research being conducted by various research groups (e.g., Praxair, Air Products, Norsk Hydro). Also optimization of an integrated boiler and oxygen production system.**

- **Measurement of furnace heat transfer in CO_2/O_2 environments in a proof of concept boiler simulation facility.**

- **Improved solvents, which require lower regeneration energy requirements and/or can be regenerated at a lower temperature level, similar to MHI's KS1 system, but for coal-firing application.**

- **Hybrid process using oxygen-enriched combustion and amine based CO_2 absorption, to accrue, simultaneously, both CO_2 capture and drastic NOx emissions reduction.**

- **Investigation of a new novel high risk CO_2 capture process that would reduce efficiency penalty and retrofit investment cost. This would likely not utilize the existing boiler.**

Volume II

The bench-scale testing consisted primarily of drop-tube (DTFS-1) and thermo-gravimetric analyses (TGA). Three cases were investigated: (1) an air-fired case (Base case); (2) a case in which the N_2 was replaced by an equal mass of CO_2 (Constant Mass case); and (3) a case in which the N_2 volume was replaced by an equal volume of CO_2 (Constant Volume case). The Constant Mass case represents the appropriate retrofit scenario in a commercial pulverized coal firing application, albeit without flue gas recycling.

Major conclusions:

(a) Based on TGA results

- ***B*oth the Conesville Unit #5 and Pittsburgh Seam #8 coals have, for air (Base case) and O_2/CO_2 environments (Constant Mass/Volume cases), similar burning characteristics, as they both go to completion within the same temperature range (550-575 ^oC). Burning either of the coals in air or any of the O_2/CO_2 mixture ratios used here would produce virtually identical results; and (2) both coals have similar reactivity characteristics.**

(b) Based on DTFS-1 results

- Residence time, as expected, has a significant impact on the overall coal combustion efficiency, and hence, on the unburned carbon emission in the fly ash for both coals under consideration. That is, while the combustion efficiency for both coals is only about 80% (dry-ash-free coal basis) at 0.2 sec., it is more than 98% at greater than 0.5 sec. Correspondingly, unburned carbon emissions range from about 65% to about 2% (dry fly ash basis).

- The impact of reaction medium is also significant. While the overall combustion efficiencies do not seem to exhibit significant differences between the baseline and Constant Mass firing, the more sensitive parameter, unburned carbon (UBC) in the fly ash, on the other hand, clearly shows better performance for the Constant Mass case. That is, the UBC at the furnace outlet was about a factor of 4.5 lower for the Constant Mass case (~2% vs. 9%). This is believed to be due to higher reactant gas temperature, and yet longer residence (0.77 sec. vs. 0.57 sec.). It appears, from the shape of the UBC profiles, that, if given enough residence time, these differences would narrow down considerably, implying that the ultimate performance in both mediums would be similar.

- The coal sample from Conesville Unit #5 appears to perform better than the sample from Pittsburgh seam #8 coal (e.g., 9% vs. 12.5% UBC for the air firing case). However, based on the shape of the UBC profiles, it can be seen that if given additional residence time, they would both be expected to perform similarly.

- NOx emission from oxy-fuel firing is about a factor of 3 (0.73 vs. 2.25 lb/MMBtu) lower than that of the Base case. Correspondingly, sulfur dioxide is lower by about 19% (3.6 vs. 4.3 lb/MMBtu), and CO is significantly lower (0.09 lb/MMBtu vs. a negligible amount).

(c) Based on CFD results

- The baseline case exhibits higher carbon-in-ash (by 1.6 percentage points), higher outlet NOx emissions (by a factor of 2), and higher outlet CO (by a factor of 2) than the Concept-B case. These same computational trends are also qualitatively exhibited by the bench-scale testing.

- The baseline case exhibits a slightly higher peak gas temperature (maximum difference of about 200 °F), and a correspondingly higher average (cross-sectional) gas temperature (difference of 90 °F at the HFOT), than that of the Concept-B case.

- The net wall absorption in the furnace region for the baseline case is larger (by less than 1%) than that of the Concept-B case. However, a significant variation in both calculated and experimental irradiation to the wall between baseline and oxy-firing scenarios was also found in the literature (e.g., +6 to –18%). Differences may be partially attributed to the sensitivity of the irradiation and local emissivities to the aerodynamic and flame patterns, which are, in turn, a function of the furnace and firing system. It is suggested that radiatively absorbing gas species can either enhance or inhibit the irradiation in the vicinity of the wall, depending upon their local temperature and their relative spatial placement.

Major recommendations:

- Carry out a systematic coal fundamentals study concerning this field of endeavor. This study should be geared toward the creation of a database of information depicting the impact of coal nature (using coals of various rank coming from both domestic and international sources) when fired in an oxy-fuel environment on:
    - Coal devolatilization and char oxidation kinetics, and unburned carbon emissions in the fly ash
    - Acid rain-related gaseous (NOx, SO_2, and CO) emissions
    - Air toxics, particularly, mercury (Hg), volatile organic compounds (VOCs), and particulate emissions.
- Derive furnace heat transfer data from natural gas and coal firing in an oxy-fired pilot-scale furnace, which simulates the temperature/time history of a boiler. Use first the data from natural gas firing to validate the CFD code, without the complication of burning particles transiting the boiler. Apply the experience learned from modeling a natural gas fired boiler to model the coal-fired boiler.

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Background

International discussions continue to debate the need for reductions in emission levels of green house gases (GHG). Because fossil fuel fired power plants are among the largest and most concentrated producers of CO_2 emissions, it stands to reason that recovery of CO_2 from the flue gas of such plants has been identified as one of the primary means for reducing CO_2 emissions.

ALSTOM Power Inc.'s Power Plant Laboratories (ALSTOM) teamed with American Electric Power (AEP), ABB Lummus Global Inc. (ABB), the US Department of Energy National Energy Technology Laboratory (DOE NETL), and the Ohio Coal Development Office (OCDO) to conduct a comprehensive study evaluating the technical and economic feasibility of three alternate CO_2 capture technologies applied to an existing US coal-fired electric generation power plant. The steam generator analyzed in this study is the Conesville No. 5 unit, operated by American Electric Power (AEP) of Columbus, Ohio. This steam generator is a nominal 450 MW, coal-fired, subcritical pressure, controlled circulation unit. It has a single cell furnace that employs corner-fired, tilting, tangential burners and which fires bituminous coal from the state of Ohio. The flue gas leaving the boiler system is cleaned of particulate matter in an electrostatic precipitator (ESP) and of SO_2 in a lime-based flue gas desulfurization (FGD) system before being discharged to the atmosphere.

ALSTOM is managing and performing the subject study from its US Power Plant Laboratories office in Windsor, CT. ABB, from its offices in Houston, Texas, is participating as a sub-contractor. AEP is participating by offering their Conesville Generating Station as the case study and cost sharing consultation, along with relevant technical and cost data. AEP is one of the largest US utilities, and as the largest consumer of Ohio coal, is bringing considerable value to the project. Similarly, ALSTOM and ABB are well established as global leaders in the design and manufacturing of steam generating equipment, petrochemical, and CO_2 separation technology. The DOE National Energy Technology Laboratory and the Ohio Coal Development Office provided consultation and funding. All participants contributed to the cost share of this project.

Objectives

The motivation for this study was to provide input to potential US electric utility actions concerning GHG emissions reduction. If the US decides to reduce CO_2 emissions, action would need to be taken to address existing power plants. Although fuel switching from coal to natural gas may be one scenario, it will not necessarily be a sufficient measure and some form of CO_2 capture for use or disposal may also be required. The output of this CO_2 capture study will enhance the public's understanding of control options and influence decisions and actions by government, regulators, and power plant owners in considering the costs of reducing greenhouse gas CO_2 emissions.

The key goals of the study were to evaluate the impacts on the plant output, efficiency, and CO_2 emissions, resulting from the addition of the CO_2-capture systems. Cost estimates were developed for the systems required to produce, extract, clean and compress 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 cost of electricity (COE) and on the mitigation cost for CO_2 (\$/ton of CO_2 avoided) was also evaluated.

Work Scope

The total work breakdown structure is encompassed within three major reports, namely: (1) Literature Survey, (2) AEP's Conesville Unit #5 Retrofit Study, and (3) Bench-Scale Testing and CFD Evaluation. The report on the literature survey results was issued earlier by Bozzuto, et al. (2000). The report entitled "AEP's Conesville Unit #5 Retrofit Study" constitutes Volume I of the final report.

"Bench-Scale Testing and CFD Evaluation" is presented herein as the Volume II report. The work performed, results obtained, and conclusions and recommendations derived therefrom are summarized in the sections that follow.

Bench-Scale Testing

The overall objective of the (Volume II) work segment was to provide critical technical information for evaluating a single option, selected from the three retrofit technology options available, which had the highest probability of realizability. The final option selected was that denoted as: "Coal Combustion in an O_2/CO_2 Environment with CO_2 Recycle." Specific objectives were two-fold: (1) bench-scale testing was to provide experimental combustion and emissions performance data on a specific Ohio coal under conditions where the fuel was to be burned in mixtures of CO_2 and O_2; and (2) a computational fluid dynamics (CFD) modeling effort was to utilize design and operating information from a commercial boiler burning Ohio coal in order to establish a calculation for the conventional burning of coal (in air), and then to evaluate performance in the same commercial unit for the burning of the same coal in mixtures of CO_2 and O_2.

Bench-scale results provided surrogate combustion kinetics, NOx, CO, SO_2 and unburned carbon results on the coal that is presently burned in the Conesville Boilers Units #5 and 6 under conditions that represent conventional combustion (in air) and in various mixtures of CO_2 and O_2. A knowledge of the actual combustion and emissions data from Conesville Unit #5 provided a means whereby the baseline bench-scale results in air were referenced and interpreted; in a sense, it allowed "calibration" of the bench-scale experimental facility to a given point of reference. Referencing the baseline bench-scale results to the actual furnace emissions data allowed a more accurate interpretation of the bench-scale results of the CO_2-O_2 mixtures and permitted a more reliable/extrapolation of how an actual retrofitted boiler would respond to the same CO_2-O_2 mixtures.

The completed literature survey, performed under an initial task of this project (Bozzuto, et al., 2000), indicated that no fundamental combustion performance had been carried out on any coal in CO_2/O_2 media. Hence, the work scope provided first-of-a-kind information to the project. Information derivation and application was carried out as follows:

- First, bench-scale testing of the Conesville coal was carried out in both air (to provide baseline information) and in two mixtures of CO_2 and O_2.

- A comparison of the information on combustion performance in air with information from ALSTOM's proprietary database permitted ALSTOM to select char oxidation kinetic parameters to be used on a surrogate basis during the calibration of the CFD cases.

Coal Characteristics

ASTM standard analyses and ash chemistry analyses were performed for coal samples from AEP's Conesville Unit #5 and Pittsburgh #8 seam. Both of these coal samples are typical of good Eastern US bituminous steaming coals, with fuel values (i.e., FC/VM ratios) of 1.4 and 1.3, respectively. Both coals have high sulfur contents and exhibit medium potentials for slagging the boiler waterwall and convective pass.

TGA Results

The coal samples were size graded to 200x400 mesh. These size-graded materials were subsequently tested in the thermo-gravimetric analyzer (TGA) and the drop-tube furnace system (DTFS-1) equipment to determine their reactivities and unburned carbon emission propensities.

Three cases were evaluated for each coal. The Base case consisted of burning the coal in air. The Constant Mass case consisted of burning the coal in an O_2/CO_2 mixture, such that the mass of O_2 used in the Base case was maintained and the mass of N_2 used in the Base case was replaced with an equal mass of CO_2. The Constant Volume case consisted burning the coal in O_2/CO_2 mixture, such that the mass of O_2 used in the Base case was again maintained and the volume of N_2 used in the Base case was replaced with an equal volume of CO_2.

The combustion efficiency profiles from the TGA indicate that both the Conesville and Pittsburgh #8 coals, for both the air (Base case) and O_2/CO_2 environments (Constant Mass/Volume cases), have similar burning characteristics, as they both go to completion within the same temperature range (550-575°C). Additional analysis of the weight loss as a function of time, as well as the temperature at which the maximum reaction rate occurs, confirms that: (1) burning either of the coals in air or any of the O_2/CO_2 mixture ratios used will produce virtually identical results; and (2) both coals have similar reactivity characteristics.

Drop-Tube Combustion Furnace System-1 Results

The DTFS-1 testing was used to derive quantitative information on the impact of various operating parameters such as particle size, stoichiometry and temperature/time history on unburned carbon loss and gaseous (NOx, CO and SO_2) emissions. The coal feed rate was adjusted for each coal such that the overall stoichiometry (λ) was ~2.50, which is the stoichiometry required to achieve reasonably low unburned carbon emissions in the drop tube. The DTFS-1 was operated in a laminar flow regime, with a Reynolds number in the 150-225 range; the maximum residence time for the conditions evaluated was about 0.8 seconds. DTFS-1 results showed the following:

- Residence time has, as expected, a significant impact on the overall coal combustion efficiency, and hence, on the unburned carbon emission in the fly ash for both coals under consideration. That is, while the combustion efficiency for both coals is only about 80% (dry-ash-free coal basis) at 0.2 sec residence time, it is more than 98% at greater than 0.5 sec. residence times. Correspondingly, unburned carbon emissions range from about 65% to less about 2% (dry fly ash basis) for the same residence times.

- The impact of reaction medium is also significant. While the overall combustion efficiencies do not seem to exhibit significant differences between the baseline and Constant Mass firing, the more sensitive parameter, unburned carbon (UBC) in the fly ash, on the other hand, clearly shows better performance for the Constant Mass case. That is, the UBC at the furnace outlet was more than a factor of 4.5 lower for the Constant Mass case (~2% vs. 9%). This is believed to be due to higher reaction temperature, and yet longer residence time (0.77 sec. vs. 0.57 sec.). It appears, from the shape of the UBC profiles, that, if given enough residence time, these differences would considerably narrow down, implying that the performance in both mediums would be similar.

- The coal sample from Conesville Unit #5 appears to perform better than the sample from Pittsburgh seam #8 coal (e.g., 9% vs. 12.5% UBC for the air firing case). However, based on the shape of the UBC profiles, it can be seen that if given additional residence times, they would both be expected to perform similarly.

- NOx emission from oxy-fuel firing is about a factor of 3 (0.73 vs. 2.25 lb/MMBtu) lower than that of the Base case. Correspondingly, sulfur dioxide is lower by about 19% (3.6 vs. 4.3 lb/MMBtu), and CO is significantly lower (0.09 lb/MMBtu vs. a negligible amount).

- The gaseous emission trends are consistent with results reported by others (Thambimuthu, et al.; 1998; Weller, et al., 1985) and by the CFD results of the present study (Section 3.0, below).

CFD Analysis

ALSTOM Power uses a commercial CFD code (denoted FLUENT™) from Fluent Inc. for its simulations of furnaces and auxiliary equipment. Case preparation of a boiler for the Fluent code was initiated with the collection of the geometry data (from blueprints and schematics), the free areas of the nozzles, and the operating conditions. A CFD grid or mesh was prepared using a combination of EXCEL spreadsheets and a grid-generation package (denoted GAMBIT™) supplied by Fluent Inc. The total number of cells in the grid was approximately 315,000.

Two cases have been compared using the FLUENT™ CFD code – a baseline, nominal 15% excess air case (Case 1 or Base case), and a recirculated flue gas, oxy-fired case (Case 2 or Concept-B case). In the Concept-B case, the nitrogen was largely replaced by carbon dioxide. The cases have been compared on the basis of equivalent nozzle/inlet areas, as well as an equivalent coal-to-oxygen feed ratio (by mass).

The baseline case was calibrated to known furnace performance and operating conditions. Calibration is required because of the uncertainty associated with wall fouling factors, particle reactivities, etc. “Implicit” boundary conditions that were matched during the baseline calibration included:

- the unburned carbon (UBC) in the fly ash (i.e., carbon-in-ash value) at the outlet of the computational domain
- the outlet NOx emissions value at the outlet of the computational domain
- the “horizontal furnace outlet temperature (HFOT),” which is the mass-weighted average temperature across the plane at the apex of the arch (i.e. the horizontal furnace outlet plane or HFOP).

Both the UBC and NOx emissions values were derived from experimental measurements (usually taken downstream of the air preheater). The HFOT value was calculated by an in-house, proprietary design package denoted the Reheat Boiler Program (RHBP). An iterative trial-and-error method was used to calibrate the baseline case by changing the input parameters for char oxidation, NOx, and waterwall conduction resistance, until (after a number of runs) the code calculations matched the desired UBC, the NOx value, and the HFOT, respectively.

Lateral and vertical waterwall heat flux or absorption distributions, as well as the overall heat transfer in the lower and upper furnace, were extracted from the CFD calculations. Total and sectional (area-weighted) waterwall heat absorption rates, in addition to peak heat absorption values, were used to quantify the heat absorption rates that constitute one of the metrics used as a basis of comparison between the cases. Metrics utilized to compare the cases include:

- the unburned carbon (UBC) in the fly ash at the outlet of the computational domain
- the outlet NOx emissions value at the outlet of the computational domain
- the horizontal (and/or vertical) furnace outlet temperature(s) (FOT)
- the outlet CO emissions value at the outlet of the computational domain
- total, sectional, and peak waterwall heat flux or absorption rates

Some of the metrics for the two cases, such as NOx, were compared on a basis that is scaled or referenced to the baseline value, since relative trends may be a more reliable indicator of the expected performance than the absolute magnitudes of the metrics. Line and contour plots for variables of interest were prepared at selected horizontal or vertical planes, as appropriate.

The predictions provided a relative indication of the changes that occurred in the oxy-firing case as a result of deviation from the baseline operating conditions. The difference in inlet nozzle velocities, mixture specific heats, and radiative properties between CO_2 and N_2 had a visible impact on the performance criteria and metrics of interest, e.g., HFOT, waterwall heat flux distributions, NOx emissions, and unburned carbon in the fly ash. Overall, the CFD runs served to provide insight into the flow and reaction patterns of the lower furnace, as well as to provide relative changes in emissions and residual carbon values at the exit plane as a result of changes in operating conditions.

CFD Results and Conclusions

The following computational results have been highlighted:

- The baseline case exhibits higher carbon-in-ash (by 1.6 percentage points), higher outlet NOx emissions (by a factor of 2), and higher outlet CO (by a factor of 2) than the Concept-B case. These same computational trends are also qualitatively exhibited by the bench-scale testing.

- The baseline case exhibits a slightly higher peak gas temperature (maximum difference of about 200 °F), and a correspondingly higher average (cross-sectional) gas temperature (difference of 90 °F at the HFOT), than that of the Concept-B case.

- The net wall absorption in the furnace region for the baseline case is larger (by less than 1%) than that of the Concept-B case. (Conversely, the net wall absorption for the baseline calculation with the RHBP was found to be about 6% lower than the Concept-B case calculated by the RHBP.)

Explanations for the observed trends must rely on physical property differences between the nitrogen, which dominates in the baseline case, and carbon dioxide, which largely supplants the nitrogen in the Concept-B case. The physical species properties and the anticipated consequences are itemized below:

1. Higher molecular weight of CO_2 relative to N_2. For equivalent inlet areas, and as a result of the molecular weight and associated density differences, the baseline case will exhibit larger inlet velocities and a higher penetration capability than the Concept-B case. The inlet velocity differences will further serve to impact the aerodynamic, particle trajectory, mixing, and flame patterns.

2. Higher specific heat of CO_2 relative to N_2. For an equivalent local heat release and heat transfer, the higher specific heat of carbon dioxide will serve to decrease the overall flame temperatures of the recycled flue gas mixture in the Concept-B case, relative to the baseline case.

3. Higher gas emissivity and absorptance of CO_2 relative to N_2. For an equivalent local gas temperature, the higher emissivity of CO_2, relative to the optically transparent N_2, should serve to increase the absorption coefficient (and the irradiation to the wall) in the recycled flue gas case (Concept-B) relative to the baseline case.

The three physical properties listed above may have compensatory rather than cumulative effects. In the present CFD calculations, the resultant, local gas temperatures are not equivalent in the two cases. The higher specific heat of the carbon dioxide, and the associated lower gas temperatures in the Concept-B case, will tend to offset somewhat or compensate for the expected increase in the wall absorption (i.e., anticipated due to the enhanced emissivity of the CO_2.) Furthermore, the spatial distribution of the carbon dioxide across the cross-section may be important; it is hypothesized that high concentrations of CO_2 in the cooler gas mixtures between the wall and the flame may actually act to inhibit the net radiative flux to the walls.

The trend calculated by the CFD code, with respect to the furnace temperature at the HFOP, appears to be qualitatively aligned with the trend calculated by the in-house RHBP code. However, the difference in wall absorption for the Concept-B case relative to the baseline case, calculated by FLUENT™ (change of –1%), does not fully agree with the results of the RHBP (change of +6%). Computed and/or experimental trends that indicate that oxy-firing may induce either an increase in wall absorption (+6%) or a decrease in wall absorption (-18%), relative to the baseline case, are supported in the literature. It is suggested that the absorptive properties of a local CO_2-containing gas mixture in the vicinity of the wall may either act to enhance or inhibit the incident radiative flux to the wall, depending upon its relative spatial placement and mixture temperature. For example, a low-temperature, CO_2-rich mixture near the wall would absorb the radiative flux from the interior "fireball" region; a portion of the energy would be re-emitted at the lower gas temperature, and the remainder of the energy would be used to heat up the gas. The result would be a net reduction in the incident radiative flux to the wall. Conversely, a high-temperature, CO_2-rich mixture near the wall would emit at its local gas temperature, thus enhancing the incident radiative flux to the wall. Therefore, it is suggested that the relative spatial relationships of the cool and hot gas mixtures, the relative composition of the absorbing media, and the proximity of that media to the wall (as induced by the furnace and firing system aerodynamics), may significantly impact the gas-phase irradiation to the wall. Indeed, the aerodynamic and flame patterns that prevail in a given air-fired or oxy-fired furnace may conceivably have as strong an impact on the net wall absorption as the physical property differences themselves (gas-phase emissivity, specific heat, etc.)

The results of computational tools (such as CFD and the RHBP) should be viewed as being qualitative in nature, particularly when they are asked to produce calculations outside of the experience base for which they have been validated. Certainly, the gas-phase and particle-phase radiation property submodels could be enhanced in both codes, and the impact of spatial gradients (caused by the interleaving of hot and cool regions of varying CO_2 composition) on radiative wall flux needs to be investigated more fully. CFD computations have been utilized to gain a more in-depth qualitative understanding of aerodynamic and flame patterns in combustion systems, but have not yet been integrated into routine boiler design procedures. At the present time, the ALSTOM design standards are based upon the experience and expertise built into the RHBP and other design protocols, and those standards must be adhered to in any new design project in order to mitigate risk. Therefore, the results of the RHBP must be regarded as the default standard, both from an engineering experience viewpoint and from a rules-based design viewpoint. However, additional validation work needs to be done in order to confirm the trends for oxy-firing scenarios.

Although differences in physical properties of the transport gases (e.g., air-firing versus oxy-firing) may initially induce undesirable deviations from the performance goals of a field unit, the designer and engineer views such alterations as challenges rather than insurmountable obstacles. Much can be done in the way of operating condition optimization (e.g., spray and nozzle tilt), without major retrofits, to realign steam temperatures and wall absorptions with their desired, target values. Design parameters are often in conflict, and the intelligent designer must work to balance various parameters and operating conditions to achieve the desired outcome. The CFD work performed here was based on certain constraints to maintain equivalency for comparison purposes, which would not necessarily be the case in the field. For example, a potential decrease in furnace wall absorption of several percentage points could be nullified through manipulation of various operating parameters. If required, firing system retrofits and surface modifications could be performed to handle more severe decreases in absorption; certainly new unit design could easily accommodate any necessary alterations in design philosophy.

Recommendations for Future Work

There is, to the authors' knowledge, no fundamental coal combustion work being carried out elsewhere to determine the impact of coal nature in an oxy-fuel environment on unburned carbon and gaseous and particulate emissions. Hence, the following is recommended:

- Carry out a systematic coal fundamentals study concerning this field of endeavor. This study should be geared toward the creation of a database of information depicting the impact of coal nature (using coals of various rank coming from both domestic and international sources) when fired in an oxy-fuel environment on:
    - Coal devolatilization and char oxidation kinetics, and unburned carbon emissions in the fly ash
    - Acid rain-related gaseous (NOx, SO_2, and CO) emissions
    - Air toxics, particularly mercury (Hg), volatile organic compounds (VOCs), and particulate emissions.
- Derive furnace heat transfer data from natural gas and coal firing in an oxy-fired pilot-scale furnace, which simulates the temperature/time history of a boiler. Use first the data from natural gas firing to validate the CFD code, without the complication of burning particles transiting the boiler. Apply the experience learned from modeling a natural gas-fired boiler to model the coal-fired boiler.

1.0 BACKGROUND/OBJECTIVES

ALSTOM Power Inc.'s Power Plant Laboratories led a team, which performed a study titled "Engineering Feasibility and Economics of CO_2 Capture on an Existing Coal-Fired Power Plant," under the auspices of the OCDO and DOE NETL. The other team members were ABB Lummus Global Inc. and American Electric Power (AEP).

The overall objective of this project was to evaluate the technical feasibility and the economics of alternate CO_2 capture and sequestration/use technologies for retrofitting an existing coal-fired power plant. Three alternate CO_2 capture systems (see Section 3 of Volume I) were evaluated, with attention being given to the impact of such systems on an existing boiler, the auxiliary components of the boiler, the boiler and overall plant performance, and the power plant costs, including the cost of electricity.

The overall objective of this work segment was to provide critical technical information for evaluating one of the three retrofit technology options, namely "Coal Combustion in an O_2/CO_2 Environment with CO_2 Recycle." Specific objectives were two-fold: (1) bench-scale testing was to provide experimental combustion and emissions performance data on a specific Ohio coal under conditions where the fuel was to be burned in mixtures of CO_2 and O_2; and (2) a computational fluid dynamics (CFD) modeling effort was to utilize design and operating information from a commercial boiler burning Ohio coal in order to establish a calculation for the conventional burning of coal (in air), and then to evaluate performance in the same commercial unit for the burning of the same coal in mixtures of CO_2 and O_2.

Bench-scale results provided surrogate combustion kinetics, NOx, CO, SO_2 and unburned carbon results on the coal that is presently burned in the Conesville Boilers Units #5 and 6 under conditions that represent conventional combustion (in air) and in various mixtures of CO_2 and O_2. A knowledge of the actual combustion and emissions data from Conesville Unit #5 provided a means whereby the baseline bench-scale results in air were referenced and interpreted; in a sense, it allowed "calibration" of the bench-scale experimental facility to a given point of reference. Referencing the baseline bench-scale results to the actual furnace emissions data allowed a more accurate interpretation of the bench-scale results of the CO_2-O_2 mixtures and permitted a more reliable/extrapolation of how an actual retrofitted boiler would respond to the same CO_2-O_2 mixtures.

The completed literature survey, performed under an initial task of this project (Bozzuto, et al., 2000), indicated that no fundamental combustion performance had been carried out on any coal in CO_2/O_2 media. Hence, the work scope provided first-of-a-kind information to the project. Information derivation and application was carried out as follows:

- First, bench-scale testing of the Conesville coal was carried out in both air (to provide baseline information) and in various mixtures of CO_2 and O_2.

- A comparison of the information on combustion performance in air with information from ALSTOM's proprietary database permitted ALSTOM to select char oxidation kinetic parameters to be used on a surrogate basis during the calibration of the CFD cases.

2.0 BENCH-SCALE TESTING

2.1 Introductory Remarks

ALSTOM Power's DTFS-1 has been used extensively over the years to derive, among other things, the following information (Nsakala, et al., 1986; Nsakala, et al., 1991; Thornock, et al., 1993):

- High temperature volatile matter yields from coals over the whole ASTM rank spectrum and from various synthetic solid fuels (petroleum cokes, refused-derived fuels, process coal gasification chars, beneficiated coal-based fuels, etc.).
- Fuel nitrogen conversion from coals over the whole ASTM rank spectrum and from the same various synthetic solid fuels given above.
- Gasification and devolatilization kinetic parameters for coals over the whole ASTM rank spectrum and for the same various synthetic solid fuels given above.
- Oxidation kinetic parameters of chars prepared from coals over the whole ASTM rank spectrum and from the same various synthetic solid fuels given above.

This information is used, where appropriate, in conjunction with coal combustion modeling with CFD and other proprietary in-house codes for the purpose of predicting (principally) unburned carbon loss and NOx emissions in pulverized fuel fired utility power boilers.

2.2 Analysis of Test Coals

Tables 2.1 and 2.2 present the ASTM standard analyses and ash chemistries of coal samples from AEP's Conesville Unit #5 and Pittsburgh #8 seam. Both of these coal samples are typical of good Eastern US bituminous steaming coals, with fuel values (i.e., FC/VM ratios) of 1.4 and 1.3, respectively. Consistent with the majority of the coal samples from this geological province, they have high sulfur contents (2.4 and 3.5 lb_m S/MMBtu, respectively). Their coal ashes are typical of "Eastern" coal ashes, in which the iron contents are greater than the sums of alkali and alkaline earth contents. Based on the commonly used ash slagging and fouling indices, in conjunction with ash contents and fusibility temperatures, these coal ashes have only medium potentials of slagging the boiler waterwall and convective pass. The plant personnel from Conesville Generating Station confirmed this behavior, with respect to their Unit #5.

The coal samples in Table 2.1 were size graded to 200x400 mesh. These size-graded materials (Table 2.3) were subsequently tested in the TGA and DTFS-1 equipment to determine their reactivities and unburned carbon emission propensities. It is important to note that the chemical compositions of these size fractions are comparatively consistent with those of the parent coal samples they were prepared from.

Table 2. 1: Standard ASTM Analysis of Test Coals

QUANTITY	CONESVILLE #5 COAL (hvAb)		PITTSBURGH #8 COAL (hvAb)	
	As Rec'd	daf	As Rec'd	daf
Coal Properties				
Moisture	10.1		8.2	
Volatile Matter	32.7	41.6	34.8	43.9
Fixed Carbon (Diff.)	45.9	58.4	44.5	56.1
Ash	11.3		12.5	
Hydrogen	4.3	5.5	3.9	4.9
Carbon	63.2	80.4	64.3	81.1
Sulfur	2.7	3.4	4.1	5.2
Nitrogen	1.3	1.7	1.2	1.5
Oxygen (Diff.)	7.1	9.0	5.8	7.3
FC/VM	1.40	---	1.28	---
Sulfur loading, lb.S/MMBtu	2.39		3.50	
HHV (Btu/lb.)	11293	14368.	11726	14787.
lb. Ash/MMBtu	10.0		10.7	
HGI	50		63	
Ash Properties				
Fusibility Temperatures, (°F)				
I.T.	2125		2140	
S.T.	2140		2185	
H.T.	2165		2240	
F.T.	2180		2290	
Ash Comp., (Wt.%, Dry)				
SiO_2	45.9		47.7	
Al_2O_3	21.7		19.9	
Fe_2O_3	20.4		22.9	
CaO	2.9		3.0	
MgO	0.9		0.7	
Na_2O	0.9		0.6	
K_2O	2.0		2.1	
TiO_2	1.1		1.0	
P_2O_5	0.3		0.5	
SO_3	3.0		1.5	
MnO	0.1		0.1	
TOTAL	99.2		99.9	

Table 2. 2: Characteristics of Test Coal Ash Samples

PARAMETER	Conesville Unit #5 Coal Data		Pittsburgh #8 Coal Data	
Fusibility Temperatures (°F)				
I.T.	2125		2140	
S.T.	2140		2185	
H.T.	2165		2240	
F.T.	2180		2290	
Composition, Wt.% (dry)				
SiO_2	45.9		47.7	
Al_2O_3	21.7		19.9	
Fe_2O_3	20.4		22.9	
CaO	2.9		3.0	
MgO	0.9		0.7	
Na_2O	0.9		0.6	
K_2O	2.0		2.1	
TiO_2	1.1		1.0	
P_2O_5	0.3		0.5	
SO_3	3.0		1.5	
Total	99.1		99.9	
Coal Sulfur Content, Wt.% (dry)	3.0		4.4	
Ash Type	**Eastern Bit.**		**Eastern Bit.**	
Slagging Index Definition				
Bituminous Ash: $R_S = B/A*S$				
Lignitic Ash: $R^*_S = [Max (H.T.)+4*Min(I.T.)]/5$				
Slagging Index & Potential	**Index**	**Potential**	**Index**	**Potential**
Bituminous Ash: $R_S = B/A*S$	**1.18**		**1.88**	
$R_S < 0.6$ = low				
$0.6<R_s<2.0$ = medium				
$2.0<R_s<2.6$ = high				
$R_s>2.6$ = severe				
Lignitic Ash: $R^*_S = [Max (H.T.)+4*Min(I.T.)]/5$	**---**	**Medium**	**---**	**Medium**
$R_s^* > 2450$ = low				
$2250<R_s^*< 2450$ = medium				
$2150<R_s^*< 2250$ = high				
$R_s^*< 2100$ = severe				
Fouling Index Definition				
Bituminous Ash: $R_F = B/A*Na_2O$				
Lignitic Ash, Na_2O, when $CaO+MgO+Fe_2O_3 >20\%$ (w/w)	**TRUE**		**TRUE**	
Lignitic Ash, Na_2O, when $CaO+MgO+Fe_2O_3 <20\%$ (w/w)	**---**		**---**	
Fouling Index & Potential	**Index**	**Potential**	**Index**	**Potential**
Bituminous Ash: $R_F = B/A*Na_2O$	**0.36**		**0.26**	
$R_F < 0.2$ = low				
$0.2 < R_F < 0.5$= medium		**Medium**		**Medium**
$0.5 < R_F < 1.0$= high				
$R_F>1.0$ = severe				
Lignitic Ash, Na_2O, when $CaO+MgO+Fe_2O_3 >20\%$ (w/w)	**0.9**		**0.6**	
$Na_2O < 3$ = low to medium		**---**		**---**
$3 <Na_2O < 6$ = high				
$Na_2O > 6$ = severe				
Lignitic Ash, Na_2O, when $CaO+MgO+Fe_2O_3 <20\%$ (w/w)	**0.9**		**0.6**	
$Na_2O < 1.2$ = low to medium		**---**		**---**
$1.2 <Na_2O < 3$ = high				
$Na_2O > 3$ = severe				

Ash Classification: $Fe_2O_3 > CaO+MgO$ ==> Bituminous ash; $Fe_2O_3 < CaO+MgO$ ==> Lignitic ash

Table 2. 3: Proximate and Ultimate Analyses of Size Graded Coal Samples

QUANTITY	CONESVILLE #5 COAL (400x400 Mesh)		PITTSBURGH #8 COAL (200x400 Mesh)	
	As Rec'd	daf	As Rec'd	daf
Moisture	3.6		1.1	
Volatile Matter	36	41.6	38.2	44.2
Fixed Carbon (Diff.)	50.5	58.4	48.1	55.8
Ash	9.9		12.7	
Hydrogen	4.5	5.2	4.4	5.1
Carbon	69.1	79.9	70.6	81.8
Sulfur	3.1	3.6	5.2	6.1
Nitrogen	1.4	1.6	1.3	1.5
Oxygen (Diff.)	8.4	9.7	4.8	5.5

2.3 Combustion Evaluation of Size-Graded Coal Samples

The coal samples, whose analyses are given in Table 2.3, were combustion tested in both a Thermo-Gravimetric Analysis (TGA) apparatus and a Drop Tube Furnace System-1 (DTFS-1). The equipment, experimental procedures, and conclusions are presented in the succeeding sections.

2.3.1 Thermo-Gravimetric Analysis (TGA)

The TGA apparatus, experimental procedure, and results are presented below.

2.3.1.1 TGA Description and Procedure

ALSTOM uses a Perkin-Elmer TGA Model 7 (Figure 2.1) to obtain specific reactivity parameters of fuels and limestones, among other things. The TGA is also used to derive the “micro-proximate” analyses of coals and coal chars, based on a test protocol developed in-house. Testing was conducted as follows. About 4-6 mg of sample was placed in the TGA sample pan. The sample amount was such that it distributed over the bottom of the pan almost as a monolayer, thus minimizing O_2 mass transfer control phenomenon during combustion. Equal amounts of nitrogen (which serves to protect the balance) and air is allowed to pass through the reaction tube containing the coal sample. The temperature control mechanism is actuated, such that the heating rate is maintained at 10 ^{o}C/min from room temperature to the completion of combustion. Both weight loss and rate of rate of weight loss are monitored and recorded throughout the combustion process. The information from this testing is used, as shown in the next section, to derive combustion efficiency curves, and “pseudo” surface reaction kinetic parameters of the test coals as a function of reaction medium.

Three cases were evaluated for each coal. Fifty cc/min of N_2 gas were admitted into the balance compartment to protect it from over-heating. The Base case consisted of burning the coal in 50 cc/min of air. The Constant Mass case consisted of burning the coal in an O_2/CO_2 mixture, such that the mass of O_2 used in the Base case was maintained and the mass of N_2 used in the Base case was replaced with an equal mass of CO_2. The Constant

Volume case consisted of burning the coal in O_2/CO_2 mixture, such that the mass of O_2 used in the Base case was again maintained and the volume of N_2 used in the Base case was replaced with an equal volume of CO_2.

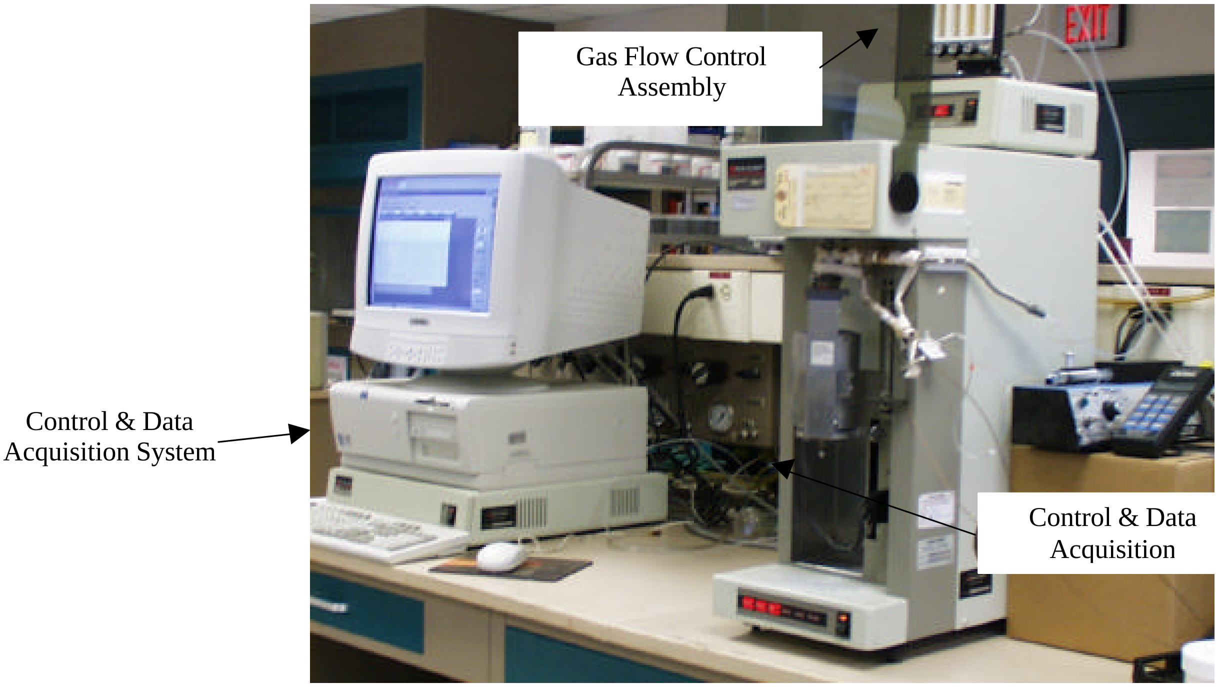


Figure 2. 1: Thermo-Gravimetric Analyzer (TGA)

Table 2. 4: Thermo-Gravimetric Analysis Test Matrix

Combustion		Balance		Combustion Air										Comb.Med. Comp.		
		Cooling N_2		Total Air		O_2		N_2		CO_2		Grand Total		(Vol. %)		
Medium	Case	(cc/min)	(g/min)	(cc/min)	(g/min)	(cc/min)	(g/min)	(cc/min)	(g/min)	(cc/min)	(g/min)	(cc/min)	(g/min)	N_2	CO_2	O_2
Air	Baseline	50	0.0573	50	0.0592	10.5	0.01373	39.5	0.04524	0.0	0.0	100.0	0.1162	89.5	---	10.5
O_2/CO_2 Mixture	Constant Mass	50	0.0573	0.0	0.0	10.5	0.01373	0.0	0.0	25.2	0.0452	85.7	0.1162	58.4	29.4	12.3
O_2/CO_2 Mixture	Constant Volume	50	0.0573	0.0	0.0	10.5	0.01373	0.0	0.0	39.5	0.071	100.0	0.1420	50.0	39.5	10.5

All flow rates @ 25 °C & 1 atm.

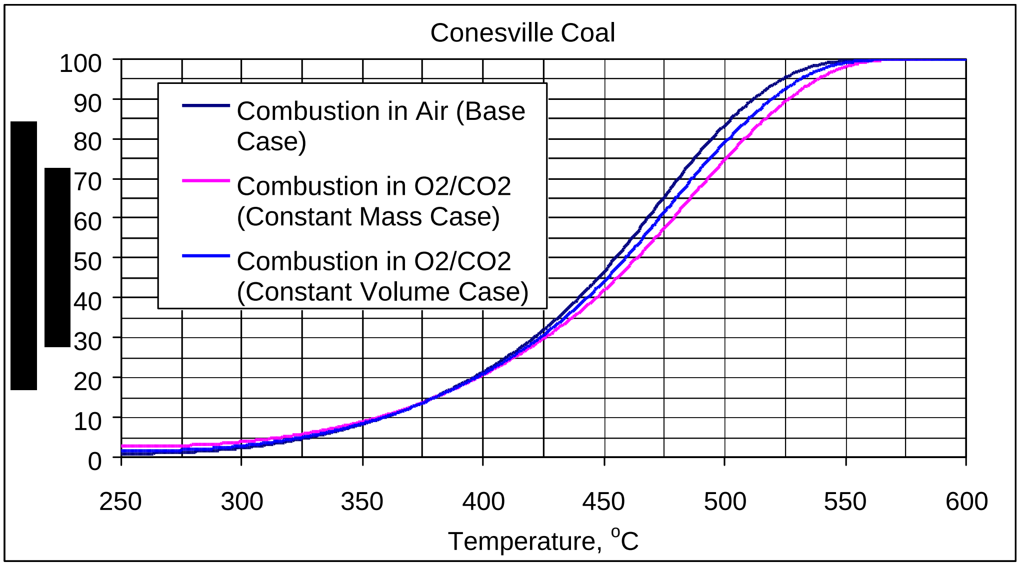
2.3.3.2 TGA Results

As stated in Section 2.2, the TGA evaluation was carried out on the 200x400-mesh samples identified in Table 2.3. Results from this evaluation are presented in Figures 2.2 to 2.4.

The combustion efficiency profiles, shown in Figure 2.2, indicate that some subtle differences seem to exist between coal firing in air and O_2/CO_2 mediums. A close examination of the data, however, shows that these differences are within experimental error for this type of testing. The results further indicate that both coals have similar burning characteristics, as they both go to completion within the same temperature range (550-575 ^oC).

The differential thermo-gravimetric analysis profiles, derived from the first derivative of weight loss with time (i.e., -dW/dt) are shown in Figure 2.3. One of the parameters traditionally used to interpret these curves is the Tkp. This parameter is, by definition, the temperature at which maximum heat evolution or reaction rate occurs. The values for the Conesville Unit #5 coal range from 472 to 479 ^oC; the corresponding values for Pittsburgh #8 coal fall in the 462 to 479 ^oC range. Clearly, these numbers are in such a narrow range that it can be concluded that: (1) burning either of the coals in air or any of the O_2/CO_2 mixture ratios used will produce virtually identical results; and (2) both coals have similar reactivity characteristics. These results are consistent with those reported in Figure 2.2.

The combustion efficiency curves in Figure 2.2 were used to derive first-order reaction kinetic parameters, assuming that the effects of O_2 mass transfer in the fixed-bed TGA were minimal and could be neglected. Results are plotted in Figure 2.4 as logKs versus 1000/, where Ks is the “pseudo” surface reactivity rate parameter (sec^{-1}), and T (^oK) is the reaction temperature. These results are consistent with those reported in Figures 2.2 and 2.3.



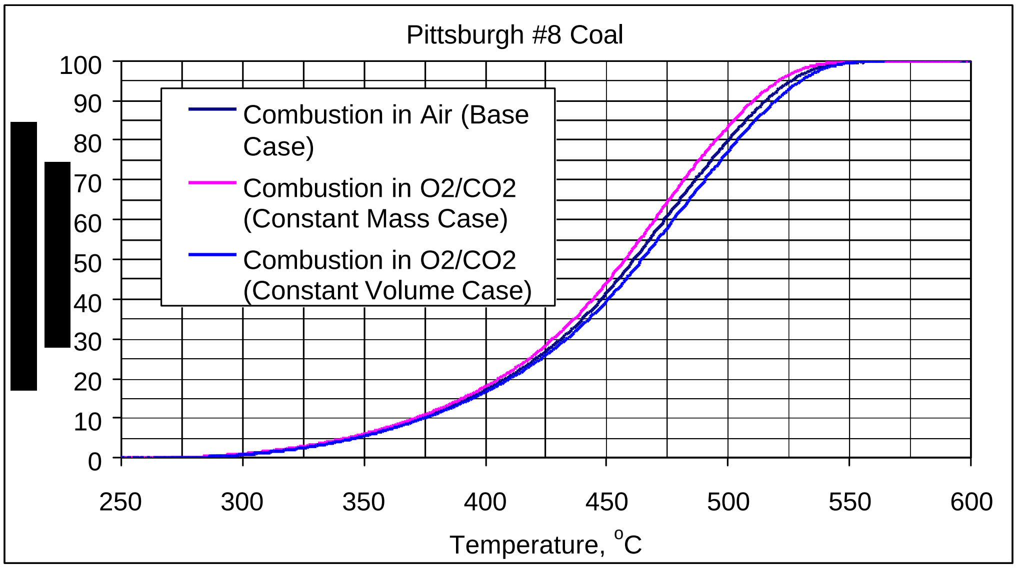


Figure 2. 2: TGA Coal Combustion Efficiency Profiles

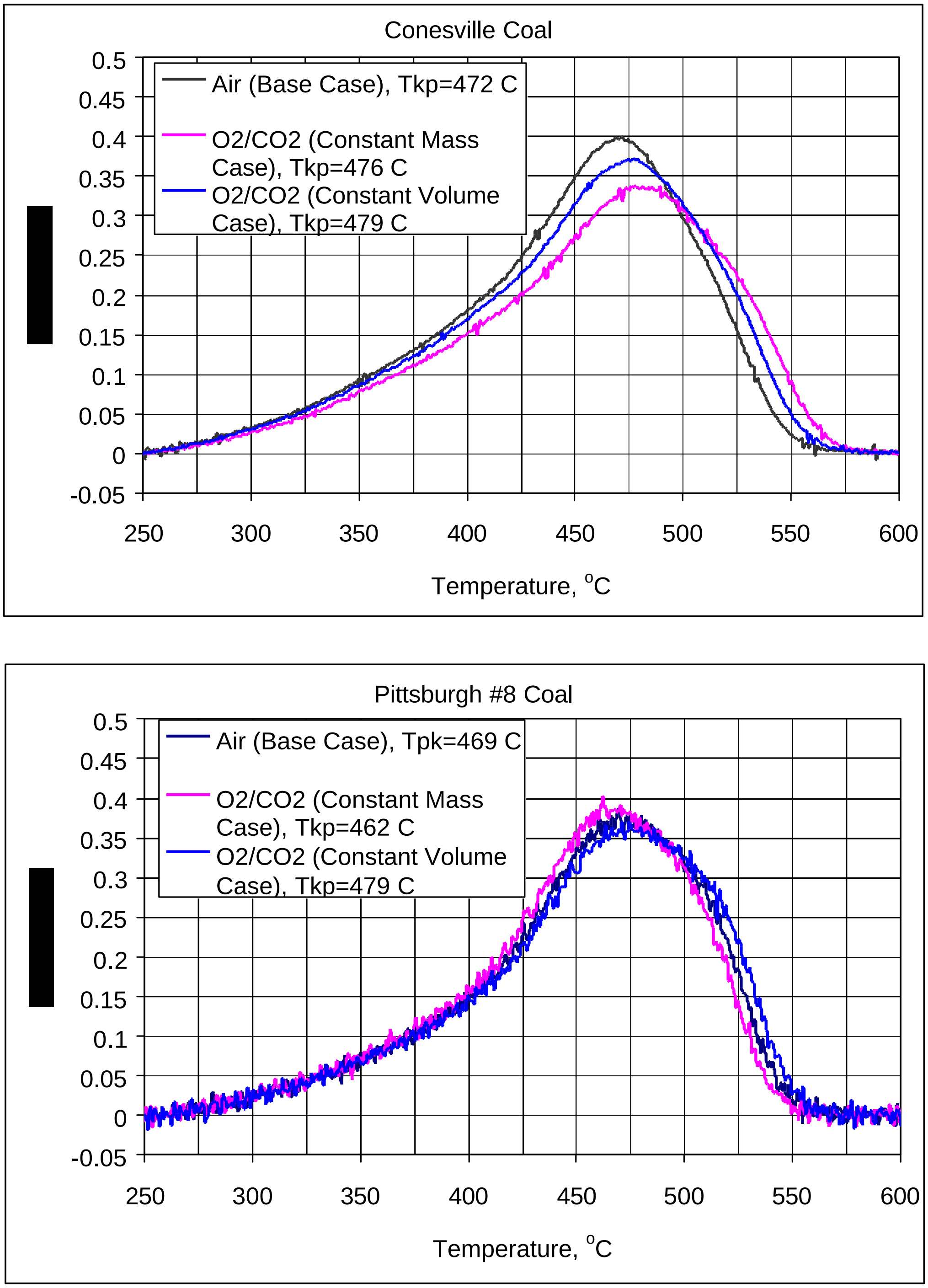


Figure 2. 3: Differential Thermo-Gravimetric Analysis Profiles

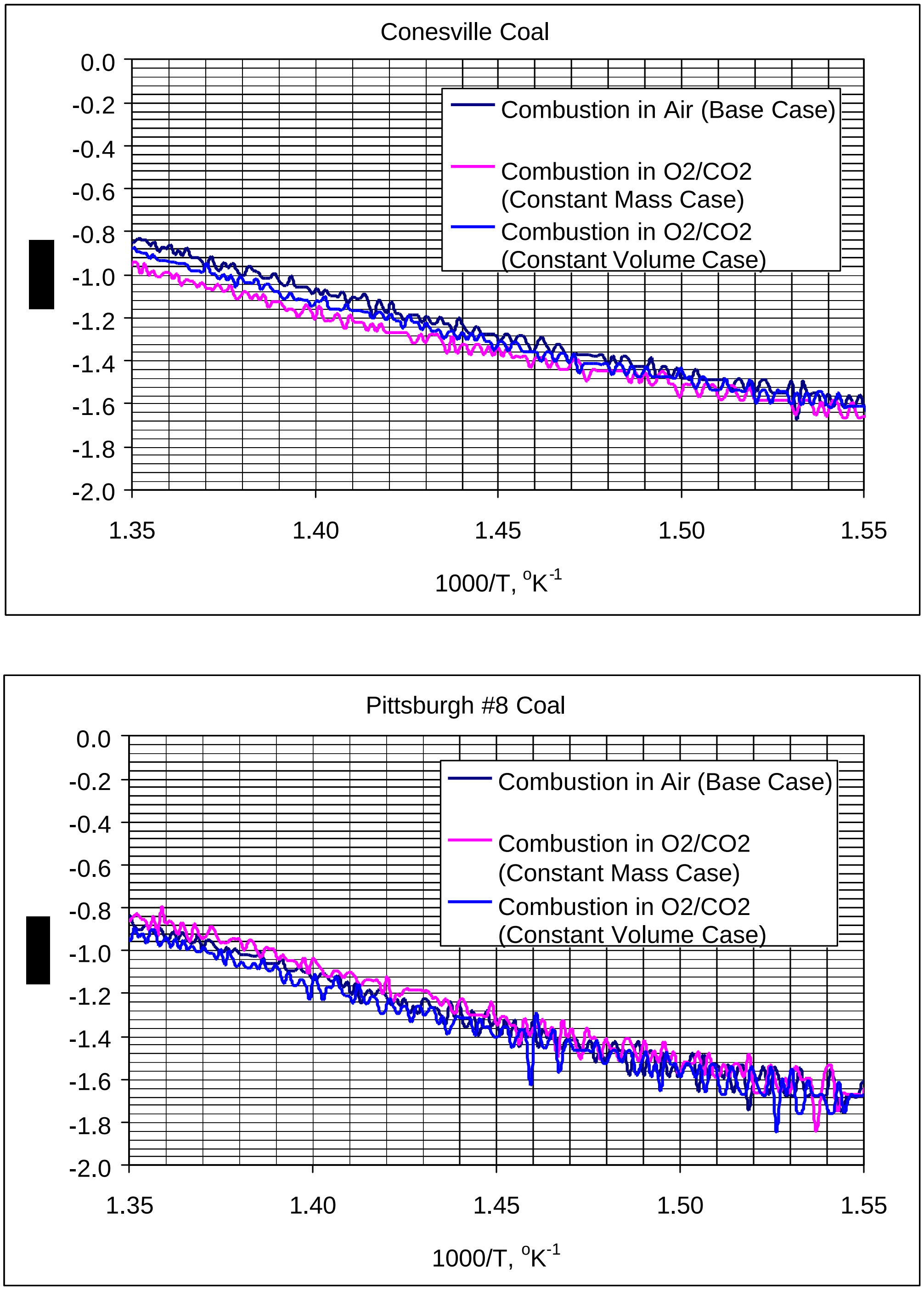


Figure 2. 4: TGA Reaction Rates of Coal Samples

2.3.2 Drop Tube Furnace System-1 (DTFS-1) Testing

The test facility, testing procedures used, and results obtained therefrom are discussed in the following sections.

2.3.2.1 DTFS-1 Description and Testing Procedure

ALSTOM's Drop Tube Furnace System-1 (DTFS-1) (Figure 2.5) is comprised of a 1-inch inner diameter, horizontal tube, gas pre-heater and a 2-inch inner diameter, vertical tube, test furnace for providing controlled temperature conditions to study devolatilization, gasification and/or combustion phenomena. This entrained flow reactor, which is electrically heated with silicon carbide elements, is capable of heating reacting particles to temperatures of up to 2700 °F and sustaining particle residence times of up to about one second to simulate the suspension firing conditions encountered in pulverized coal-fired boilers.

The DTFS-1 testing procedure entails the following: (1) the fuel is fed at a precisely known rate through a water-cooled injector into the test furnace reaction zone; (2) the fuel and its carrier gas are allowed to rapidly mix with a pre-heated down-flowing secondary gas stream; (3) devolatilization, gasification, or combustion is allowed to occur for a specific time (dictated by the transit distance); (4) reactions are rapidly quenched by aspirating the mixture into a water-cooled sampling probe; (5) the solids are separated from gaseous products in a filter medium; and (6) an aliquot of the effluent gas stream is sent to a dedicated Gas Analysis System (GAS) for on-line determination of NOx, SO_2, O_2, CO_2, CO, and THC (total hydrocarbons) concentrations. Minor gaseous components (HCN, NH_I, CH_I, VOCs, PAHs, etc.) can be measured, if required, with existing ancillary equipment (a gas chromatograph and / or a Fourier transform infrared spectrometer). A LabView-based data acquisition system records, on demand, all relevant test data for subsequent retrieval and processing. The LabView software package is also used to control the gas flow rates into the furnace.

An ash tracer technique (Nsakala, et al., 1977; Nsakala, et. Al., 1986) is used in conjunction with the proximate analyses of feed samples and chars subsequently generated in the DTFS-1 to calculate the devolatilization, gasification or combustion efficiency as a function of operational parameters (particle temperature, particle residence time, fuel fineness, reaction medium, etc.). A proprietary software package can, alternatively, use the information on concentrations of CO_2, CO, and THC (if available) in the effluent gas streams to calculate carbon conversion rates under prevailing conditions.

The DTFS-1 testing is normally used to derive quantitative information on the impact of various operating parameters such as particle size, stoichiometry and temperature/time history on unburned carbon loss and gaseous (NOx, CO and SO_2) emissions.

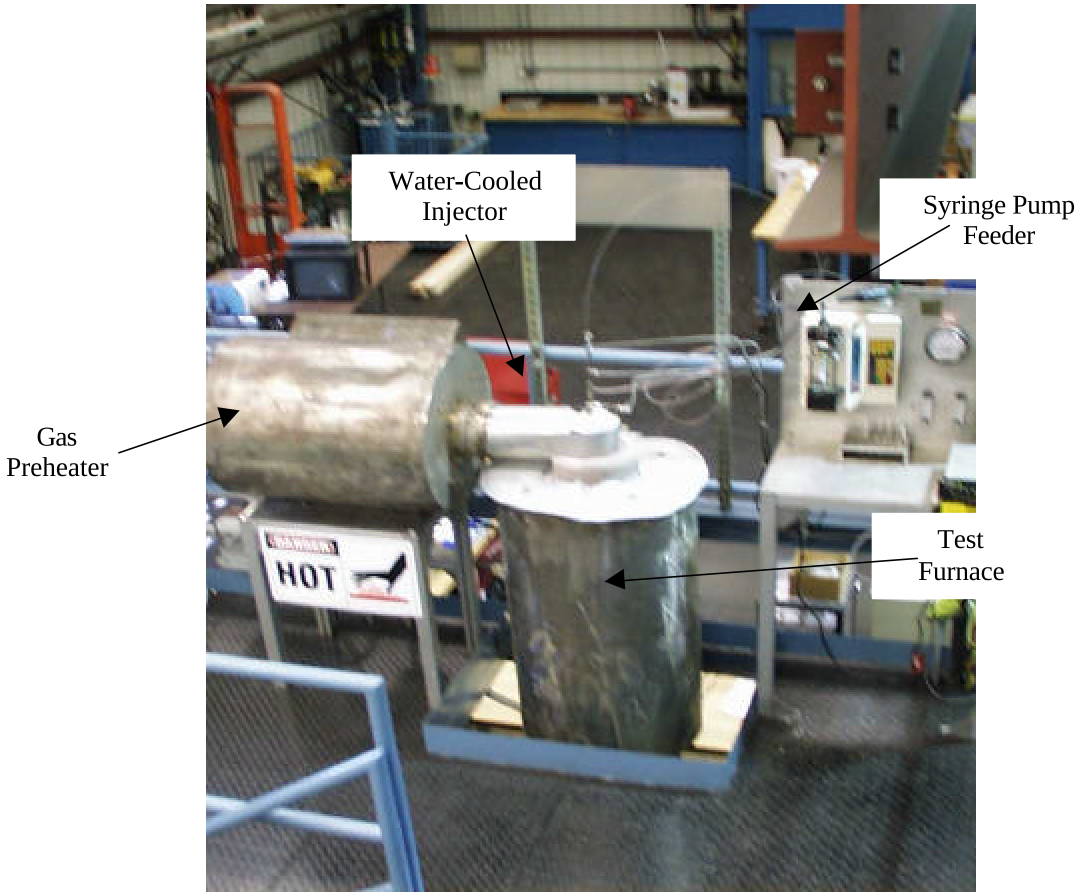


Figure 2. 5: ALSTOM Power's Drop-Tube Furnace System-1 (DTFS-1)

Three cases were evaluated for each coal (Table 2.5). The Base case consisted of burning the coal in 15 l/min of air. The Constant Mass case consisted of burning the coal in an O_2/CO_2 mixture, such that the mass of O_2 used in the Base case was maintained and the mass of N_2 used in the Base case was replaced with an equal mass of CO_2. The Constant Volume case consisted of burning the coal in an O_2/CO_2 mixture, such that the mass of O_2 used in the Base case was again maintained and the volume of N_2 used in the Base case was replaced with an equal volume of CO_2.

The rationale for selecting these three test cases is as follows. The Base case represents coal combustion in air, consistent with a "business as usual" scenario. The Constant Mass case represents a realistic scenario if coal were to be fired in an O_2/recycled flue gas environment. The Constant Volume case also represents coal combustion in an O_2/recycled flue gas environment, although in this case the CO_2 volume is conserved vis-à-vis the N_2 it would displace from the air. Table 2.5 shows that the total gas volumetric flow rates for the Constant Mass and Constant Volume cases are, respectively, lower than and equal to that of the Base case (10.7 l/min and 15.0 l/min versus 15.0 l/min, at 25 ^{o}C and 1 atm.). Hence, all other things being equal, the gas residence times in the furnace for the Constant Mass and Constant Volume cases would be, respectively, longer than and equal to the gas residence time for the Base case.

The coal feed rate was adjusted for each coal such that the overall stoichiometry (λ) was ~2.50. This value is more than a factor of 2 higher than the typical value encountered in pulverized coal firing applications. It was known, from previous experience, that in order to achieve reasonably low unburned carbon emissions in the drop tube, it was necessary to conduct these experiments in such oxygen-rich environments. This mode of operation is necessitated for two principal reasons, namely: (1) the DTFS-1 operates in a laminar flow regime,

with a Reynolds number in the 150-225 range; and (2) the maximum residence time for the conditions evaluated was about 0.8 seconds.

Table 2. 5: DTFS-1 Test Matrix

Combustion Medium	Case	Combustion Air						CO_2		Grand Total		Overall Stoichiometry (l)	Comb.Med. Comp. (Vol. %)		
		Total Air		O_2		N_2									
		(l/min)	(g/min)	(l/min)	(g/min)	(l/min)	(g/min)	(l/min)	(g/min)	(l/min)	(g/min)		N_2	CO_2	O_2
Air	Baseline	15.00	17.76	3.15	4.12	11.85	13.57	0.00	0.00	15.00	17.69	2.5	79.0	---	21.0
O_2/CO_2 Mixture	Constant Mass	0.00	0.00	3.15	4.12	0.00	0.00	7.55	13.57	10.70	17.69	2.5	---	70.5	29.5
O_2/CO_2 Mixture	Constant Volume	0.00	0.00	3.15	4.12	0.00	0.00	11.85	21.31	15.00	25.43	2.5	---	79.0	21.0

All flow rates @ 25 °C & 1 atm.

2.3.2.2 Reactant Gas Temperature Measurements

A suction pyrometer was used to measure, for each of the three study cases, the gas temperature profile along the axis of the DTFS-1 reaction chamber. In addition to the in-furnace measurements, heat transfer calculations were performed for each case to predict the impact of the reactant gas characteristics on its temperature profile along the axis of the DTFS-1 reaction chamber. The predictions were based on the exact DTFS-1 operating conditions for the same three study cases. The measurements (Figure 2.6) indicate the following gas temperature profile trend: $T_{\text{Base case}} < T_{\text{Constant Volume case}} < T_{\text{Constant Mass case}}$. The reason for this occurrence is that CO_2, unlike N_2, is capable of absorbing energy radiated from the furnace walls. CO_2 has a greater specific heat than N_2 (Figure 2.7), but as is evident from the Constant Mass case, the ability of the CO_2 to absorb/emit radiation dominates and causes the CO_2-rich gas to achieve higher temperatures than the N_2-rich case. The trends exhibited by the theoretical heat transfer model, developed to predict gas temperatures, supported those found experimentally (Figure 2.6).

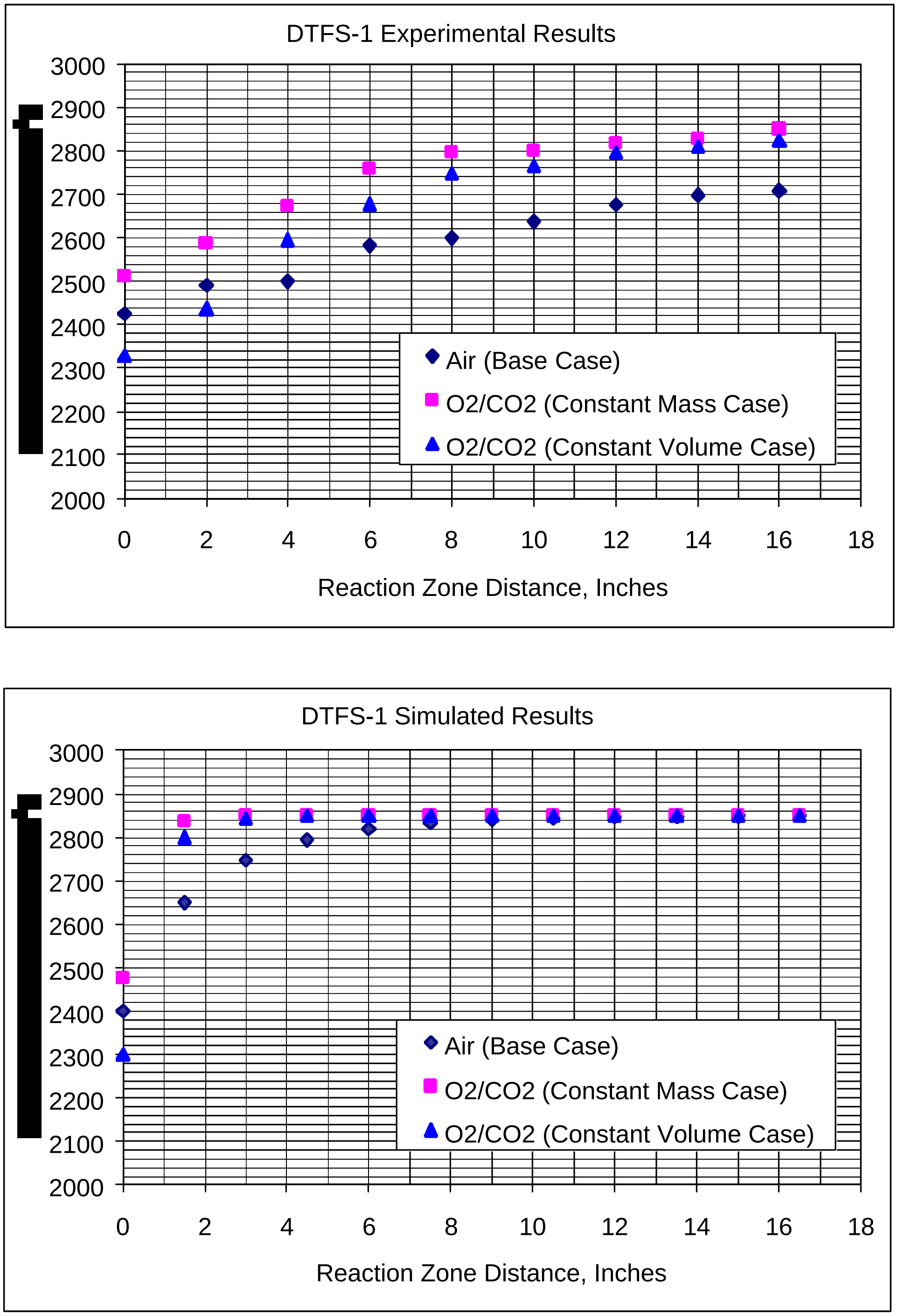


Figure 2. 6: Measured and Predicted Gas Temperatures in the DTFS-1

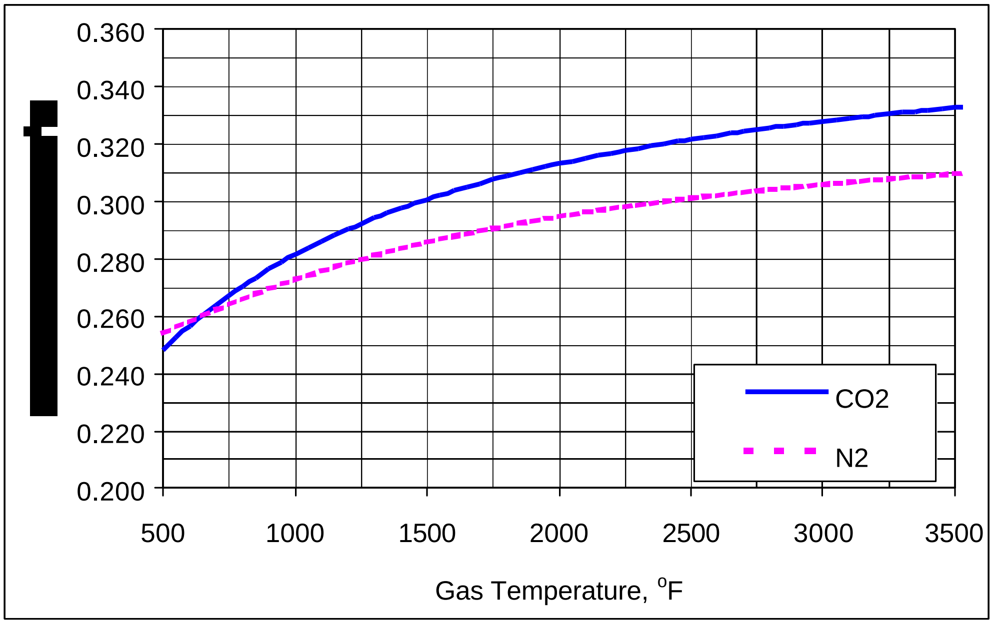


Figure 2. 7: Specific Heats of Nitrogen and Carbon Dioxide as a Function of Temperature

It should be borne in mind that the heat transfer calculations were made only to validate the trends of the gas temperatures measured in the DTFS-1. No attempts were made to calibrate those calculations to the measurements.

When gas mass flow rates, temperatures, and furnace geometry are taken into consideration together the corresponding gas residence times (t) indicated the following trend: $t_{\textbf{Constant Volume case}} < t_{\textbf{Base case}} < t_{\textbf{Constant Mass case}}$**. The impacts of these trends or lack thereof, on unburned carbon and gaseous emissions, will be discussed in Section 2.3.2.3.**

2.3.2.3 Unburned Carbon and Gaseous Emission Results

It is important to reiterate at the outset (See Section 2.3.2.1) that: (1) the DTFS-1 is operated in a laminar flow regime, with a Reynolds number in the 150-225 range; and (2) the maximum residence time for the conditions evaluated was about 0.8 sec. Hence, in order to achieve reasonably low unburned carbon emissions at the furnace outlet, the DTFS-1 was operated at a rather high overall oxygen/fuel stoichiometric ratio (i.e., l ~2.5 versus the typical value of ~1.2 encountered in pulverized coal firing applications). Consequently, the unburned carbon and gaseous emission results reported here should be interpreted strictly on a relative basis only, as no attempt was made to correlate them with field (i.e., Conesville Unit #5) performance results.

Table 2.6 summarizes the unburned carbon (UBC) results obtained from firing the coal samples from Conesville Unit #5 and Pittsburgh #8 seam under the three study cases discussed above. Figures 2.8 and 2.9 are plots of coal combustion efficiency and unburned carbon in fly ash as a function of gas residence time.

Table 2. 6: Summary of DTFS-1 Test Results on Coal Combustion Efficiencies and Unburned Carbon in Fly Ashes

Reaction Zone Length (inches)	Gas Residence Time (sec)	Combustion in Air			
		Conesville Unit #5 Coal		Pittsburgh #8 Seam Coal	
		UBC in Fly Ash (% dry)	Coal Combustion Efficiency (%daf)	UBC in Fly Ash (% dry)	Coal Combustion Efficiency (%daf)
4	0.15	64.52	78.74		
8	0.30	47.70	89.18		
12	0.44	20.48	97.20		
12	0.44	22.71	96.90		
14	0.51	11.34	98.59		
16	0.58	8.47	99.00	12.53	87.55
16	0.58	8.49	99.10		
16	0.58	9.58	98.70		

Reaction Zone Length (inches)	Gas Residence Time (sec)	Combustion in O_2/CO_2, Constant Mass Case			
		Conesville Unit #5 Coal		Pittsburgh #8 Seam Coal	
		UBC in Fly Ash (% dry)	Coal Combustion Efficiency (%daf)	UBC in Fly Ash (% dry)	Coal Combustion Efficiency (%daf)
4	0.20	51.07	85.23		
8	0.40	39.13	93.01		
12	---	---	---		
14	0.68	3.41	99.63		
16	0.78	2.27	99.75	6.16	93.56
16	0.78	1.92	99.81	7.23	92.90

Reaction Zone Length (inches)	Gas Gas Residence Time (sec)	Combustion in O_2/CO_2, Constant Volume Case			
		Conesville Unit #5 Coal		Pittsburgh #8 Seam Coal	
		UBC in Fly Ash (% dry)	Coal Combustion Efficiency (%daf)	UBC in Fly Ash (% dry)	Coal Combustion Efficiency (%daf)
4	0.15	66.61	75.32		
8	0.29	38.67	93.14		
12	0.43	16.05	98.00		
14	---	---	---		
16	0.57	1.23	99.91	16.05	84.33
16	0.57	3.00	99.67		
16	0.57	5.29	99.64		

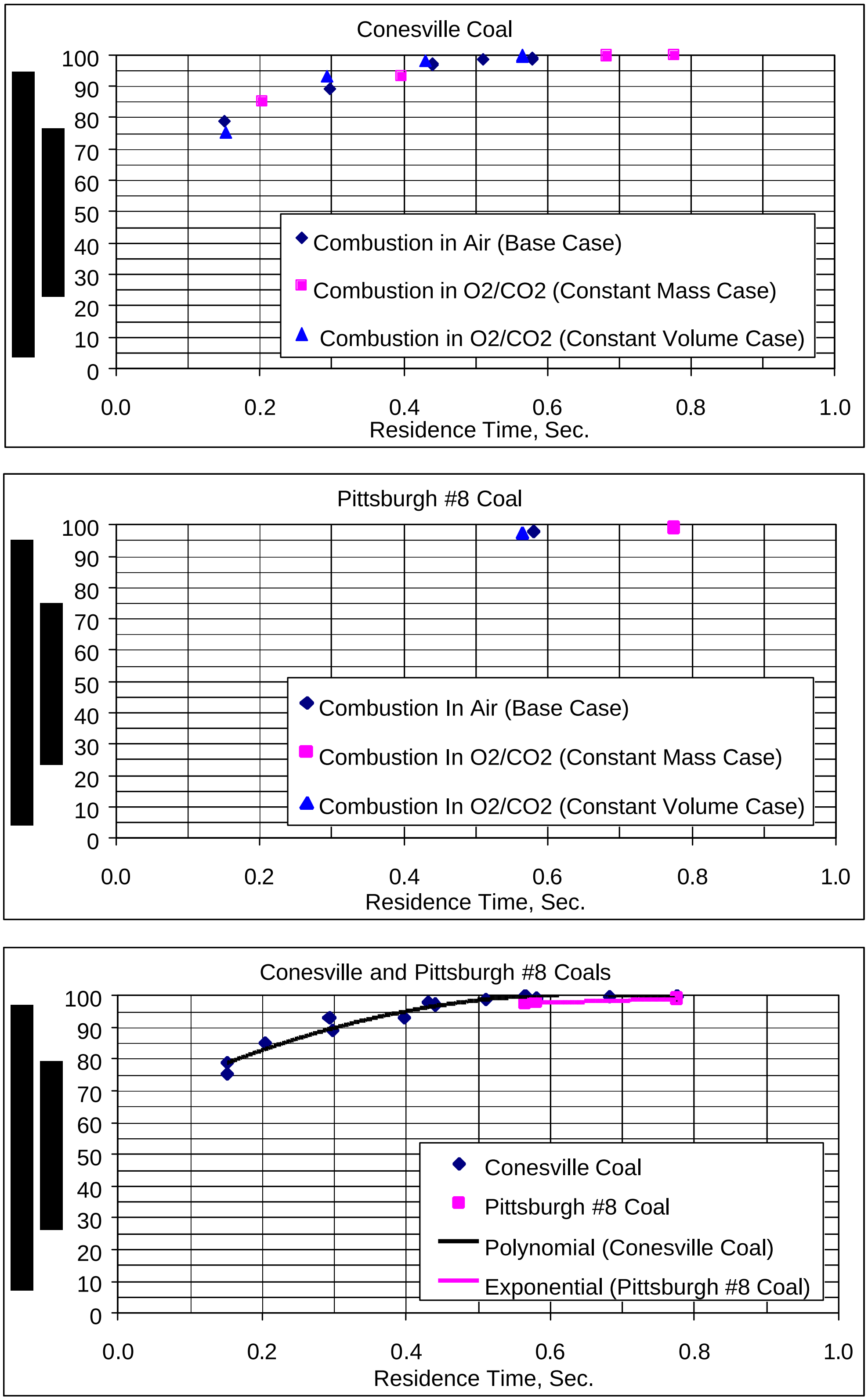


Figure 2. 8: DTFS-1 Generated Coal Combustion Efficiency Profiles

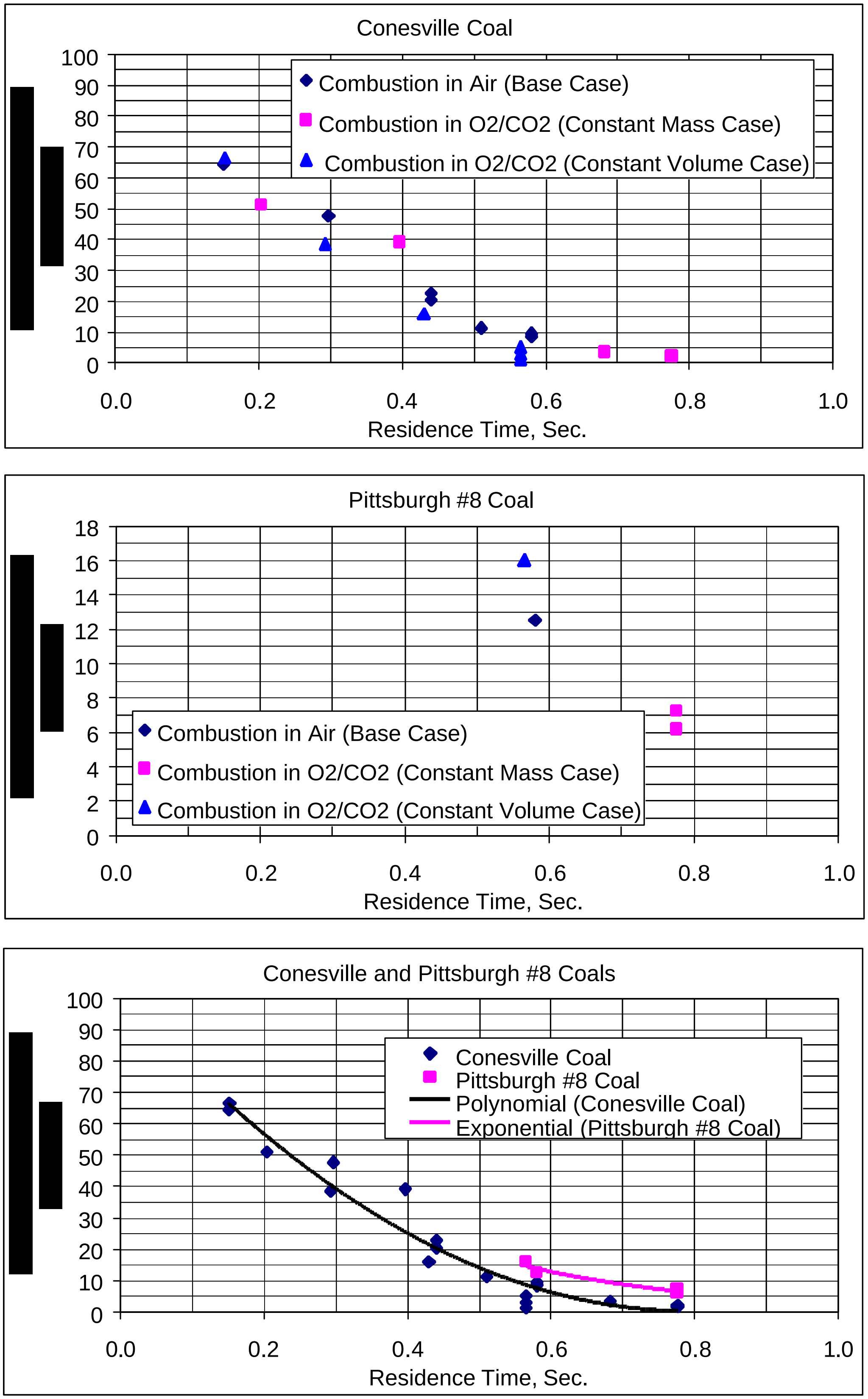


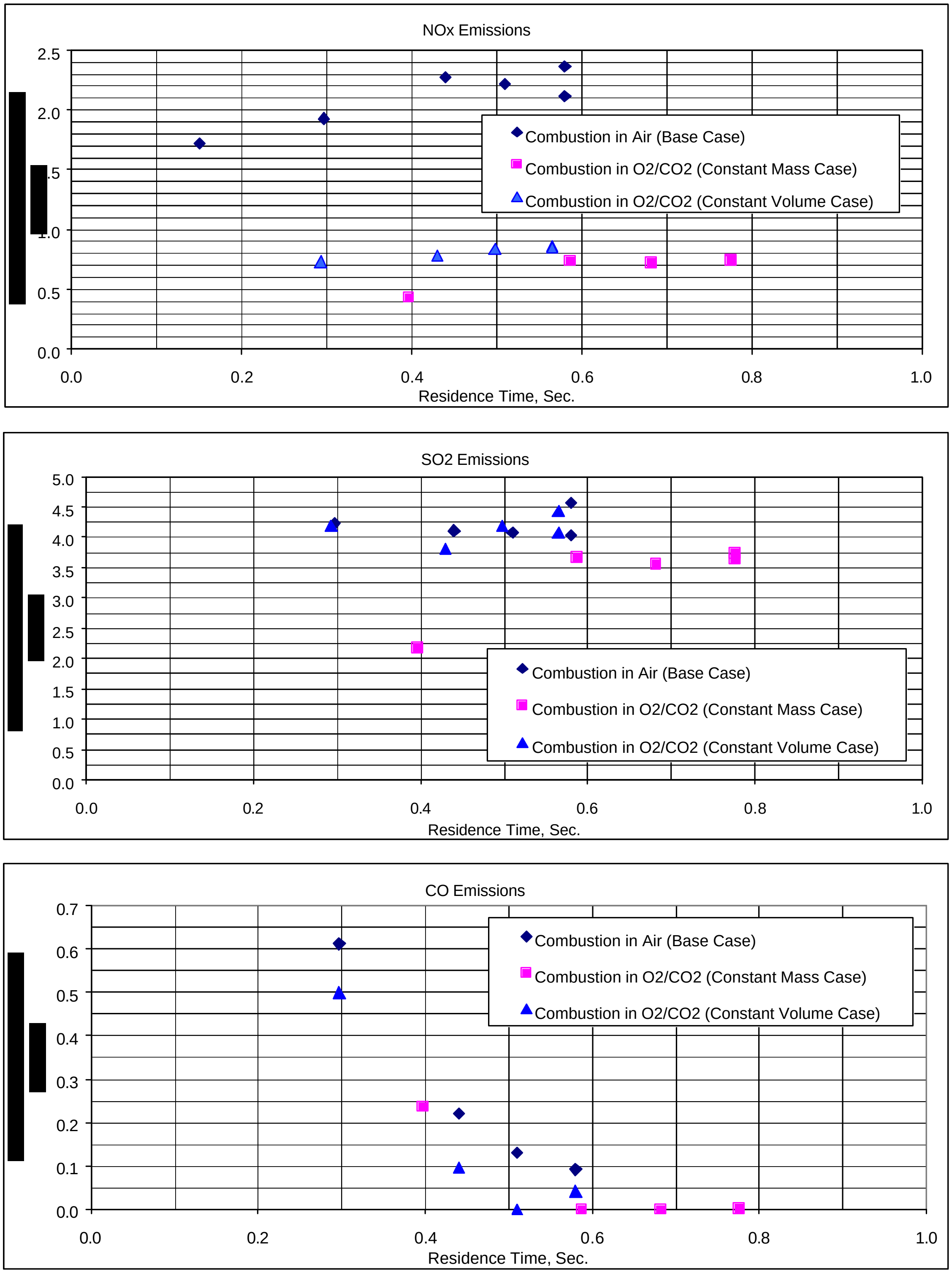
Figure 2. 9: DTFS-1 Generated Unburned Carbon in Fly Ash Samples

Figures 2.8 and 2.9 indicate that residence time has, as expected, a significant impact on the overall coal combustion efficiency, and hence, on the unburned carbon emission in the fly ash for both coals under consideration. That is, while the combustion efficiency for both coals is only about 80% (dry-ash-free coal basis) at 0.2 sec residence time, it is more than 98% at 0.8 sec. residence time. Correspondingly, unburned carbon emissions range from about 65% to less about 2% (dry fly ash basis) for the same residence times. The second point that can be made is that for the given thermal regime in the DTFS-1, the impact of reaction medium is significant. While the overall combustion efficiencies do not seem to exhibit significant differences between the baseline and Constant Mass firing, the more sensitive parameter, unburned carbon (UBC) in the fly ash, on the other hand, clearly shows better performance for the Constant Mass case. That is, the UBC at the furnace outlet was more than a factor of 4.5 lower for the Constant Mass case (~2% vs. 9%). This is believed to be due to higher reaction temperature, and yet longer residence by 35% (0.77 sec. vs. 0.57 sec.). It appears, from the shape of the UBC profiles, that, if given enough residence times, these differences would considerably narrow down. Finally, it appears that the coal sample from Conesville Unit #5 performs better than the sample from Pittsburgh #8 coal seam (e.g., 9% vs. 12.5% UBC for the air firing case). However, if given additional residence times, they would both be expected to perform similarly.

The ultimate conclusion is that unburned carbon emission for the Constant Mass case, which represents the appropriate retrofit scenario in a commercial pulverized coal firing application (i.e., the mass of CO_2 in the O_2/CO_2 combustion medium displaces the same mass of N_2 in the combustion air), provides a significant advantage over air firing. This is, presumably, because of the relatively higher temperature, and yet longer combustion process time that the Constant Mass case provides.

Table 2.2.7 and Figure 2.10 show gaseous (NOx, CO, and SO_2) emissions results obtained from burning the coal sample from Conesville Unit #5 in air and O_2/CO_2 combustion mediums. These results indicate significant increases in NOx, slight increases in SO_2, and significant decreases in CO emissions as a function of increasing residence time. This is due to, as expected, correspondingly improved overall coal combustion efficiency with time. The NOx emission reduction from air firing (Base case) to oxy-fuel firing (i.e., O_2/CO_2 combustion mediums) is about a factor of 3 (0.73 vs. 2.25 lb/MMBtu). The trend is consistent with results reported by others (Thambimuthu, et al.; 1998; Weller, et al., 1985) and by the CFD results of the present study (Section 3.0, below). However, Thambimuthu, et al. and Weller, et al. reported reductions ranging up to only about 30%. Hence, it is not clear whether the superstoichiometry utilized in the present experiment (λ ~2.5) exaggerates the difference. Sulfur dioxide for oxy-fuel firing is lower by about 19% than the corresponding value for air firing (3.6 vs. 4.3 lb/MMBtu). CO for oxy-fuel firing was lower than the corresponding values for air firing (0.09 lb/MMBtu vs. a negligible amount).

The ultimate conclusion from gaseous emission findings is that the Constant Mass case does seem to provide (1) a significant advantage over air firing, with respect to NOx and CO emissions; and (2) a slight advantage over air firing, with respect to SO_2 emission.



NOx Emissions

SO2 Emissions

CO Emissions

Figure 2. 10: DTFS-1 Generated NOx, CO and SO_2 Emissions

Table 2. 7: Summary of DTFS-1 Test Results on Gaseous emissions from Conesville Unit #5 Coal Combustion

Reaction Zone Length (inches)	Gas Residence Time (sec)	**Combustion in Air**				
		UBC in Fly Ash (% dry)	Coal Comb. Eff. (%daf)	NOx (lb/MMBtu)	SO_2 (lb/MMBtu)	CO (lb/MMBtu)
4	0.15	64.52	78.74	1.72	4.56	
8	0.30	47.70	89.18	1.93	4.22	0.61
12	0.44	20.48	97.20			0.22
12	0.44	22.71	96.90	2.27	4.11	
14	0.51	11.34	98.59	2.22	4.08	
16	0.58	8.47	99.00	2.36	4.57	0.13
16	0.58	8.49	99.10	2.11	4.04	0.09
16	0.58	9.58	98.70			

Reaction Zone Length (inches)	Gas Residence Time (sec)	**Combustion in O_2/CO_2, Constant Mass Case**				
		UBC in Fly Ash (% dry)	Coal Comb. Eff. (%daf)	NOx (lb/MMBtu)	SO_2 (lb/MMBtu)	CO (lb/MMBtu)
4	0.20	51.07	85.23			0.54
8	0.40	39.13	93.01	0.43	2.17	0.24
12	---	---	---	0.74	3.67	
14	0.68	3.41	99.63	0.72	3.56	0.00
16	0.78	2.27	99.75	0.73	3.63	0.00
16	0.78	1.92	99.81		3.74	

Reaction Zone Length (inches)	Gas Residence Time (sec)	**Combustion in O_2/CO_2, Constant Volume Case**				
		UBC in Fly Ash (% dry)	Coal Comb. Eff. (%daf)	NOx (lb/MMBtu)	SO_2 (lb/MMBtu)	CO (lb/MMBtu)
4	0.15	66.61	75.32			
8	0.29	38.67	93.14	0.73	4.19	0.50
12	0.43	16.05	98.00	0.78	3.82	0.10
14	---	---	---	0.84	4.18	0.00
16	0.57	1.23	99.91	0.85	4.08	0.04
16	0.57	3.00	99.67		4.43	
16	0.57	5.29	99.64			

2.3.3 Conclusions and Recommendations for Future Work

The conclusions and recommendations that were derived from this bench-scale testing campaign are given below.

2.3.3.1 Conclusions

A series of conclusions can be drawn from the results from combustion testing of the coal samples from Conesville Unit #5 and Pittsburgh #8 seam in both TGA and DTFS-1 equipment, namely:

- TGA results indicate that subtle differences exist between: (1) coal firing in air and in O_2/CO_2 mediums; and (2) Conesville Unit #5 coal and Pittsburgh Seam #8 coal. A close examination of the data, however, indicates that these differences are within experimental errors. Hence, it can be concluded that both coals can be expected to perform similarly under the experimental constraints specified in this study.

- DTFS-1 results showed the following:

    - Residence time has, as expected, a significant impact on the overall coal combustion efficiency, and hence, on the unburned carbon emission in the fly ash for both coals under consideration. That is, while the combustion efficiency for both coals is only about 80% (dry-ash-free coal basis) at 0.2 sec residence time, it is more than 98% at greater than 0.5 sec. residence times. Correspondingly, unburned carbon emissions range from about 65% to less about 2% (dry fly ash basis) for the same residence times.

    - The impact of reaction medium is also significant. While the overall combustion efficiencies do not seem to exhibit significant differences between the baseline and Constant Mass firing, the more sensitive parameter, unburned carbon (UBC) in the fly ash, on the other hand, clearly shows better performance for the Constant Mass case. That is, the UBC at the furnace outlet was more than a factor of 4.5 lower for the Constant Mass case (~2% vs. 9%). This is believed to be due to higher reaction temperature, and yet longer residence (0.77 sec. vs. 0.57 sec.). It appears, from the shape of the UBC profiles, that, if given enough residence times, these differences would considerably narrow down, implying that the performance in both mediums would be similar.

    - The coal sample from Conesville Unit #5 appears to perform better than the sample from Pittsburgh seam #8 coal (e.g., 9% vs. 12.5% UBC for the air firing case). However, base on the shape of the UBC profiles, it can be seen that if given additional residence times, they would both be expected to perform similarly.

    - NOx emission from oxy-fuel firing is about a factor of ~3 (0.73 vs. 2.25 lb/MMBtu) lower than that of Base case. Correspondingly, sulfur dioxide is lower by about 19% (3.6 vs. 4.3 lb/MMBtu), and CO is significantly lower (0.09 lb/MMBtu vs. a negligible amount).

    - The gaseous emission trends are consistent with results reported by others (Thambimuthu, et al.; 1998; Weller, et al., 1985) and by the CFD results of the present study (Section 3.0, below).

2.3.3.2 Recommendations for Future Work

There is, to the authors' knowledge, no fundamental coal combustion work being carried out elsewhere to determine the impact of coal nature in an oxy-fuel environment on unburned carbon and gaseous and particulate emissions. Hence, the following is recommended:

- Carry out a systematic coal fundamentals study concerning this field of endeavor. This study should be geared toward the creation of a database of information depicting the impact of coal nature (using coals of various rank coming from both domestic and international sources) when fired in an oxy-fuel environment on:

    - Coal devolatilization and char oxidation kinetics, and unburned carbon emissions in the fly ash

  - ➢ Acid rain-related gaseous (NOx, SO_2, and CO) emissions
  - ➢ Air toxics, particularly mercury (Hg), volatile organic compounds (VOCs), and particulate emissions.

- Derive furnace heat transfer data from natural gas and coal firing in a pilot-scale furnace, which simulates the temperature/time history of a boiler. Use first the data from natural gas firing to validate the CFD code, without the complication of burning particles transiting the boiler. Apply the experience learned from modeling a natural gas flame to model the coal-fired flame.

3.0 CFD Evaluation Of Conesville Boiler No. 5

3.1 Introduction

ALSTOM Power has made a strong commitment to computational fluid dynamics (CFD) and has effectively used CFD in many applications, e.g., design and optimization of coal-, oil-, and gas-fired boilers, pulverizers, air heaters, electrostatic precipitators, cyclones, and scrubbers. At any given time, there are 8 to 10 individuals performing CFD simulations on approximately 40 CPU's, including both workstations and multi-processor servers.

With respect to boiler performance, significant CFD and combustion modeling research efforts have been conducted at ALSTOM Power to characterize and optimize the performance of tangentially-fired and wall-fired furnaces. These projects have demonstrated the impact of fluid dynamics, gas phase stoichiometry, and heat input on the combustion efficiency and boiler emissions for a variety of commercial boiler applications. Synopses of recent studies that have been permitted to enter the public domain include the prediction of furnace gas temperatures and species concentrations for NOx reduction with recirculated flue gas (Richards, et al., 1997), furnace waterwall species concentrations for improvement to deposition and slagging performance (Hart, et al., 1998), and design of a low NOx burner (Haynes, et al., 1996). (Overview papers on CFD at ALSTOM Power include Marion, et al., 1995 and Fiveland, 2000.)

CFD modeling is considered to be the most cost-effective approach to assess flow and mixing related phenomena for reacting flow conditions, and was therefore proposed for this work.

3.2 Grid Preparation

American Electric Power's Conesville Unit #5 boiler has been described in detail in Volume I of this report (Bozzuto, et al., 2001); however, certain features will be reviewed below to introduce the grid and case preparation.

In general, the Conesville unit can be described as a nominal 450 MW-gross, tangentially coal fired, subcritical pressure, controlled circulation, radiant reheat unit. The furnace is a single cell design utilizing five elevations of tilting tangential coal burners. The unit fires a mid-western bituminous coal. The coal is supplied to the five elevations with five RP-903 coal pulverizers. The unit is designed to generate about 3.1 x 10^6 lb_m/hr of steam at 2400 psi and 1005 °F with reheat also to 1005 °F. The boiler is equipped with a number of superheater surfaces, including a low temperature superheater (horizontal), finishing superheater, a radiant superheat division panel section, and the superheater platen section. The reheater consists of a low temperature radiant wall section followed by a spaced finishing section. The economizer section consists of four banks of spiral finned tubes. Outlet steam temperature control is provided with de-superheating spray and burner tilt. The dimensions of the boiler unit are 51' 8" in width, 44' 3-15/16" in depth (to the rear wall), and 170' 10-1/4" in height. A sectional side view of this boiler is shown in Figure 3.1.

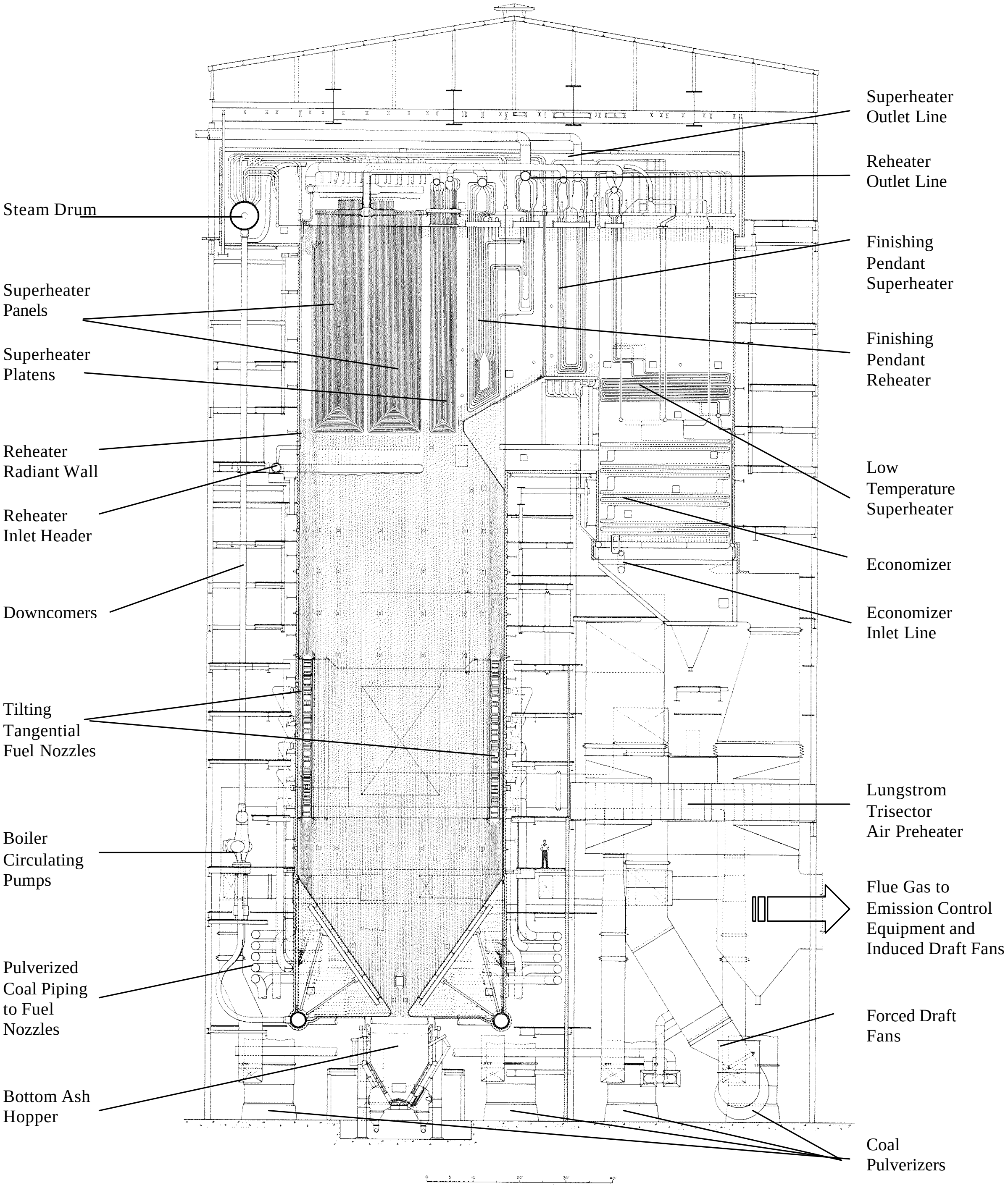


Figure 3. 1: Sectional Side View of the Conesville Unit #5 Steam Generator

ALSTOM Power uses a commercial CFD code (denoted FLUENT™) from Fluent Inc. (Lebanon, NH) for its simulations of furnaces and auxiliary equipment. Case preparation of a boiler for the Fluent code is initiated with the collection of the geometry data (from blueprints and schematics), the free areas of the nozzles, and the operating conditions. A CFD grid or mesh is prepared using a combination of EXCEL spreadsheets and a grid-generation package (denoted GAMBIT™) supplied by Fluent Inc. Skeletal and external isometric views of the computational domain are shown in Figure 3.2.

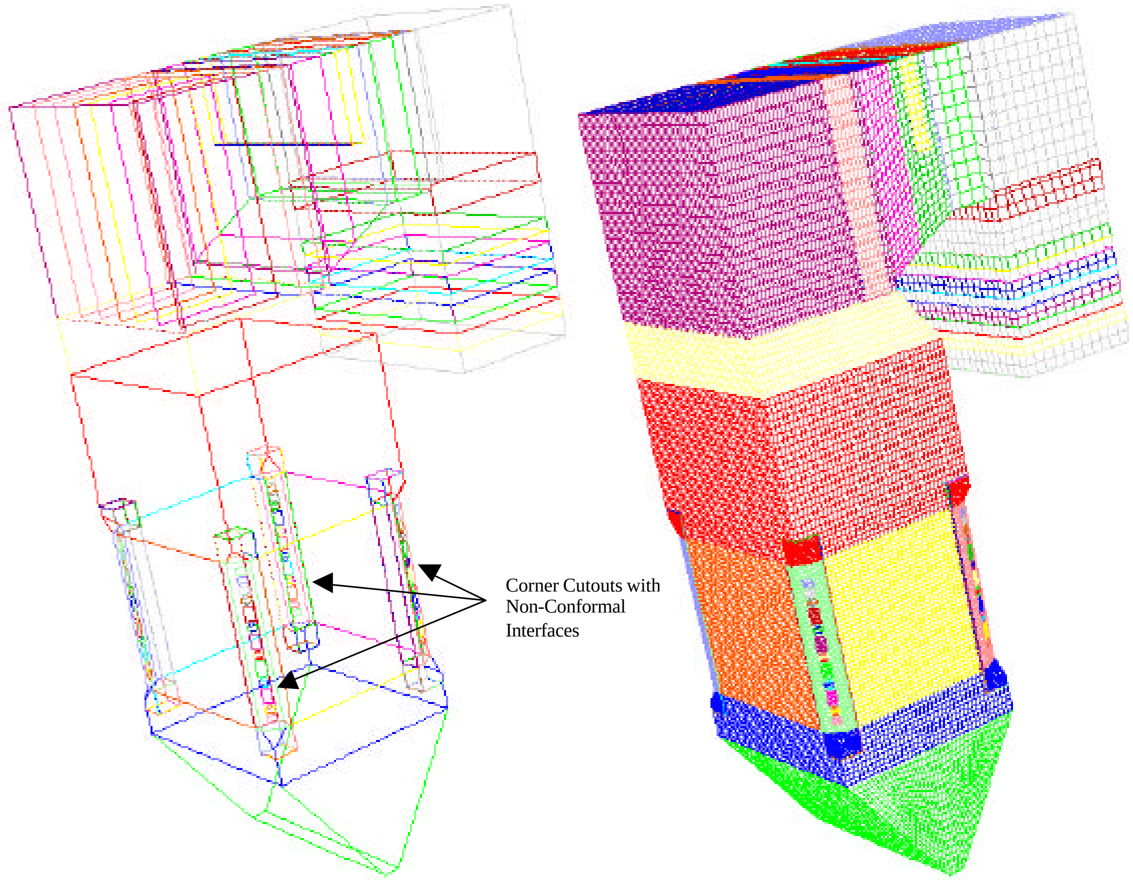

Corner Cutouts with Non-Conformal Interfaces

Figure 3. 2: Isometric Views of the Conesville Unit #5 Computational Domain

The framework or skeletal view above shows square cutouts in the corners that encompass the windboxes. These square cutouts contain cells that are smaller (by a factor of 2 in each coordinate direction) than the cells in the interior of the domain, thus allowing the windbox region to receive a greater concentration of cells than in the interior. The enhanced cell concentration in the windbox region permits greater resolution of the relatively small nozzle openings and the high gradient regions associated with discrete jets. The interfaces between each of the corner cutouts and the domain interior are referred to (by Fluent Inc.) as “non-conformal” interfaces. Such interfaces permit grids of dissimilar type and size to seamlessly interact with each other along any given plane and preserve the conservation of all relevant quantities (e.g., mass, momentum, and energy).

Close-ups of the grid from a side view of the boiler are shown in Figure 3.3. Besides the non-conformal interfaces around the windboxes, another non-conformal grid interface is utilized in the convective pass, permitting the grid to coarsen (by a factor of 2 in each coordinate direction) into the backpass. The grid in the computational domain is composed primarily of hexahedral elements. Tetrahedral elements are utilized only at the top and bottom of the windbox to assist in a geometric transition. Pyramidal elements are used at the interface between the tetrahedrons and the hexahedrons. The total number of cells in the grid is 314,176.

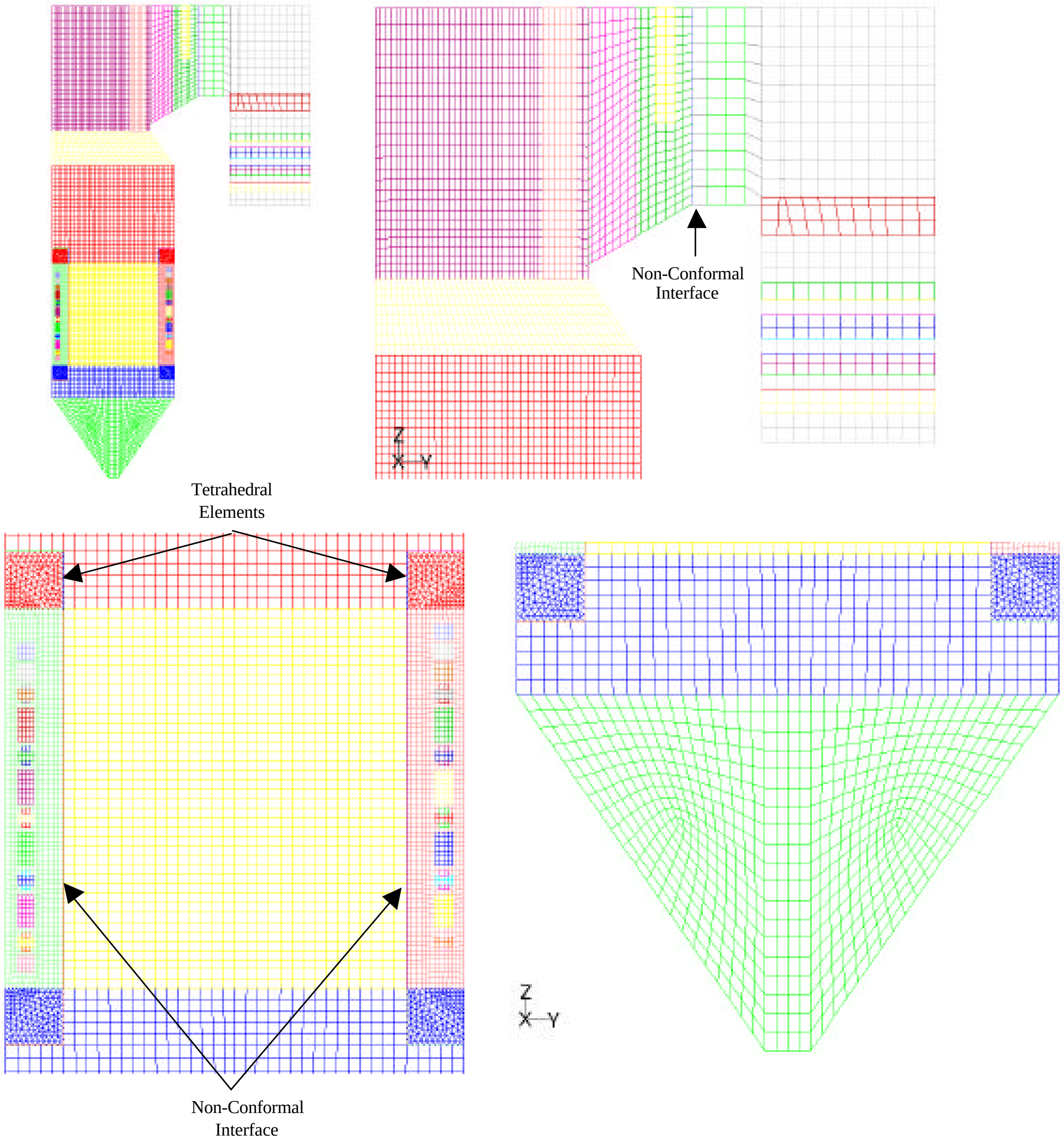


Figure 3. 3: Side Views of the Conesville Unit #5 Exterior Grid

Conesville Unit No. 5 is equipped with a tangential firing system, consisting of 3 close-coupled overfire air (CCOFA) register elevations, and 5 coal nozzle elevations. The main and CCOFA windbox widths are equal to 26 inches and the coal nozzle ID is equal to 19.25 inches. The tip/nozzle free areas were extracted from the firing system windbox selection sheets and the nozzle elevations were calculated from the available blueprints. A computational representation of the windbox and nozzles, which is necessarily crude in comparison to the details of the actual hardware, is shown in Figure 3.4. The computational inlet designations (i.e., CCOFA-C, etc.) are also shown. The small inlets on the top and bottom of each computational coal nozzle represent the flow area associated with the fuel-air stream, which in actuality, flows into the furnace around the periphery of the primary coal pipe/tip.

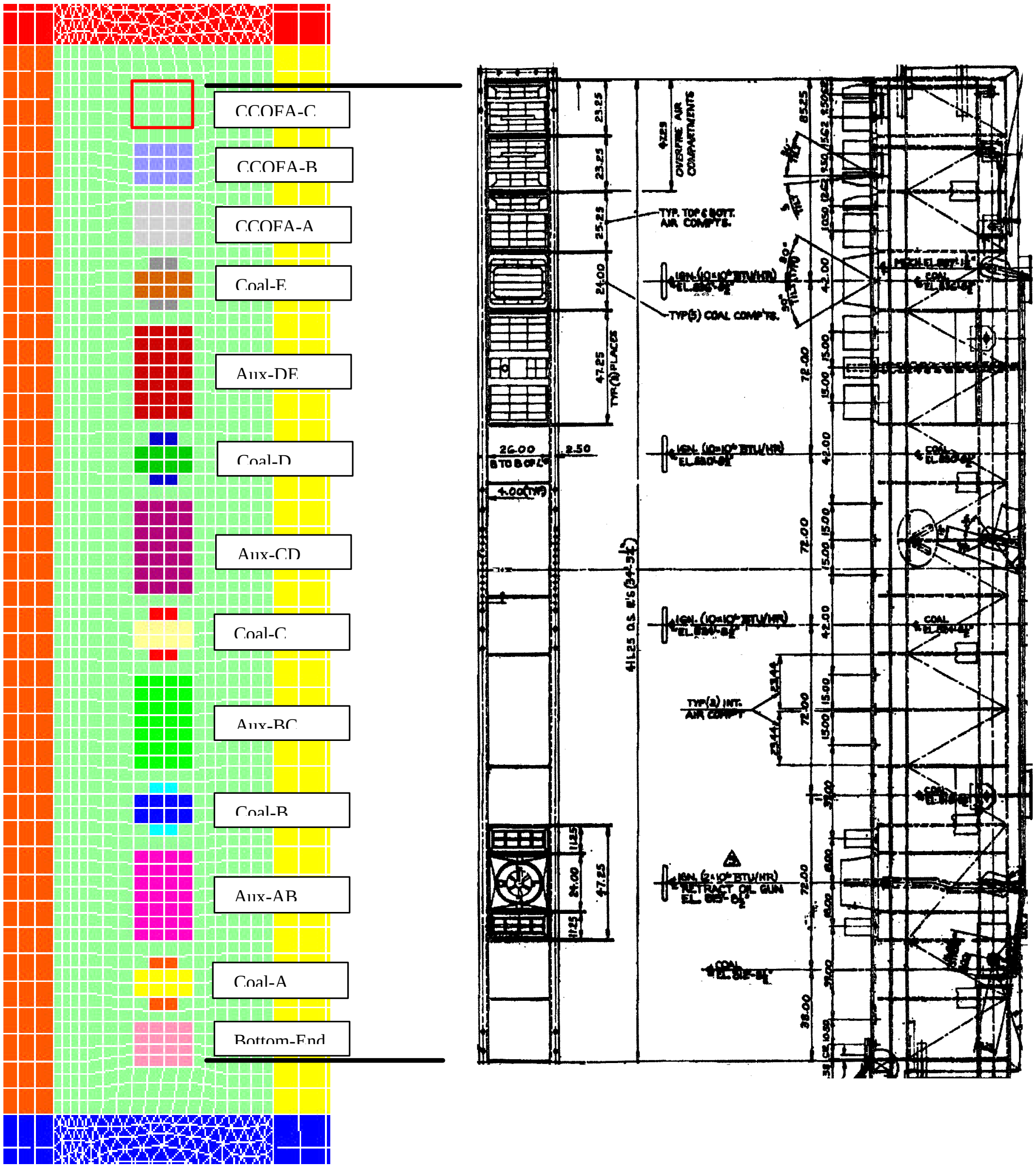


Figure 3. 4: Detailed Drawing of Windbox and Corresponding Side View of Grid Depicting Nozzle Openings

The direction of swirl rotation in Conesville Unit 5 (as viewed from the top of the boiler), is shown below in Figure 3.5. The direction of the "fireball rotation" is counter-clockwise. The normal firing angles are at 44 degrees and at 36 degrees for the pairs of opposed jets.

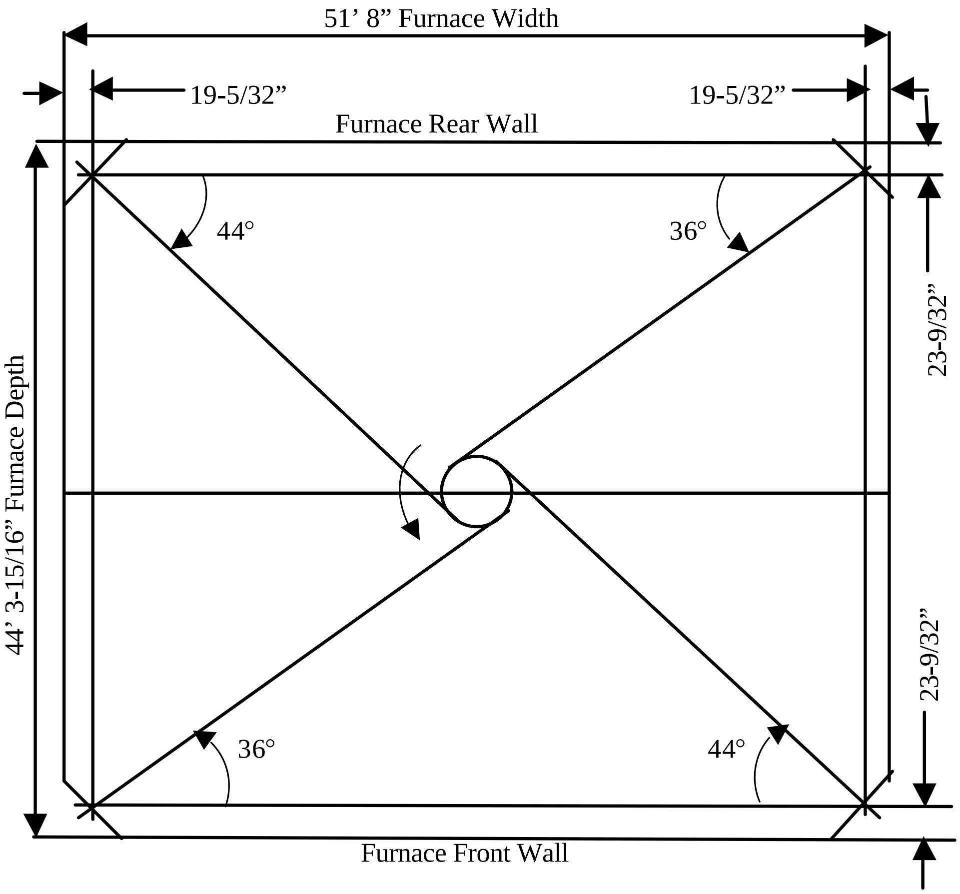


Figure 3. 5: Cross-Sectional View of Normal Firing Angles for Conesville Unit 5

The corresponding grid on the cross-section, in the vicinity of the windbox, is shown in Figure 3.6. As explained before, a 2:1 change in cell dimension occurs across the non-conformal interface in the windbox region, thus allowing a greater grid density to be concentrated in the immediate vicinity of the nozzle inlets. The grid in the windbox region consists of hexahedral cells, although they may be slightly skewed to accommodate the corner fillets (comprising the windbox).

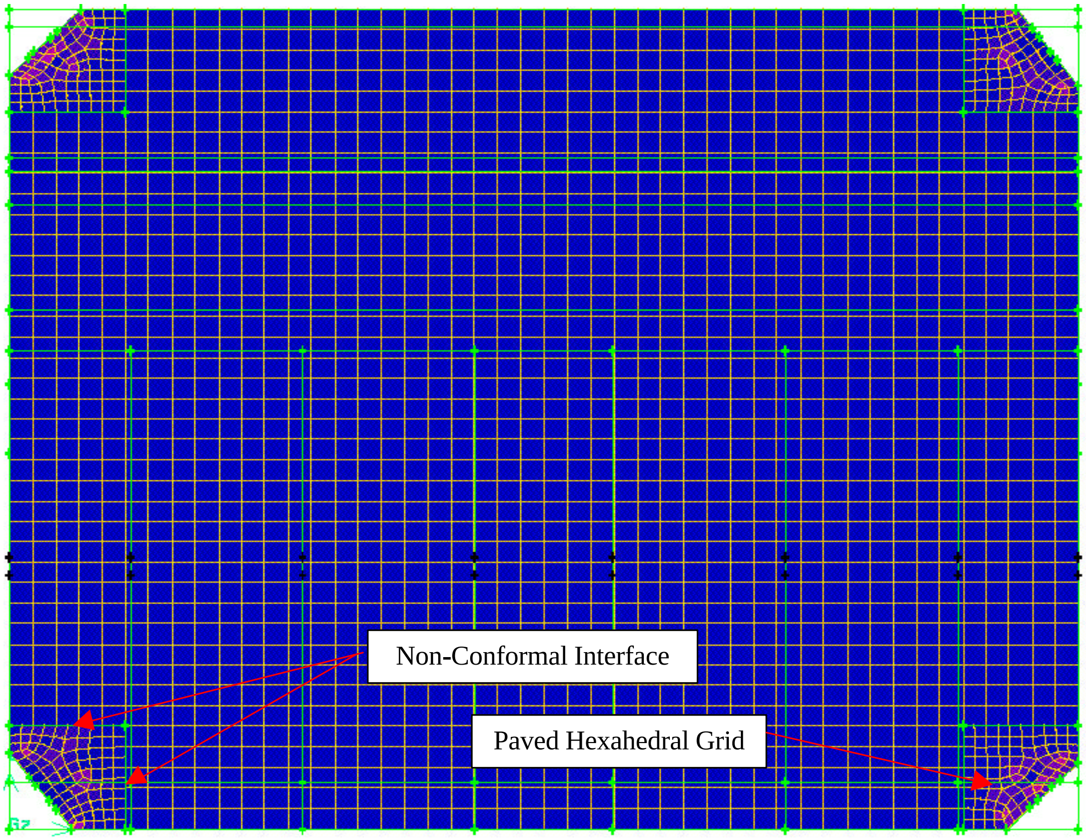


Figure 3. 6: Cross-Sectional View of Grid, in the Vicinity of the Windbox, for Conesville Unit 5

3.3 Baseline Calibration Procedure and Strategy

Typically, in an applications project, as in the present study, the first CFD case that is prepared and run is used to evaluate the furnace performance under its baseline combustion conditions. Many of the boundary conditions, such as velocity and temperature, constitute “explicit” boundary conditions and are defined as part of the input. However, a few boundary conditions (or equality constraints) are “implicitly” defined and are derived either from experimental measurements or external calculations. While “implicit” boundary conditions are not explicitly defined as inputs, it is understood that such boundary conditions must nevertheless be matched by the post-run results. Such “implicit” boundary conditions include:

- the unburned carbon (UBC) in the fly ash (i.e., carbon-in-ash value) at the outlet of the computational domain
- the outlet NOx emissions value at the outlet of the computational domain
- the “horizontal furnace outlet temperature (HFOT),” which is the mass-weighted average temperature across the plane at the apex of the arch (i.e. the horizontal furnace outlet plane or HFOP).

Both the UBC and NOx emissions values originate from experimental measurements (usually taken downstream of the air preheater). The HFOT value is calculated by an in-house, proprietary design package denoted the Reheat Boiler Program (RHBP). The Reheat Boiler Program, running in its reverse/calibration mode, is fed the excess air, windbox nozzle tilts, all known steam-side parameters, the fuel flow rate, and other general and geometric parameters, and in turn, calculates the HFOT (as well as the wall heat transfer fouling factors, the tube bank surface effectiveness factors, the gas flow rates, and the amount of superheat/reheat spray, if applicable).

In the CFD calculation, the char oxidation input parameters control the final amount of residual carbon in the fly ash, the NOx kinetic input parameters largely control the outlet NOx values, and the overall resistance to heat transfer through the waterwall and superheater surfaces and deposits controls the HFOT. An iterative trial-and-error method is used to calibrate the baseline case by changing the input parameters for char oxidation, NOx, and waterwall conduction resistance, until (after a number of runs) the code calculations match the desired UBC, the NOx value, and the HFOT, respectively. Such an iterative procedure for calibration is required since many of the input parameters and boundary conditions are not known with sufficient accuracy to be specified in an *a priori* manner, and calibration tends to compensate for the accrued uncertainty.

In the present study, the above procedure was used to calibrate the baseline case to the desired UBC (as dictated by experimental field data) and to the desired HFOT (as dictated by the RHBP results). Since the CFD code (i.e., the FLUENT™ code) did not provide sufficient access to the NOx kinetic parameters, the calculated baseline NOx results were scaled to the baseline experimental values.

3.4 CFD Test Matrix

Two main cases were analyzed in this CFD modeling study. The scope of the modeling work consisted of the baseline case and one subsequent case entailing oxy-firing (i.e., firing the coal in O_2- and CO_2-rich recycled flue gas. A description of each of the cases is provided in Table 3.1.

Table 3. 1: Matrix of CFD Cases

Case	Case Description	Nominal Inlet O_2 Mole Percent (Dry)	Nominal Inlet CO_2 Mole Percent (Dry)	Nominal Inlet N_2 Mole Percent (Dry)
1	Base Case: Nominally 15% Excess Air	21	0	79
2	Concept B Case: FGR / Oxy-Firing	29	65	6

Case 1 is the Base case or baseline case, which corresponds to a nominal 15% excess air and maximum continuous rating (MCR) load conditions. As discussed above, specific input parameters have been iteratively modified until calibration with the selected implicit boundary conditions was achieved to a reasonable approximation.

The FLUENT™ code, calibrated with the baseline combustion conditions, was subsequently used to evaluate the impact of oxy-firing the coal, with a prescribed CO_2/O_2 mole ratio. In comparison with the baseline run (Case 1), the Concept-B case (Case 2) displaces the mass of N_2 by an equivalent mass of CO_2 (approximately). Since the molecular weight of CO_2 is greater than that of N_2 by a factor of approximately 1.57, the density of CO_2 will be correspondingly greater than that of N_2, and the velocity of the recirculated flue gas streams through the windbox nozzles will therefore be smaller (by, e.g., 36% for pure streams) than that of the baseline case (for the same nozzle free area). Based on the presumption that windbox retrofits should not be required in any application of the proposed advanced FGR technology, a decision was made to preserve the same (computational) inlet nozzle area for both cases; therefore, both cases were compared on an equivalent inlet-area basis, rather than on an equivalent inlet-velocity basis.

Lateral and vertical waterwall heat flux or absorption distributions, as well as the overall heat transfer in the lower and upper furnace, have been extracted from the CFD calculations. Total and sectional (area-weighted) waterwall heat absorption rates, in addition to peak heat absorption values, have been used to quantify the heat absorption rates that constitute one of the metrics used as a basis of comparison between the cases. Metrics utilized to compare the cases include:

- the unburned carbon (UBC) in the fly ash at the outlet of the computational domain
- the outlet NOx emissions value at the outlet of the computational domain
- the horizontal (and/or vertical) furnace outlet temperature(s) (FOT)
- the outlet CO emissions value at the outlet of the computational domain
- total, sectional, and peak waterwall heat flux or absorption rates

Some of the metrics for the two cases, such as NOx, will be compared on a basis that is scaled or referenced to the baseline value, since relative trends may be a more reliable indicator of the expected performance than the absolute magnitudes of the metrics. Line and contour plots for variables of interest will be shown at selected horizontal or vertical planes, as appropriate.

The predictions provide a relative indication of the changes that may occur as a result of deviation from the baseline operating conditions. The difference in inlet nozzle velocities, mixture specific heats, and radiative properties between CO_2 and N_2 have an impact on the performance criteria and metrics of interest, e.g., HFOT, waterwall heat flux distributions, NOx emissions, and unburned carbon in the fly ash. Overall, the CFD runs serve to provide insight into the flow and reaction patterns of the lower furnace, as well as to provide relative changes in emissions and residual carbon values at the exit plane as a result of changes in operating conditions.

3.5 Input and Boundary Conditions

The case selected as the “Base case” is not strictly a replication of the experimental test data at one operating condition of the Conesville Unit No. 5. Rather, the Base case is defined as being a composite case which is representative of the available control room or board data over multiple test days. The board data encompasses variations in excess air conditions (14.1% to 18.1%), tilts, and burners designated as being out of service (BOOS)). From the array of operating conditions available in the board data, specific operating parameters were selected, which collectively constituted the “Base case” operating conditions. A RHBP simulation of the selected Base case test conditions helped to further delineate and refine some of the operating condition parameters.

Based on the project goals and requirements for flue gas recirculation and oxy-firing, the operating conditions of the Concept-B case were also subsequently defined. A RHBP run was conducted to study and evaluate the Concept-B case.

The mass flow rates, excess air conditions, inlet temperatures, tilts, etc. calculated in the RHBP runs were utilized directly in the CFD runs. A summary of the overall input parameters applied to the two cases are shown in Table 3.2.

Table 3. 2: Overall Mass Flow Rate Parameters

	Base Case	Concept B Case
Parameters	**(15% Excess Air)**	**(Oxy-Firing)**
Total Secondary Gas (lb_m/hr) =	2,968,935	2,933,877
Total Primary Gas (lb_m/hr) =	724,580	671,503
Total Gas (lb_m/hr) =	3,693,515	3,605,380
Total (As Rec'd) Coal (lb_m/hr) =	374,455	366,603
Total (Moisture-Free) Coal (lb_m/hr) =	336,635	329,576
Total (DAF) Coal (lb_m/hr) =	294,322	288,150
O_2:Coal Feed Ratio (by Mass) =	2.253	2.251
% Excess O_2 (by Mass) =	13.75	13.64
For Complete Combustion:		
% Exit O_2 (by Volume, Dry) =	2.594	3.599
% Exit O_2 (by Volume, Wet) =	2.351	3.083
% Exit O_2 (by Mass, Dry) =	2.687	2.698
% Exit O_2 (by Mass, Wet) =	2.534	2.520

As shown in the above table, the primary criterion used to put the two cases on an equivalent basis for comparison purposes was the oxygen-to-coal ratio (by mass), designated to be approximately 2.252. Additional input parameters, including the inlet gas composition, are provided in Table 3.3.

Table 3. 3: Overall Gas Property and Input Parameters

Parameters	Base Case (15% Excess Air)	Concept B Case (Oxy-Firing)
Primary Gas Mass Fraction:		
N_2	75.872	3.967
O_2	22.845	2.627
H_2O	1.283	2.867
CO_2	0.000	90.513
SO_2	0.000	0.026
Secondary Gas Mass Fraction:		
N_2	75.872	4.011
O_2	22.845	27.528
H_2O	1.283	2.115
CO_2	0.000	66.327
SO_2	0.000	0.019
Inlet Primary Gas Temperature (°F) =	156	140
Inlet Secondary Gas Temperature (°F) =	614	552
Nozzle Tilt (Degrees)=	-10	-10

The same coal composition for the mid-western bituminous fuel was utilized for both cases and is shown in Table 3.4.

Table 3. 4: Coal Composition and Heating Value

Coal Ultimate Analysis		Coal Proximate Analysis	
	As-Received Basis (Weight %)		As-Received Basis (Weight %)
Carbon	63.2		
Hydrogen	4.3		
Nitrogen	1.3		
Sulfur	2.7	Volatile Matter	32.7
Oxygen	7.1	Fixed Carbon	45.9
Moisture	10.1	Moisture	10.1
Ash	11.3	Ash	11.3
Total	100.0	Total	100.0
Higher Heating Value (Btu/lb$_m$) =		11,293	(As-Rec'd Basis)

The mass flow rates for the secondary gas feed streams are shown in Table 3.5.

Table 3. 5: Mass Flow Rates for the Secondary Gas Feed Streams

		Base Case		**Concept B Case**	
		(15% Excess Air)		**(Oxy-Firing)**	
Compartment / Nozzle Name	Damper Position (% Open)	Mass Flow Rate (lb_m/hr/elev)	Average Velocity (ft/s)	Mass Flow Rate (lb_m/hr/elev)	Average Velocity (ft/s)
CCOFA-C	0.0	74,671	48.1	73,789	33.8
CCOFA-B	49.7	286,760	184.8	283,374	130.0
Top Air / CCOFA-A	20.0	121,270	72.2	119,838	50.8
Coal-E (Fuel-Air)	60.0	108,016	217.7	106,741	153.1
Aux Air DE	33.0	412,925	115.0	408,049	80.9
Coal-D (Fuel-Air)	60.0	108,016	217.7	106,741	153.1
Aux Air CD	33.0	412,925	115.0	408,049	80.9
Coal-C (Fuel-Air)	60.0	108,016	217.7	106,741	153.1
Aux Air BC	33.0	412,925	115.0	408,049	80.9
Coal B (Fuel-Air)	60.0	108,016	217.7	106,741	153.1
Aux Air AB	33.0	397,744	115.0	393,048	80.9
Coal A (Fuel-Air)	0.0	23,874	48.1	23,593	33.8
Bottom-End Air	70.0	393,774	234.6	389,124	165.0
Totals:			% of Total Gas		% of Total Gas
Total Sec Through Windbox		2,968,935	80.38	2,933,877	81.37
CCOFA		482,701	13.07	477,001	13.23
Total Windbox - CCOFA		2,486,234	67.31	2,456,876	68.14
Aux Air		1,636,520	44.31	1,617,196	44.86
Fuel-Air/Gas		455,940	12.34	450,556	12.50
Bottom-End Air/Gas		393,774	10.66	389,124	10.79

The damper positions for the top two CCOFAs were extracted from field board data. The other damper positions were estimated, based upon ALSTOM Power's experience. A proprietary EXCEL spreadsheet calculator, denoted the Firing Zone Stoichiometry Calculations (FZSC) spreadsheet, was used to modify the damper positions slightly until the desired windbox-to-furnace Δp of 6.4 inches-of-water-gauge and the desired excess air (XA) were achieved. (The desired windbox-to-furnace Δp of 6.4 inches-of-water-gauge corresponds to a field measurement from the Conesville Unit No. 5 board data at a relatively low excess air condition of 14.1%.)

The mass flow rates for the primary gas and coal feed streams are shown in Table 3.6.

Table 3. 6: Mass Flow Rates for the Primary Gas and Coal Feed Streams

	1.1.1.1.1.1 Base Case **(15% Excess Air)**			*1.1.1.1.1.2 Concept B Case* **(Oxy-Firing)**		
Nozzle	Primary Air Mass Flow Rate (lb$_m$/hr/elev)	Average Air Velocity (ft/s)	Coal Mass Flow Rate (lb$_m$/hr/elev)	Primary Gas Mass Flow Rate (lb$_m$/hr/elev)	Average Gas Velocity (ft/s)	Coal Mass Flow Rate (lb$_m$/hr/elev)
Coal E	178,909	96.2	93,614	165,803	60.8	91,650.8
Coal D	178,909	96.2	93,614	165,803	60.8	91,650.8
Coal C	178,909	96.2	93,614	165,803	60.8	91,650.8
Coal B	178,909	96.2	93,614	165,803	60.8	91,650.8
Coal A	8,945	4.8	0	8,290	3.0	0.0
Total:	724,580		374,455	671,503		366,603.0

It should be noted that the bottom coal elevation (Coal A) has been designated as being out of service (OOS) for these two cases. For the Coal-A nozzle, the coal mass flow rate is completely turned off and only a small amount (e.g., 5% of the normal allocation) of air or gas is permitted to pass through the coal nozzle (i.e., to help purge and cool the nozzle).

The inlet turbulence intensity was arbitrarily set at 10% of the mean velocity. The dissipation length scale for the turbulence was calculated using an effective hydraulic diameter for the inlet.

As mentioned previously, the overall wall absorption and the HFOT constitute implicit boundary conditions. The RHBP is the external resource which is used to provide values for the tube bank heat extraction, the overall wall absorption, and the HFOT. The overall resistance to wall heat transfer in the CFD computation is iteratively changed until the calculated implicit boundary condition constraint (i.e., wall absorption and HFOT) matches the RHBP values to within acceptable accuracy. Since the wall absorption computed by the CFD must ultimately match the wall absorption calculated by the RHBP, a brief tutorial of the RHBP nomenclature is required. The RHBP arbitrarily divides the furnace and convective pass into the sections shown in Figure 3.7.

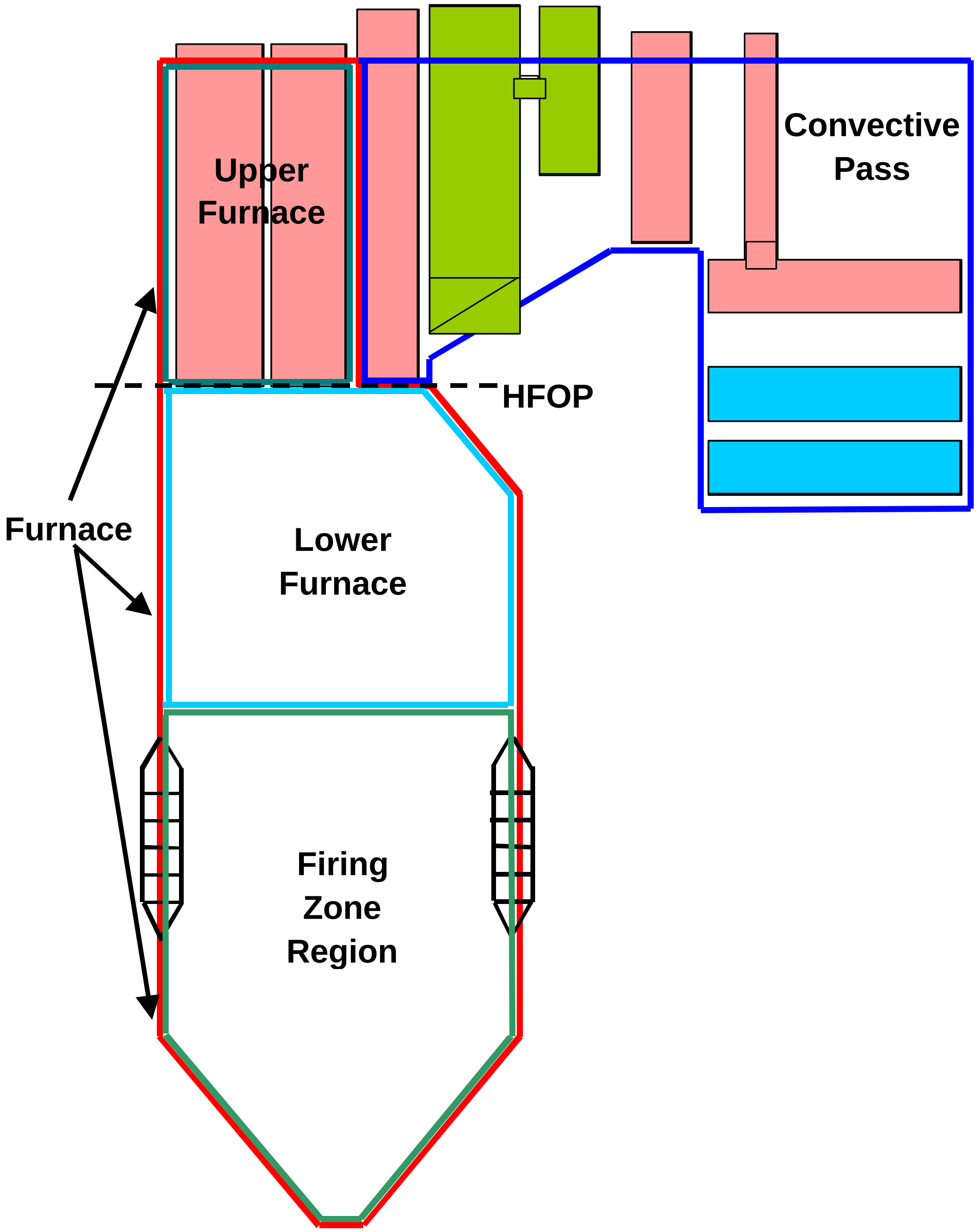


Figure 3. 7: Furnace and Backpass Sections Specific to the RHBP

The “furnace” volume (outlined in Figure 3.7) is defined within the RHBP to include most of the radiant wall sections and extends from the bottom of the coutant to the roof of the upper furnace; it does not include the convective pass section, nor the superheater platen tube bank. The “furnace” volume encompasses 3 partitioned sections. The first section is the “firing zone” region, which extends from the bottom of the coutant to several feet above the top coal elevation. The “lower furnace” section extends from the top of the firing zone section to the HFOT plane. The “upper furnace” section extends from the HFOT to the roof and encompasses the superheater division panels, but not the superheater pendant platens. For all surfaces in the “furnace” volume, the RHBP generates the appropriate fouling factors.

The convective pass and backpass volume is defined by the RHBP to encompass all of the tube banks and their bounding surfaces, with the exception of the division panels. For convenience sake, the RHBP treats the convective pass volume differently than it does the furnace volume. For the convective section, the RHBP generates surface effectiveness factors, rather than fouling factors.

The HFOP is at the apex of the arch. The VFOP (not shown) is defined by the RHBP as being at the inlet to the superheater platen (both along the top and along the bottom of the platen), but this definition will be modified later on. The nomenclature utilized by the RHBP (with the exception of the VFOP definition), will be adhered to in the ensuing discussion.

The material properties and relative amounts of slag and fouling deposits found locally on the superheater division panels and the bounding tube walls are usually not known *a priori*. Consequently, the surface resistance to heat transfer in the CFD computation must be iteratively adjusted throughout the furnace volume in order to match the horizontal furnace outlet temperature (HFOT) calculated by the RHBP. For heat transfer to the computational waterwalls, radiant reheat surfaces, and superheat surfaces, the boundary conditions were prescribed using an effective overall surface resistance (or overall fouling factor) and a back-side (steam) temperature. It should be noted that the fouling factors calculated by the RHBP are not utilized directly in the CFD computation. Since the radiation and heat transfer submodels in the RHBP are significantly different than those found in a CFD code, it follows that the fouling factors and emissivities generated by the RHBP may also be quite different from those calculated by the CFD code.

In the waterwalls, the backside (two-phase) steam-water temperature corresponds to saturated conditions and was assigned a temperature of 680 °F. Since the fouling factor must be iteratively changed in the CFD calculation in order to calibrate the wall absorption and HFOT to the RHBP value, all of the backside steam temperatures were equated to the (constant) saturation steam temperature, including the radiant reheat and superheat surfaces. Although the superheat temperatures in the bounding surfaces will be higher than the saturation temperature, the goal of the calibration exercise was principally to match the overall absorption in the furnace proper, the convective pass being secondary, and the approximation was thought to be acceptable. The wall heat transfer parameters found to calibrate the Base case to the RHBP results are shown in Table 3.7.

Table 3. 7: Wall Heat Transfer Parameters in the CFD Cases

Interior Wall Surface	Internal Wall Emissivity	Overall Fouling Factor (hr-ft^2-°F/Btu)
Coutant and Wall Below Windbox	0.80	0.0175
Firing Zone	0.80	0.0232
Lower Furnace Waterwall (Above Windbox)	0.75	0.0258
Radiant Reheat Walls (Lower Furnace)	0.75	0.0247
Upper Furnace Waterwalls/Radiant Reheat	0.65	0.0378
Upper Furnace Roof	0.65	0.0258
Superheater Division Panels	0.65	0.0378
Convective Pass and Backpass	0.75	0.0232
Roof of Convective Pass and Backpass	0.75	0.0142
Backside Wall / Steam Temperature (°F) =	680	

One of the quantifiable metrics selected as a means of comparing the Base case and the Concept-B case is the overall wall absorption. In order to compare the two cases on an equivalent basis, it was deemed necessary to maintain the same fouling factors throughout the "furnace" volume in both cases. Therefore, the two corresponding RHBP runs were generated with the same fouling factors. However, convective section surface effectiveness factors and tube bank absorptions could not also be forced to be identical for the two cases in order to permit the RHBP to converge to a physical solution based on its solution constraints (i.e., prescribed superheat and reheat temperatures).

The tube banks are approximated as porous media in the CFD cases; parameters describing the pressure drop through each tube bank are calculated from industry design standards. The RHBP is used to evaluate the steam side performance and to estimate the furnace outlet temperature and the heat duty of the various tube bank sections. The amount of heat extracted from each tube bank, as calculated from the RHBP, with the exception of the superheater division panels (which are calculated with the resistance and backside temperature approach), are shown in Table 3.8. The tube bank absorption quantities also include the absorption through any nearby screen tubes.

Table 3. 8: Energy Extracted from Tube Banks

	Base Case (15% Excess Air)	**Concept B Case (Oxy-Firing)**
Tube Bank	Heat Transferred to Steam in Banks (Btu/hr)	Heat Transferred to Steam in Banks (Btu/hr)
Superheater Pendant Platen	2.930E+08	2.942E+08
Reheater Pendant Platen (Front) Assemblies	3.957E+08	3.753E+08
Reheater Finishing Section	1.424E+08	1.286E+08
Superheater Finishing Section	2.397E+08	2.194E+08
Superheater Low Temp (Vertical) Pendant Section	1.298E+08	1.253E+08
Superheater Horizontal Spaced Rear Assemblies	2.998E+08	2.807E+08
Economizer Assembly	5.118E+08	4.594E+08
Totals:	2.012E+09	1.883E+09

Although the fouling factors in the “furnace” volume were forced to be identical in the two RHBP cases, the heat absorption through each tube bank in the convective pass was allowed to self-adjust to meet the RHBP solution and convergence criteria. Therefore, the tube bank heat extractions for the two cases are somewhat different. The tube bank absorption for the Concept-B case is generally lower than (but within 10% of) the tube bank absorption in the Base case.

The above absorptions were specified in the CFD calculation for each tube bank in the form of a sink term to the energy transport equation.

3.6 Submodel Parameters and Case Preparation

Calculations of the two Conesville cases were run using FLUENT™ Version 5.4.8. FLUENT™ is a general purpose CFD code with a range of physical model capabilities. For these simulations, the important fluid flow, heat transfer, gas and solid phase combustion processes were represented in a steady-state calculation. Submodel selections and their parameter settings are described below.

The SIMPLE pressure-correction algorithm was used with first-order upwind differencing. The turbulence model was the traditional or standard k-ε turbulence model. The turbulence-chemistry interaction was modeled using the composite Magnussen eddy-breakup (mixing-limited / fast-chemistry) and Arrhenius rates (kinetically-limited) combustion submodel, in which the smaller of the two rates (eddy-breakup or Arrhenius rates) takes local precedence. The gas-phase chemistry was based on two, irreversible, global steps: (1) volatile combining with oxygen to form carbon monoxide, water, nitrogen, and sulfur dioxide, and (2) carbon monoxide combining with oxygen to form carbon dioxide.

The Discrete Ordinates radiation model was used to calculate the local radiative fluxes. The particle absorption and scattering efficiencies were specified via a proprietary user-defined function (UDF). An implementation inconsistency was found in the weighted-sum-grey-gas (WSGGM) model in the FLUENT™ code (which ultimately forced the CFD runs to be repeated and delayed the final report). To circumvent the use of the WSGGM submodel, a gas-phase absorption coefficient, calculated from the Edwards wide-band model, was calculated and utilized instead. Unfortunately, the polynomial prescriptions for the gas-phase absorption coefficient in the FLUENT™ input permitted the coefficient to be a function of local temperature only, whereas it should ideally be a dual function of both local concentration and temperature. Consequently, the gas-phase absorption coefficient was input as a linear function of temperature, calculated from a single CO_2 and H_2O partial pressure that prevailed in the computed upper furnace region. (While the complete wide-band model cannot be incorporated in the time frame required to impact the present CFD runs, a recommendation will subsequently be made to Fluent Inc. to upgrade their gas-phase radiative properties submodel to incorporate a full and reliable wide-band model.)

The particles were modeled as discrete trajectories using the stochastic Lagrangian particle model. Particles were injected from 16 inlet surfaces (8 cells per injection surface), corresponding to each of the primary coal inlets (4 coal elevations), with 10 particle size bins per distribution and 6 stochastic tries, for a total of 7,680 discrete particles that were tracked through the domain. Conesville Unit No. 5 mill classifier or sieve analysis data was available, which were averaged to produce a composite analysis with 98.35% through 50 mesh and 88.55% through 100 mesh. A Rosin-Rammler fit of the data produced a mean particle diameter of 66 microns and a slope/spread parameter of near unity. Maximum and minimum particle sizes were chosen as 200 microns and 1 micron, respectively. This information was utilized to calculate the particle size distribution through the Rosin-Rammler utility in FLUENT™.

Each of the particles followed a stochastic particle trajectory, progressing through the sequential stages of heating, devolatilization and char oxidation according to local flame conditions. The mill module, one of the modules within the company proprietary Reheat Boiler Program (RHBP), estimated that the amount of moisture in the as-fired coal would be approximately 4.5% (from an as-received moisture value of 10.1%). However, to simplify the CFD calculations, all of the coal moisture was presumed to have evaporated completely in the mill and was given to the primary transport air. Devolatilization was calculated from the Kobayashi-type two-competing-rates submodel. The high-temperature volatile yield parameter was estimated using the commercially available Niksa Energy Associates (NEA) PC COAL LAB software (i.e., FLASHCHAIN model). The high-temperature volatile yield was specified to be 67.6% on a DAF basis (as calculated for 55 micron particles at 25 °C entrained in nitrogen at 1600 °C in a furnace at 1700 °C and atmospheric pressure). The same high-temperature volatile yield was specified for all particle sizes. Based on the results of an in-house, proprietary study, the Kobayashi devolatilization parameters were modified somewhat to give accelerated devolatilization rates which more closely approximated the FLASHCHAIN rates.

Parameters for the char oxidation kinetics of the mid-western bituminous coal fired in Conesville Unit No. 5 were set initially through comparison of the proximate and ultimate analyses of the coal with other coals tabulated in an in-house, drop-tube, kinetics database. Based on a comparison with the database, a reasonable surrogate fuel was found to be Pittsburgh No. 8. Consequently the database kinetics for Pittsburgh No. 8 were used for the Conesville case. The activation energy was not subsequently changed, but the pre-exponential factor was iteratively adjusted as necessary to match the desired unburned carbon in the ash at the exit (of approximately 2%) for the baseline case. The char oxidation submodel in FLUENT™ is a composite kinetics/diffusion-limited char combustion model. In the char oxidation submodel, the char is oxidized via heterogeneous reaction to form CO, which is then oxidized further in the gas phase to form CO_2.

The species (gaseous) properties, such as thermal conductivity, viscosity, and specific heat, were calculated at each cell in the domain based on polynomial functions of temperature. The species specific heats did not account for dissociative effects. The gas mixture properties were based on composition dependent formulations of the properties of the pure species.

The NOx model is run in a post-processor fashion. NOx predictions were performed for a combination of both thermal and fuel NOx submodels. Destruction through the reburn mechanism is permitted only for the presumed probablility distribution function (PDF) combustion model, which was not utilized in this case. For thermal NOx, the partial equilibrium assumption was activated for atomic oxygen, and a beta function PDF was utilized for the convolution over temperature. For the fuel NOx model, atomic nitrogen fractions in both the volatile and the char were specified. The char nitrogen was presumed to go to HCN, rather than directly to NO. NOx and CO emissions are typically more difficult to predict than gross flow features or overall heat release because they are more sensitive to accurate calculations of local temperature and require more detailed chemistry. Consequently the NOx predictions were ultimately scaled to the measured Base case value.

Further grid refinement or adaption of the grid was not performed for the two cases. It is not known to what extent numerical diffusion, typically associated with coarse or skewed meshes, has impacted the quantitative accuracy in the present study. Since the focus of the present investigation was to discriminate between the two cases on a relative basis only, the present grid was presumed to be acceptable.

3.7 Calculation Results

The baseline case (Case 1) and the Concept-B case (Case 2) will be compared on the basis of the metrics discussed previously in Section 3.3. All of the gas-phase, planar integrated scalars, such as temperature and species, will be reported on a mass-averaged basis.

The calculated residual carbon in the fly ash, as well as the NOx and CO emissions are shown in Table 3.9.

Table 3. 9: Comparison of Predicted Residual Carbon and Emissions for Cases 1 and 2

	Base Case (15% Excess Air)	**Concept B Case (Oxy-Firing)**
Residual Carbon in Ash (%)	2.4	0.8
NOx (ppmm, uncorrected, dry) [Note 1]	190	104
CO (ppmv, dry)	141	82
Note 1:	Baseline value was scaled to measurement	

The carbon-in-ash measured in the field was between 2% and 3%; therefore for the baseline case, the pre-exponential factor for the char oxidation kinetics was adjusted until such a representative value was calculated. Relative to the baseline case, the carbon-in-ash value for the Concept-B case was calculated to be significantly lower. The same trend is also exhibited by the emissions calculations. If the calculated NOx value is scaled to the representative measured value of approximately 190 ppm, and assuming that the Concept-B calculated value can be proportionately scaled as well, then the Concept-B NOx value may be expected to decrease by about a factor of two relative to the baseline value. The calculated CO value for Concept-B is also much lower than its baseline counterpart. Since experimental CO values were reported as being “under range”, and are therefore not available, and since the inclusion of reliable CO kinetics within the CFD computation is not considered to be practical at present, the CO results, as well as the unburned carbon results, may be taken as an indication of the extent of mixing, the temperature history, and the residence time differences between the two cases.

The planar temperatures at the HFOP, VFOP, and economizer outlet plane constitute implicit boundary conditions or target temperatures, provided by the RHBP, that the CFD calculations attempts to match through modification of its wall heat transfer boundary conditions. The aforementioned planes are depicted in Figure 3.8, to show their relative positioning.

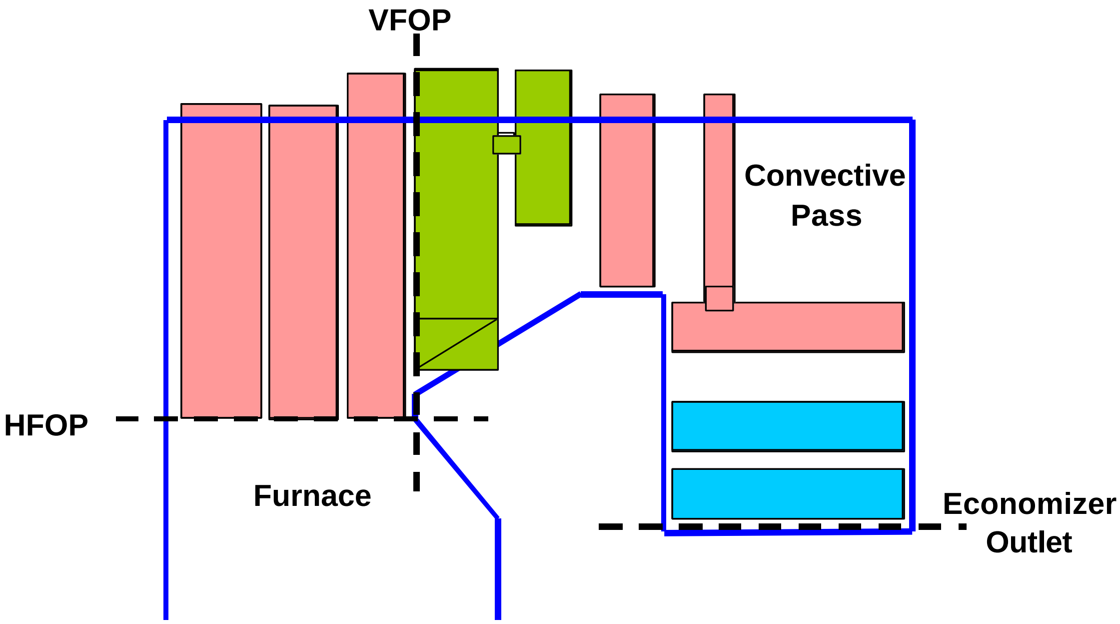


Figure 3. 8: Depictions of Specific Planes for Comparisons of Averaged Temperatures

The temperatures at various cross-sections, as well as the integrated wall absorption over various furnace wall sections (largely in accordance with the RHBP partitioning and definitions in Figure 3.7), are shown in Table 3.10. The definition of the “upper furnace” section, prescribed by the RHBP to exclude the superheater platen, is inconvenient for CFD computations. Hence, the term has been modified in the table below, such that the heat extracted in the superheater platen (via a sink term to the energy equation), and its bounding surfaces, is now included in the definition of the “upper furnace” term.

Table 3. 10: Comparison of Predicted Gas Temperature and Sectional Wall Absorption Values for Cases 1 and 2

	Base Case		**Concept B Case**	
	(15% Excess Air)		**(Oxy-Firing)**	
	RHBP	FLUENT™	RHBP	FLUENT™
Average Planar Temperature (°F):				
HFOP	2,756	2,765	2,729	2,676
VFOP [Note 1]	2,185	2,125	2,027	2,040
Economizer / Domain Outlet	706	666	683	696
Furnace Absorption (Btu/hr):				
Firing Zone Region	5.06E+08	4.37E+08	4.83E+08	4.33E+08
Lower Furnace	5.28E+08	5.92E+08	5.85E+08	5.88E+08
Total Absorption Below HFOP	1.03E+09	1.03E+09	1.07E+09	1.02E+09
Upper Furnace (Note 2)	8.95E+08	8.96E+08	9.74E+08	8.88E+08
Total Furnace	1.93E+09	1.93E+09	2.04E+09	1.91E+09
Percent Change of Absorption from Baseline			+5.8	-0.8
Note 1:	Definition of "VFOP" modified to be at the platen outlet			
Note 2:	Definition of "upper furnace" modified to include platen			

The wall absorption throughout the furnace dictates the resultant temperatures at the HFOP and VFOP. (The domain outlet temperature is a less reliable indicator, since the heat extracted from the tube banks is different for the two cases.) The FLUENT™ code does not presently permit the user to interrogate the net radiative flux through an arbitrary interior plane, so it is difficult to know to what extent radiation losses through each section, and through the porous media (representing the tube banks), are impacting the above results. Therefore, while it may not be possible to fully ascertain the energy entering and leaving each section in the CFD computation, and thus provide a more detailed explanation as to the magnitudes of the temperatures and fluxes, some general conclusions about the relative trends may be drawn.

In the baseline case, the HFOT and wall absorptions were calibrated to the RHBP values (as shown above in Table 3.10). Although the individual absorptions in the firing zone region and the lower furnace region are not duplicated, the total absorption below the HFOP, as well as the total absorption in the upper furnace are matched to an acceptable degree. The reason that the individual absorptions in the firing zone and lower furnace regions are not matched is because the CFD computation is much more sensitive to the location of the burner-out-of-service (BOOS) than the RHBP (as will be apparent later when the absorption profiles are shown). Since the lowest coal elevation has been dictated to be out of service (OOS), a smaller number of burning particle trajectories traverse the coutant region than would occur if the lowest coal elevation were in service; hence the heat release and wall absorption in the region below the windbox is also commensurately less. Alternatively, the RHBP effectively treats all of the firing zone region (below the "lower furnace") as a single, well-mixed, global volume (with an allowance for a radiation boundary layer). Since the RHBP has a long history of successful designs that encompass a wide variation in firing system configurations, it is suspected, although not proven, that the CFD computation may be overly sensitive to the location of the BOOS.

When the RHBP run for Case 1 is compared with the RHBP run of Case 2, it is evident from the calculations that the HFOT for Concept-B decreases by 29 °F and the VFOT decreases by 158 °F (for an average decrease of about 94 °F) from the corresponding baseline values. The overall furnace wall absorption for Concept-B is expected to increase over that of the baseline (by 5.8%). This result is in accordance with a literature source (Wilkinson, et al., 2001) for refinery gas fired into a wall-fired boiler, in which a one-dimensional, slice-type model (i.e., RHBP-type model) was used to perform the analysis. The authors found that the oxyfuel concept experienced an increase in the incident radiative heat flux of about 4% to 6%, and a corresponding increase in the furnace heat absorption of approximately 9%, relative to the baseline air concept. This increase was attributed by the authors to "the increase in radiative power of the carbon dioxide and water-rich combustion gases".

The CFD computations exhibited a similar temperature trend to that of the RHBP. Both the HFOT and VFOT for the Concept-B case decreased by 85 and 89 °F, respectively, relative to the baseline results. On the other hand, the CFD cases exhibited a trend in wall absorption which was not as strong as the RHBP results. The wall absorption for the Concept-B case decreased by less than 1% relative to the furnace absorption of the baseline CFD case. Additional CFD runs, in which the fuel was changed from coal to an equivalent gas-only fuel (i.e., volatile), confirmed the trend (i.e., the Concept-B absorption was slightly lower than the baseline absorption). This same CFD trend is partially supported in the literature (Chui, et al., 2001) for coal burned in a cylindrical, down-fired vertical combustor, in which both CFD and experimental data were utilized to assess performance. The authors found that both the CFD calculation and the measured data showed that the oxygen-enriched, recycle-gas case exhibited a significant decrease of about 18% in the incident radiant flux, relative to the baseline case. Chui, et al. attributed this decrease to:

(a) the change in aerodynamic and temperature patterns due to the lower inlet velocity and penetration of the recycled flue gas (i.e., due to the higher molecular weight of CO_2 relative to N_2),
(b) a decrease in the overall flame temperatures of the recycled flue gas case due to the higher specific heat of CO_2 relative to N_2, and principally to
(c) the increase in the absorption coefficient in the recycled flue gas case because of the higher concentrations of the optically absorbing CO_2 and H_2O, relative to the optically transparent N_2.

The analysts of both the (1) one-dimensional/slice or global mixing codes (like the RHBP) and the (2) CFD codes rely on the enhanced radiative absorption coefficient associated with CO_2 to partially explain the furnace absorption results, although the calculated trends exhibit different apparent sensitivities for the two types of codes. Apart from differences in the gas-phase radiation properties themselves, a potential explanation for this apparent dichotomy is not readily available. However, in the CFD calculations, the high flame temperatures and patterns are a function of the local aerodynamics and may be effectively isolated from the wall by a significant presence of colder gases. Any CO_2 (and H_2O) in the near-wall gases will absorb radiative energy at the higher temperature of the "fireball", but will re-emit at the lower near-wall temperature, thus acting to increase the temperature of the near-wall gases and decreasing the net wall heat absorption. Slice or global-mixing codes will try to mimic this same effect and may, for example, presume a uniform firing zone or slice temperature radiating through a (colder) "radiation boundary layer" adjacent to the wall. The quantitative impact of the near-wall or "radiation boundary layer" gases on the net radiative flux to the wall may be significantly different between the two types of codes. The CFD computations may simply exhibit a greater "insulative" effect to radiation in the near-wall regions than that provided in RHBP-type models due to aerodynamic (case-specific) and submodel differences.

As discussed earlier, due to an implementation inconsistency in the FLUENT™ code, the gas-phase radiation absorption coefficient utilized in the CFD computational was a linear function of local temperature only, and not a function of concentration. While it is anticipated that the gas-phase absorption coefficient will be sensitive to the local concentrations of CO_2 and H_2O, it is not known at this juncture whether the difference in overall wall absorption between the two cases would increase or decrease relative to its present values, and what the difference in magnitude would be. The calculations should be repeated with a gas-phase radiation absorption coefficient submodel that uses the full wide-band model (and that is more reliable than the WSGGM model presently in FLUENT™).

Vertical profiles of the maximum and average net wall heat flux for the baseline case and the Concept-B case are shown in Figures 3.9 and 3.10, respectively. At any given elevation, the net wall heat flux value represents the area-weighted average of all the cells positioned “circumferentially” around the furnace (at that elevation). The maximum local net wall heat flux is simply the maximum value of all of the “circumferential” cells (at a given elevation). The radiative portion of the net wall heat flux, the portion of the net wall heat flux due to conduction, and the incident radiative flux are also shown. A geometry profile is also provided, showing the coutant bendline, the windbox, and the arch bendline. The recessed line in the geometry profile represents the windbox, while the slanted line with the positive slope represents the coutant, and the slanted line with the negative slope represents the arch bendline. The maximum and average values of the net wall heat flux calculated from “idealized” in-house design standards correspond to the base case conditions only (on both Figures 3.9 and 3.10) and are provided simply for comparative reference purposes.

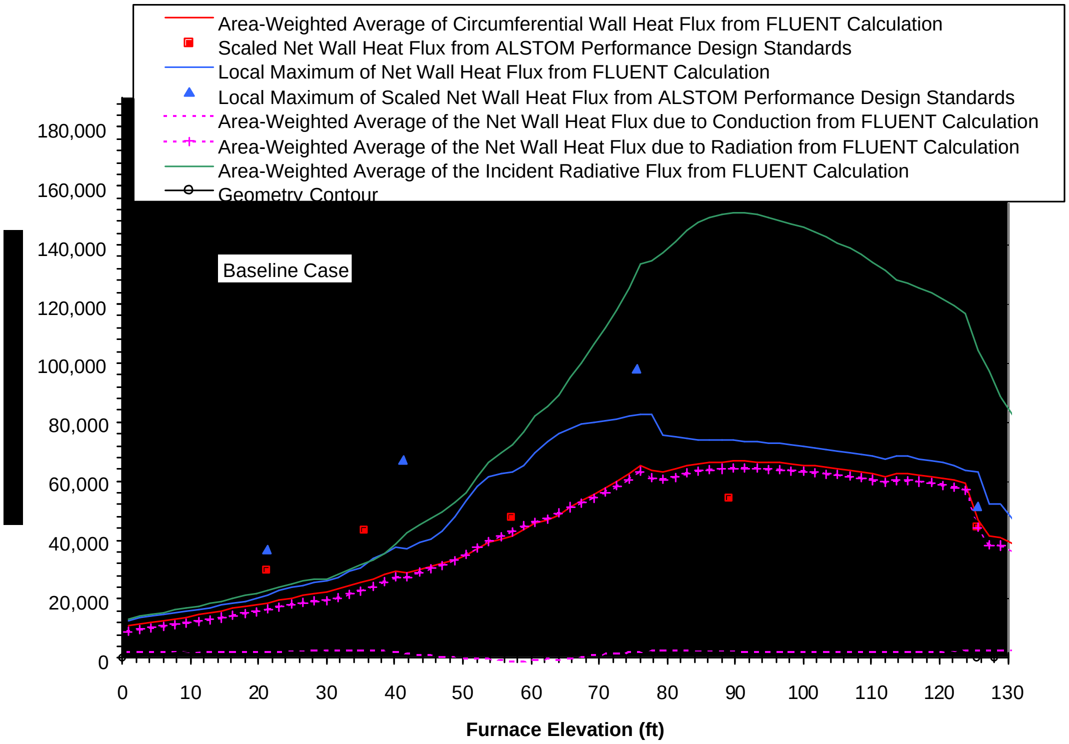


Figure 3. 9: Vertical Profiles of the Local Peak and Area-Weighted Net Wall Heat Flux for the Base Case

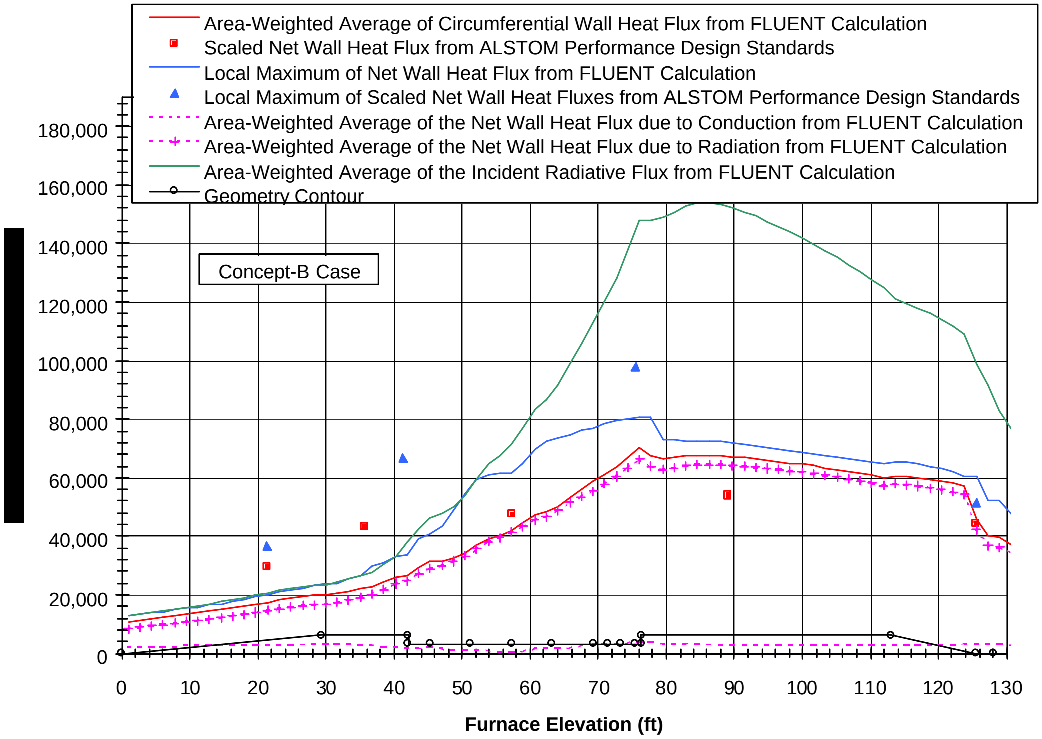


Figure 3. 10: Vertical Profiles of the Local Peak and Area-Weighted Net Wall Heat Flux for the Concept-B Case

For the Base case (Figure 3.9), it can be seen that the CFD-calculated average net wall heat flux is less than the idealized design standard in the furnace volume below the windbox, and is above the design standard in the furnace volume above the windbox. This is due again to the fact that the lower coal is OOS, which means that fewer particles are carried down toward the bottom of the furnace, and the heat release generated in the coutant region is correspondingly smaller than it otherwise would be. (Computational runs with the lower coal in service show a heat release in the coutant region which is similar in magnitude to the design standard profile.)

In a comparison of Figures 3.9 and 3.10, it can be seen that the Base case has a slightly higher peak and average wall heat flux, over most of the furnace elevation, than the Concept-B case, as has already been established. However, differences in the overall absorption between the two cases are rather small (i.e., less than 1%) and may perhaps be construed to be part of the “noise” in the CFD calculations.

Vertical profiles of the peak, standard deviation, and mass-averaged, planar, gas temperatures are provided in Figures 3.11 and 3.12. The area-averaged interior wall temperature is also shown in both figures. The peak temperatures and standard deviations of the local gas temperature (i.e., a type of unmixedness) for the baseline case are somewhat higher than those of the Concept-B case. The peak temperature differences are higher in the windbox region, with a delta of approximately 200 °F or more for the peak temperature in the mid-windbox region, which then diminishes slightly to about 150 °F as the arch elevation is approached. The standard deviation of the local gas temperature for the baseline case is about 150 °F higher than that of the Concept-B case in the windbox region, although the difference then diminishes and becomes negligibly small as the HFOP is approached. (It is difficult to say whether the difference in standard deviation between the two cases reflects a significant difference in mixing efficiencies or simply a difference in local peak temperature values.) Over most of the furnace height, except in the immediate vicinity of the CCOFA jets, the mean temperature and the interior wall temperature are generally higher in the baseline case than in the Concept-B case. (Table 3.10 shows that the baseline HFOT is approximately 90 °F higher than that of Concept-B.)

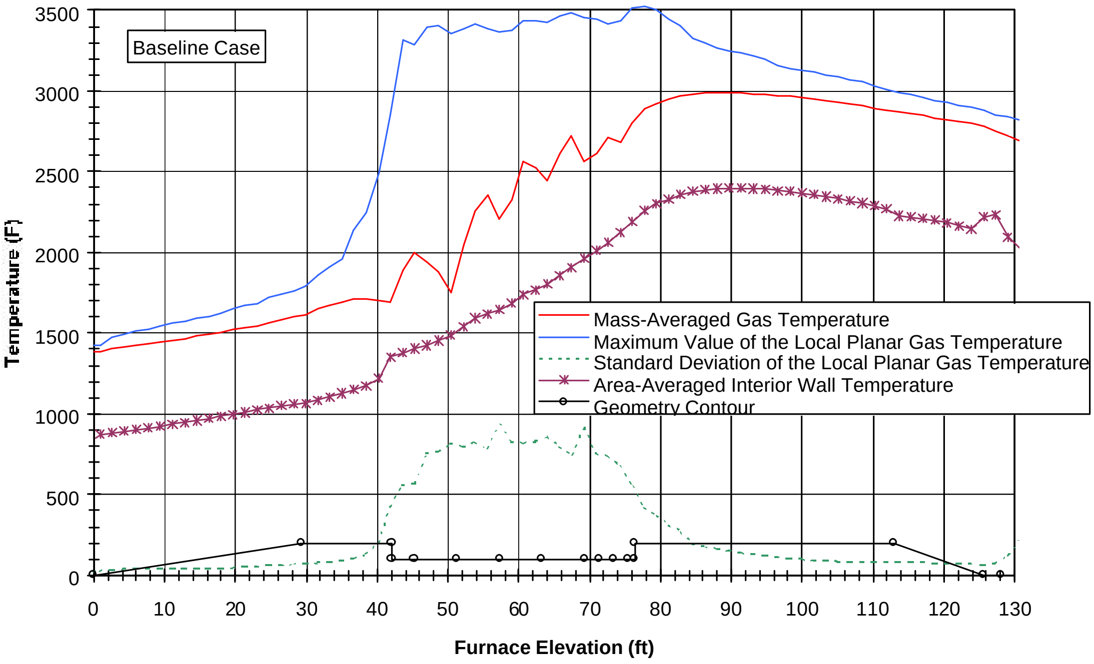


Figure 3. 11: Vertical Profiles of the Local Peak and Mass-Averaged Gas Temperatures for the Base Case

The specific heat of CO_2 is about 6% to 7% greater than that of N_2 over the range of 1340 °F (1000 K) to 3140 °F (2000 K). Therefore, for equivalent heat release and wall absorption, the Concept-B case would be expected to have a slightly lower mean gas temperature. The specific heat of H_2O (gaseous) is a factor of 2 to 3 times that of CO_2, but the mass fraction of H_2O in the inlet mixtures of the Concept-B case is only about a percentage point higher than that in the baseline case; thus its impact on lowering the Concept-B HFOT is minimal. Consequently, it is expected that the specific heat consideration of CO_2 (alone) should be a factor which would ordinarily cause the HFOT for the Concept-B case to decrease relative to that of the baseline case.

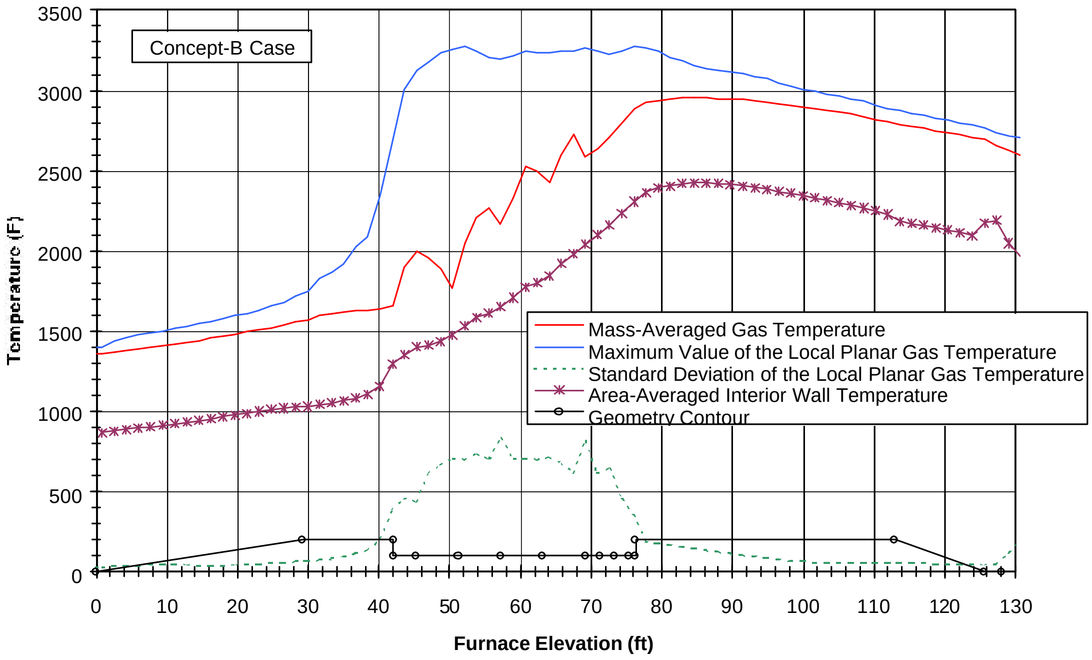


Figure 3. 12: Vertical Profiles of the Local Peak and Mass Averaged Gas Temperatures for the Concept B Case

3.8 Contour Plots

Contour plots of temperature at selected elevations are given in Figure 3.13. The plots show reasonable symmetry of the flowfield, with a relatively cool inner core, surrounded by an annular, high-temperature region. In certain regions, cooler near-wall temperatures may separate the high-temperature annular "ring" from the walls.

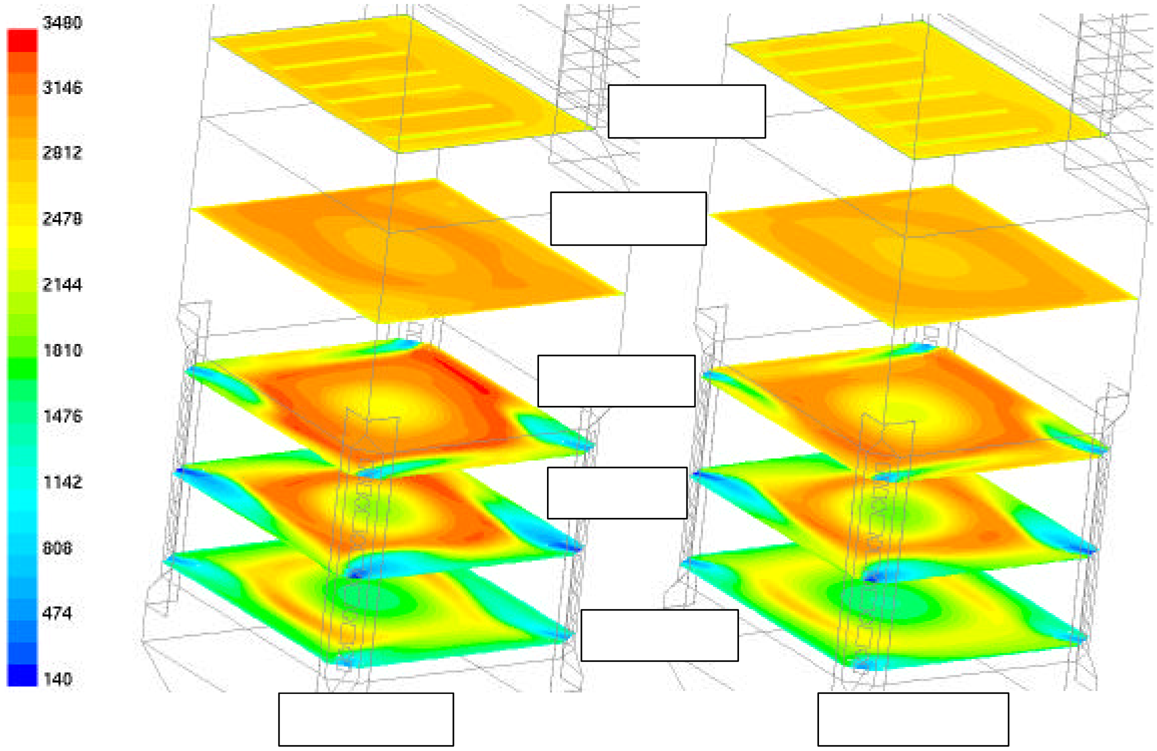


Figure 3. 13: Planar Contour Plots of Temperature at Various Elevations for the Base Case and Concept-B Case

A cross-sectional view of the temperature contours at two planes is given in Figure 3.14.

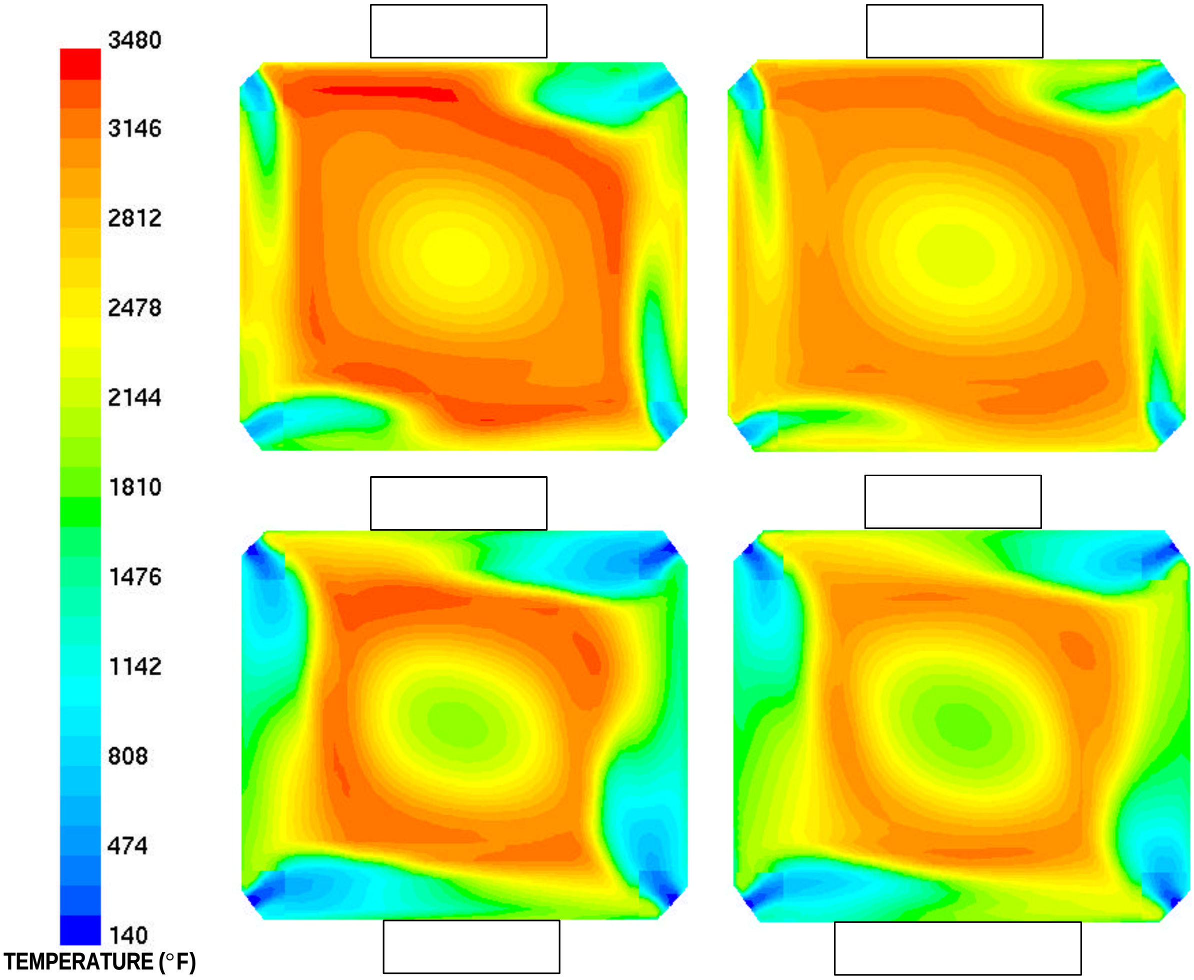


Figure 3. 14: Planar Contour Plots of Temperature at two Elevations for the Base Case and Concept-B Case

As can be seen in Figure 3.14, the aerodynamics and flame regions are rather similar in shape and appearance, with minor differences in temperature magnitude. The higher inlet velocities and penetration of the baseline case (see Tables 3.5 and 3.6) are apparent, in that a longer traverse distance is required before the jet is either swept upward out of the plane or mixed out to equivalent temperatures. The swirl is sufficiently strong to prevent the jets from penetrating radially to the core; rather the jets are quickly "bent over" in the direction of the swirl crossflow vectors (in the cross-sectional plane) and any residual jet penetration is in that same direction. It should be noted that although the difference in inlet velocities between the two jets is a factor in the resultant aerodynamic penetration, a perhaps equally influential factor is the velocity acceleration due to combustion. The coal jets, for example, enter the domain at 96.2 ft/s and 60.8 ft/s, for the baseline case and the Concept-B case respectively, but the jets are rapidly accelerated to approximately 170 ft/s (figure not shown) within a few nozzle diameters of the entrance.

The peak temperatures in the annular high-temperature annular "ring" are higher in the baseline case than in the Concept-B case. By the time the CCOFA is reached, the core temperature of the baseline case also appears to be higher than that in the Concept-B case (contributing ultimately to a higher HFOT for the Base case).

Temperature plots at the mid-plane of the x-coordinate direction are shown in Figure 3.15. The plots are again qualitatively very similar in appearance.

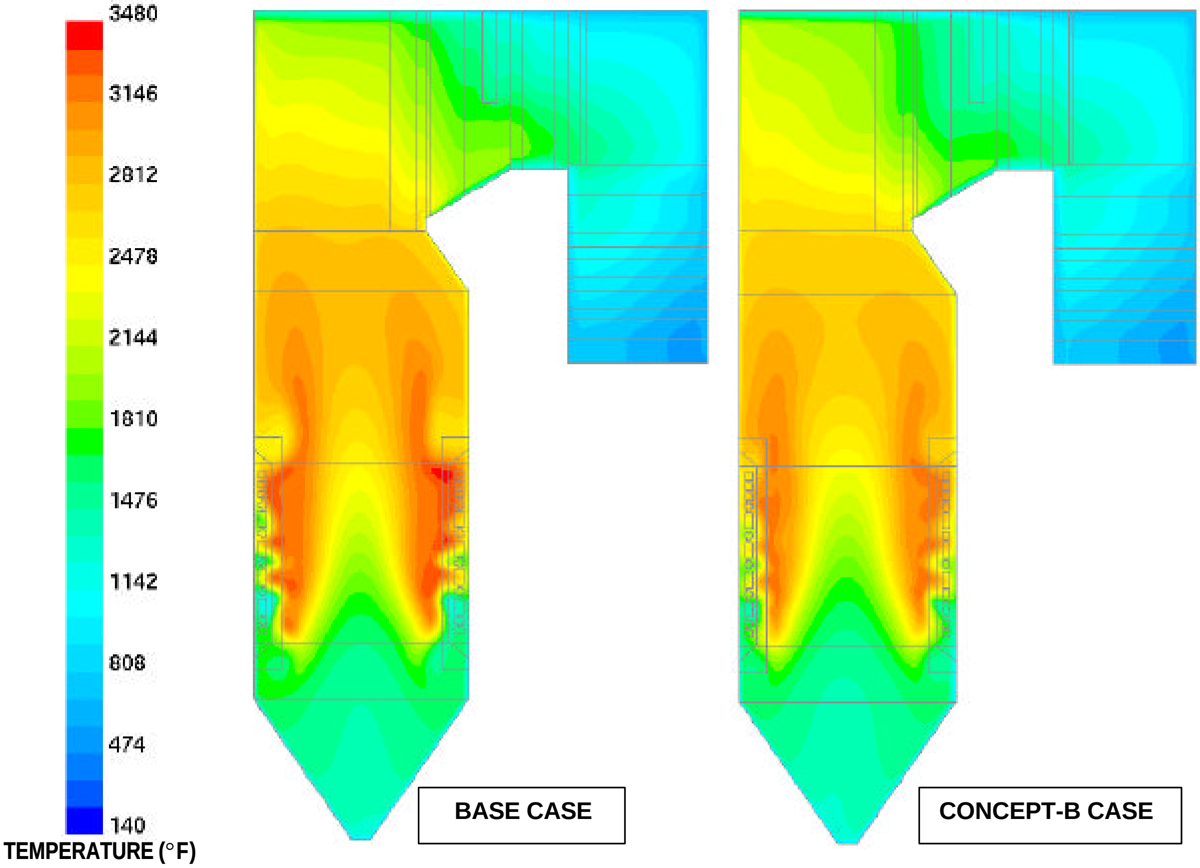
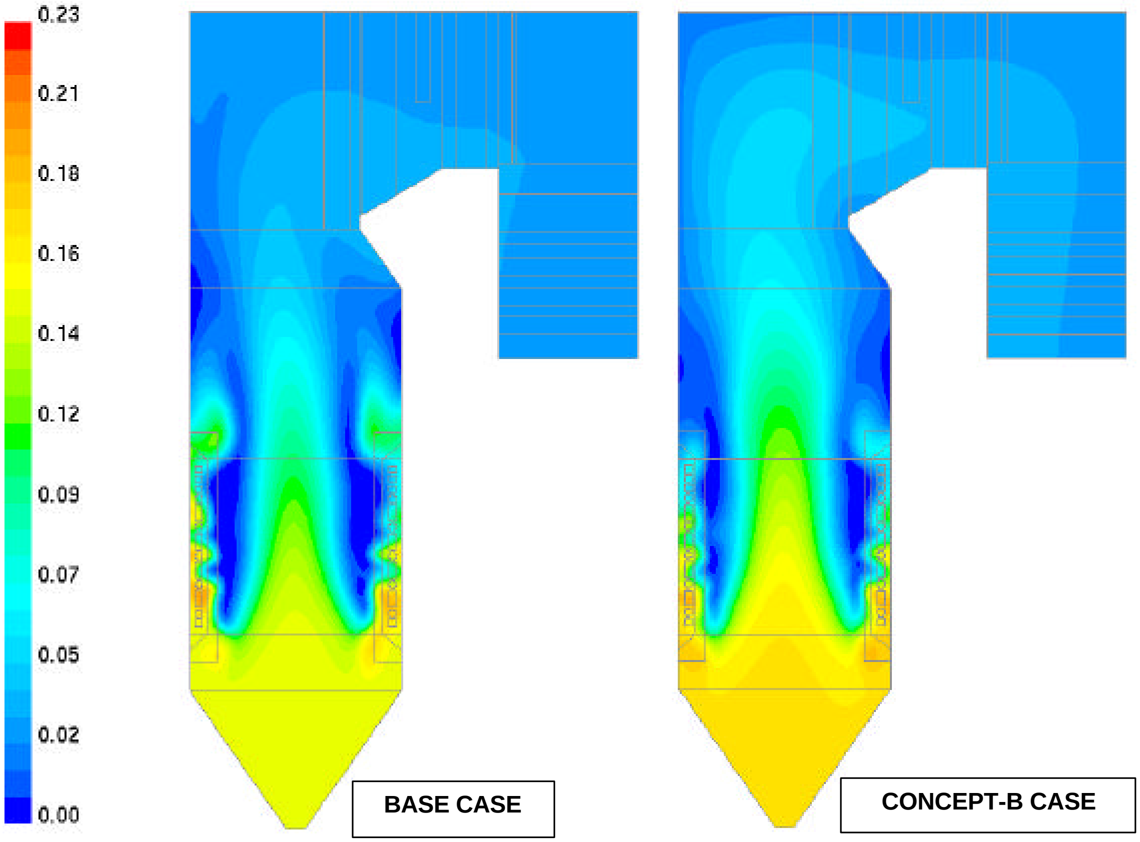


Figure 3. 15: Planar Contour Plots of Temperature at the X-Mid-Plane for the Base Case and Concept-B Case

The corresponding contour plots of the oxygen mass fraction (wet) at the x-mid-plane are shown in Figure 3.16.



1.1.1.1.1.1.9 MASS

Figure 3. 16: Contour Plots of O_2 Mass Fraction (Wet) at the X-Mid-Plane for the Base Case and Concept-B Case

The Concept-B case appears to have a slightly greater concentration of oxygen in the coutant region and in the furnace core than the Base case does. The amount of oxygen will be partially controlled by the relative local concentrations of the particles and the gas, as well as the particle burning rates and trajectory history. Assuming complete combustion, the exit O_2 concentrations on a mass fraction basis are almost equivalent (see Table 3.2); on a molar basis, the Concept-B case will have a higher exit mole fraction (3.6%, dry) compared to that of the Base case (2.6%, dry). Since the composite kinetic/diffusion surface reaction rate model for char oxidation in FLUENT™ is a direct function of the partial pressure of oxygen, assuming an equivalent particle distribution, the higher oxygen mole fraction in the Concept-B case should help the particles to burn out more completely than the particles in the Base case (as shown in Table 3.9).

Figure 3.17 shows the z-component (i.e., the upward component) of velocity on the mid-plane.

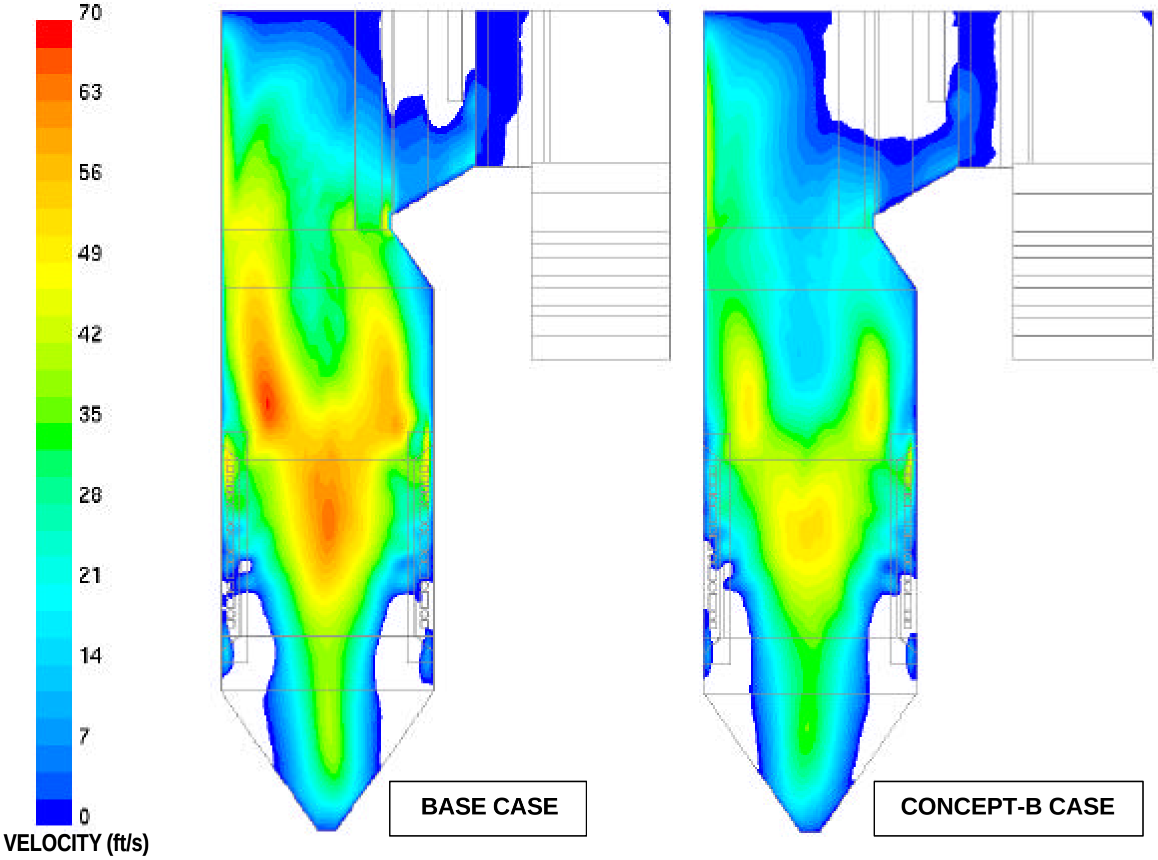


Figure 3. 17: Contour Plots of the (Upward) Z-Velocity at the X-Mid-Plane for the Base Case and Concept-B Case

In Figure 3.17, only the positive (upward) z-component velocity is shown. The upward velocity appears to be somewhat greater in the core of the furnace for the baseline case than for the Concept-B case. For an equivalent upward mass flow rate, at any cross-section, it would be expected that the baseline case velocity would be higher than that of the Concept-B case due simply to the difference in molecular weights (between CO_2 and N_2) and the associated densities. In addition, the baseline case has higher peak temperatures which would also accelerate the local velocity field. However, the presence of a higher velocity does not necessarily mean that the integrated planar flow rates are higher. The CFD results also show that between the bottom of the coutant and the top of the windbox, the planar integrals (not shown) of negative mass flow rate, positive mass flow rate, and absolute mass flow rate are on the order of 10% greater for the Concept-B case than for the baseline case. Above the windbox, the differences in integrated mass flow rate become small (since they simply reflect the nearly equivalent overall inlet flow rates).

Both cases evidence the presence of an adverse pressure gradient (i.e., low velocity region) in the core between the arch-bendline elevation and the top-of-the-windbox elevation. However, the higher upward core velocities associated with the baseline case appear to almost overcome the low-velocity, adverse pressure gradient region. Planar integrals of the swirl number (not shown) indicate that the Concept-B case has a slightly higher swirl magnitude at all elevations except over the upper windbox region. As a result, the Concept-B case appears to be closer to incipient reverse flow zone formation than the baseline case in the upper core region below the arch bendline.

The net wall heat flux is shown in Figure 3.18. (It should be noted that the wall absorption of greatest magnitude is represented by the blue color (i.e., value on the order of –70,000).)

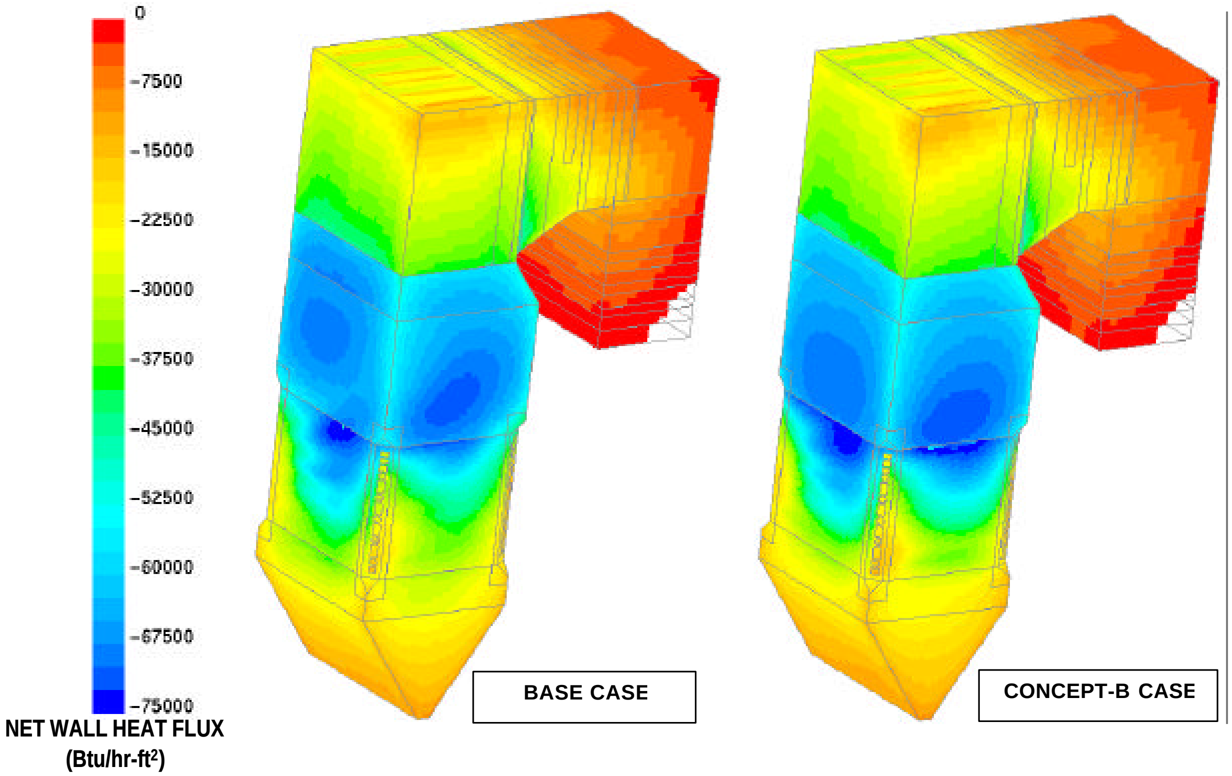


Figure 3. 18: Net Wall Heat Flux for the Base Case and Concept-B Case

As discussed earlier, the wall absorption is very similar for both cases. The peak net wall heat flux is between the top coal nozzle elevation and the arch bendline, in the mid-span region of the waterwall.

3.9 Conclusions

Two cases have been compared using the FLUENT™ CFD code -- a baseline, nominal 15% excess air case (Case 1 or Base case), and a recirculated flue gas, oxy-fired case (Case 2 or Concept-B case). In the Concept-B case, the nitrogen was largely replaced by carbon dioxide. The cases have been compared on the basis of equivalent nozzle/inlet areas, as well as an equivalent coal-to-oxygen feed ratio (by mass). The following computational results have been highlighted:

- The baseline case exhibits higher carbon-in-ash (by 1.6 percentage points), higher outlet NOx emissions (by a factor of 2), and higher outlet CO (by a factor of 2) than the Concept-B case. These same computational trends are also qualitatively exhibited by the bench-scale testing.
- The baseline case exhibits a slightly higher peak gas temperature (maximum difference of about 200 °F), and a correspondingly higher average (cross-sectional) gas temperature (difference of 90 °F at the HFOT), than that of the Concept-B case.
- The net wall absorption in the furnace region for the baseline case is larger (by less than 1%) than that of the Concept-B case. (Conversely, the net wall absorption for the baseline calculation with the RHBP was found to be about 6% lower than the Concept-B case calculated by the RHBP.)

Explanations for the observed trends must rely on physical property differences between the nitrogen, which dominates in the baseline case, and carbon dioxide, which largely supplants the nitrogen in the Concept-B case. The physical species properties and the anticipated consequences are itemized below:

(1) Higher molecular weight of CO_2 relative to N_2. For equivalent inlet areas, and as a result of the molecular weight and associated density differences, the baseline case will exhibit larger inlet velocities and a higher penetration capability than the Concept-B case. The inlet velocity differences will further serve to impact the aerodynamic, particle trajectory, mixing, and flame patterns.

(2) Higher specific heat of CO_2 relative to N_2. For an equivalent local heat release and heat transfer, the higher specific heat of carbon dioxide will serve to decrease the overall flame temperatures of the recycled flue gas mixture in the Concept-B case, relative to the baseline case.

(3) Higher gas emissivity and absorptance of CO_2 relative to N_2. For an equivalent local gas temperature, the higher emissivity of CO_2, relative to the optically transparent N_2, should serve to increase the absorption coefficient (and the irradiation to the wall) in the recycled flue gas case (Concept-B) relative to the baseline case.

The three physical properties listed above may have compensatory rather than cumulative effects. In the present CFD calculations, the resultant, local gas temperatures are not equivalent in the two cases. The higher specific heat of the carbon dioxide, and the associated lower gas temperatures in the Concept-B case, will tend to offset somewhat or compensate for the expected increase in the wall absorption (i.e., anticipated due to the enhanced emissivity of the CO_2.) Furthermore, the spatial distribution of the carbon dioxide across the cross-section may be important; it is hypothesized that high concentrations of CO_2 in the cooler gas mixtures between the wall and the flame may actually act to inhibit the net radiative flux to the walls.

The trend calculated by the CFD code, with respect to the furnace temperature at the HFOP, appears to be qualitatively aligned with the trend calculated by the in-house RHBP code. However, the difference in wall absorption for the Concept-B case relative to the baseline case, calculated by FLUENT™ (change of –1%), does not fully agree with the results of the RHBP (change of +6%). Computed and/or experimental trends that indicate that oxy-firing may induce either an increase in wall absorption (+6%) or a decrease in wall absorption (-18%), relative to the baseline case, are supported in the literature. It is suggested that the absorptive properties of a local CO_2-containing gas mixture in the vicinity of the wall may either act to enhance or inhibit the incident radiative flux to the wall, depending upon its relative spatial placement and mixture temperature. For example, a low-temperature, CO_2-rich mixture near the wall would absorb the radiative flux from the interior “fireball” region; a portion of the energy would be re-emitted at the lower gas temperature, and the remainder of the energy would be used to heat up the gas. The result would be a net reduction in the incident radiative flux to the wall. Conversely, a high-temperature, CO_2-rich mixture near the wall would emit at its local gas temperature, thus enhancing the incident radiative flux to the wall. Therefore, it is suggested that the relative spatial relationships of the cool and hot gas mixtures, the relative composition of the absorbing media, and the proximity of that media to the wall (as induced by the furnace and firing system aerodynamics), may significantly impact the gas-phase irradiation to the wall. Indeed, the aerodynamic and flame patterns that prevail in a given air-fired or oxy-fired furnace may conceivably have as strong an impact on the net wall absorption as the physical property differences themselves (gas-phase emissivity, specific heat, etc.)

The results of computational tools (such as CFD and the RHBP) should be viewed as being qualitative in nature, particularly when they are asked to produce calculations outside of the experience base for which they have been validated. Certainly, the gas-phase and particle-phase radiation property submodels could be enhanced in both codes, and the impact of spatial gradients (caused by the interleaving of hot and cool regions of varying CO_2 composition) on radiative wall flux needs to be investigated more fully. CFD computations have been utilized to gain a more in-depth qualitative understanding of aerodynamic and flame patterns in combustion systems, but have not yet been integrated into routine boiler design procedures. At the present time, the ALSTOM design standards are based upon the experience and expertise built into the RHBP and other design protocols, and those standards must be adhered to in any new design project in order to mitigate risk. Therefore, the results of the RHBP must be regarded as the default standard, both from an engineering experience viewpoint and from a rules-based design viewpoint. However, additional validation work needs to be done in order to confirm the trends for oxy-firing scenarios.

Although differences in physical properties of the transport gases (e.g., air-firing versus oxy-firing) may initially induce undesirable deviations from the performance goals of a field unit, the designer and engineer views such alterations as challenges rather than insurmountable obstacles. Much can be done in the way of operating condition optimization (e.g., spray and nozzle tilt), without major retrofits, to realign steam temperatures and wall absorptions with their desired, target values. Design parameters are often in conflict, and the intelligent designer must work to balance various parameters and operating conditions to achieve the desired outcome. The CFD work performed here was based on certain constraints to maintain equivalency for comparison purposes, which would not necessarily be the case in the field. For example, a potential decrease in furnace wall absorption of several percentage points could be nullified through manipulation of various operating parameters. If required, firing system retrofits and surface modifications could be performed to handle more severe decreases in absorption; certainly new unit design could easily accommodate any necessary alterations in design philosophy.

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