NATIONAL ENERGY TECHNOLOGY LABORATORY

# Cost and Performance Baseline for Fossil Energy Plants
# Volume 2: Coal to Synthetic Natural Gas and Ammonia

July 5, 2011

**DOE/NETL- 2010/1402**

U.S. DEPARTMENT OF ENERGY

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# COST AND PERFORMANCE BASELINE FOR FOSSIL ENERGY PLANTS VOLUME 2: COAL TO SYNTHETIC NATURAL GAS AND AMMONIA

**DOE/NETL-2010/1402**

**Final Report**

**July 5, 2011**

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## LIST OF ACRONYMS AND ABBREVIATIONS

| | |
|---|---|
| AACE | Association for the Advancement of Cost Engineering |
| AEO | Annual Energy Outlook |
| AF | Availability factor |
| AFUDC | Allowance for funds used during construction |
| AGR | Acid gas removal |
| Aspen | AspenPlus® |
| ASU | Air separation unit |
| BEC | Bare erected cost |
| BFD | Block flow diagram |
| BFP | Boiler feed pump |
| BFW | Boiler feedwater |
| Btu | British thermal unit |
| Bscf | Billion standard cubic foot |
| Btu/lb | British thermal unit per pound |
| Btu/scf | British thermal unit per standard cubic foot |
| $C_4$+ | Butane + higher paraffins |
| CCF | Capital Charge Factor |
| CCPI | Clean Coal Power Initiative |
| CCS | Carbon capture and sequestration |
| CFBC | Circulating fluidized bed combustor |
| cfm | Cubic feet per minute |
| CL | Closed-loop |
| cm | Centimeter |
| CO | Carbon monoxide |
| $CO_2$ | Carbon dioxide |
| CoP | Conoco Phillips |
| COP | Cost of production |
| COS | Carbonyl sulfide |
| CRT | Cathode ray tube |
| CS | Carbon steel |
| $CS_2$ | Carbon disulfide |
| CT | Combustion turbine |
| CWP | Circulating water pump |
| CWS | Circulating water system |
| CWT | Cold water temperature |
| DCS | Distributed control system |
| DI | De-ionized |
| DOE | Department of Energy |
| EIA | Energy Information Administration |
| EM | Electromagnetic |
| EOR | Enhanced oil recovery |
| EPA | Environmental Protection Agency |

| | |
|---|---|
| EPC | Engineer/Procure/Construct |
| EPRI | Electric Power Research Institute |
| EPCM | Engineering/Procurement/Construction Management |
| FBG | Fluidized bed gasifier |
| FEED | Front-End Engineering Design |
| ESPA | Energy Sector Planning and Analysis |
| FERC | Federal Energy Regulatory Commission |
| FGT | Florida gas transmission |
| FOAK | First-of-a-kind |
| ft | Foot, Feet |
| FW | Feedwater |
| FYCOP | First year cost of production |
| gal | Gallon |
| GDP | Gross domestic product |
| GEE | General Electric Energy |
| GHG | Greenhouse gas |
| GJ/hr | Gigajoule per hour |
| gpm | Gallons per minute |
| GRE | Great River Energy |
| $H_2$ | Hydrogen |
| $H_2S$ | Hydrogen sulfide |
| Hg | Mercury |
| HHV | Higher heating value |
| hp | Horsepower |
| HP | High-pressure |
| HVAC | Heating, ventilating, and air conditioning |
| HWT | Hot water temperature |
| Hz | Hertz |
| ICR | Information Collection Request |
| IDGCC | Integrated Drying and Gasification Combined Cycle |
| IGCC | Integrated gasification combined cycle |
| IP | Intermediate-pressure |
| kg/GJ | Kilogram per gigajoule |
| kg/hr | Kilogram per hour |
| kJ/Nm$^3$ | Kilojoule per normal cubic meter |
| kJ/kg | Kilojoule per kilogram |
| km | Kilometer |
| KO | Knockout |
| kV | Kilovolt |
| kW, kWe | Kilowatts electric |
| kWth | Kilowatt thermal |
| lb/hr | Pound per hour |
| lb/MMBtu | Pound per million British thermal units |

| | |
|---|---|
| LCOP | Levelized cost of production |
| LF | Levelization Factor |
| LNG | Liquefied natural gas |
| LHV | Lower heating value |
| LP | Low-pressure |
| lpm | Liters per minute |
| m | Meters |
| $m^3$/min | Cubic meter per minute |
| md | Millidarcy |
| MMBtu | Million British thermal units (also shown as $10^6$ Btu) |
| MMBtu/hr | Million British thermal units (also shown as $10^6$ Btu) per hour |
| MMkJ | Million kilojoules (also shown as $10^6$ kJ) |
| MMscf | Million standard cubic feet |
| $MNm^3$ | Thousand normal cubic meters |
| $BNm^3$ | Billion normal cubic meters |
| mol% | Mole percent |
| MPa | Megapascal |
| MVA | Mega volt-amp |
| MW | Megawatt |
| MWh | Megawatt-hour |
| MWth | Megawatts thermal |
| N/A | Not applicable |
| NDL | North Dakota Lignite |
| NETL | National Energy Technology Laboratory |
| NGCC | Natural gas combined cycle |
| $NH_3$ | Ammonia |
| $Nm^3$ | Normal cubic meter |
| $Nm^3$/min | Normal cubic meter per minute |
| NOAK | $N^{th}$-of-a-kind |
| No. | Number |
| NOx | Oxides of nitrogen |
| O eqv | Oxygen equivalents |
| O&GJ | Oil and Gas Journal |
| O&M | Operation and maintenance |
| $O_2$ | Oxygen |
| $OC_{Fn}$ | Category n fixed operating cost for the initial year of operation |
| PAS | PAS, Inc. |
| PC | Pulverized coal |
| PM | Particulate matter |
| POTW | Publicly Owned Treatment Works |
| ppm | Parts per million |
| ppmv | Parts per million volume |
| PRB | Powder River Basin |

| | |
|---|---|
| PSA | Pressure Swing Adsorption |
| PSFM | Power Systems Financial Model |
| psi | Pound per square inch |
| psia | Pound per square inch absolute |
| psig | Pound per square inch gage |
| R&D | Research and development |
| RDS | Research and Development Solutions, LLC |
| ROE | Return on equity |
| RWE | Rheinisch-Wesfälisches Elektrizitatswerk (German company) |
| SC | Supercritical |
| scf | Standard cubic feet |
| scfm | Standard cubic feet per minute |
| SCOT | Shell Claus Off-gas Treating |
| SFG | Siemens Fuel Gasification |
| SG | Specific gravity |
| SGC | Synthesis gas cooler |
| SGS | Sour gas shift |
| SNG | Synthetic Natural Gas |
| $SO_2$ | Sulfur dioxide |
| SRU | Sulfur recovery unit |
| STG | Steam turbine generator |
| Syngas | Synthesis gas |
| TASC | Total as-spent cost |
| TGTU | Tail gas treating unit |
| TOC | Total overnight cost |
| Tonne | Metric Ton (1000 kg) |
| TPC | Total plant cost |
| TPH | Ton per hour |
| TS&M | Transport, storage, and monitoring |
| USDA | United States Department of Agricultural |
| vol% | Volume percent |
| WB | Wet bulb |
| WGS | Water gas shift |
| wt% | Weight percent |
| WTA | German acronym for “fluidized bed dryer with integrated waste heat recovery” |
| ZnO | Zinc oxide |
| $/GJ | Dollars per gigajoule |
| $/MMBtu | Dollars per million British thermal units |
| $/MMkJ | Dollars per million kilojoule |
| $/ton | Dollars per ton |
| °C | Degrees Celsius |
| °F | Degrees Fahrenheit |
| 5-10s | 50-hour work-week |

## EXECUTIVE SUMMARY

The objective of this report is to present an accurate, independent assessment of the cost and performance of bituminous, subbituminous, and lignite coal conversion systems, specifically for the production of synthetic natural gas (SNG) and ammonia, using a consistent economic approach that accurately reflects current market conditions for plants starting operation in the near term. This is Volume 2 of a four volume report. The four volume series consists of the following:

- Volume 1: Bituminous Coal and Natural Gas to Electricity
- Volume 2: Coal to Synthetic Natural Gas and Ammonia (Various Coal Ranks)
- Volume 3: Low Rank Coal and Natural Gas to Electricity
- Volume 4: Bituminous Coal to Liquid Fuels with Carbon Capture

Volume 1 entitled *Cost and Performance Baseline for Fossil Energy Power Plants Study, Volume 1: Bituminous Coal and Natural Gas to Electricity* was recently revised and re-released in November 2010. Volume 3 was published in 2011 and Volume 4 is scheduled for publication later in 2011.

The cost and performance of a coal-to-SNG plant, with and without ammonia co-production, were evaluated in this study. The results are expected to be valuable to policy makers and will help guide research and development (R&D) efforts necessary to make SNG more cost competitive. Selection of new technologies will depend on many factors, including:

- Capital and operating costs;
- Overall conversion efficiency;
- Fuel prices;
- Cost of production (COP);
- Availability, reliability, and environmental performance; and
- Current and potential regulation of air, water, and solid waste discharges from fossil fuel-based production facilities, including a potential cap or tax on carbon dioxide ($CO_2$) emissions.

Eight coal-based plants using the Siemens gasifier were analyzed as listed in Exhibit ES-1. The list includes six SNG production cases using various coal ranks and two SNG and ammonia co-production cases using bituminous coal only. All cases require carbon capture prior to the methanation reactors and the ammonia ($NH_3$) plant in co-production cases. In some cases the removed $CO_2$ is sequestered, and in other cases it is released to the atmosphere. Normally cases are designated as $CO_2$ capture and non-capture, but since all cases are capture in this study, the designation used to differentiate between cases is sequestration and non-sequestration. Each specific scenario is presented without and with carbon sequestration.

While input was sought from various technology vendors, the final assessment of performance and cost was determined independently, and may not represent the views of the technology vendors. The extent of collaboration with the technology vendors varied in scope for each of the processes within the plant, with minimal or no collaboration obtained from some vendors.

The gasification scheme used in this study includes a partial quench followed by a syngas cooler. Heat recovery was necessary to produce enough power to satisfy internal auxiliary loads with some excess power for export in most cases. This configuration is not currently a commercial offering by Siemens, but based on personal communications, it is a configuration they are planning to offer for future coal-to-SNG projects.

The methodology included performing steady-state simulations of the various technologies using the AspenPlus® (Aspen) modeling program. The resulting mass and energy balance data from Aspen were used to size major pieces of equipment. These equipment sizes formed the basis for cost estimating. Performance and process limits were based upon published reports, information obtained from vendors and users of the technology, performance data from design/build projects, and/or best engineering judgment. Capital and operating costs were estimated based on simulation results and through a combination of vendor quotes and scaled estimates. Baseline fuel costs for this analysis were determined using data from the Energy Information Administration's (EIA) 2008 Annual Energy Outlook (AEO). The first year of capital expenditure (2007) costs used are 1.55 dollars per gigajoule ($/GJ) (1.64 dollars per million British thermal units [$/MMBtu]) for Illinois No. 6 coal, $0.84/GJ ($0.89/MMBtu) for Powder River Basin (PRB) coal, and $0.79/GJ ($0.83/MMBtu) for North Dakota lignite (NDL). All coal prices are on a higher heating value (HHV) basis and in 2007 United States (U.S.) dollars. The Illinois No. 6 and PRB coal prices include transportation costs while the NDL coal is assumed to be used at a minemouth location. Anhydrous ammonia selling prices were determined from the historic correlation between the cost of industrial natural gas (EIA) and the price of ammonia fertilizers as paid by U.S. farmers (United States Department of Agriculture [USDA] Economic Research Service). The first year selling price of ammonia for the non-sequestration SNG and ammonia co-production facility was determined to be $880/metric ton (tonne) ($799/ton) and $913/tonne ($828/ton) with carbon sequestration. All plant configurations are evaluated on installation at a greenfield site. SNG production is limited by a nominal thermal input of 500 megawatt-thermal ($MW_{th}$) to the gasifier. Based on vendor input, the actual gasifier thermal input varies according to the type of coal feed and ranges from 506 to 550 MWth. The nominal net plant production capacity of ammonia is 2,000 tonnes/day (2,204 tons/day) for the co-production cases. All cases are based on a 90 percent plant capacity factor (CF). Exhibit ES-2 shows the cost, performance, and environmental profile summary for all cases. These plants are expected to operate at full capacity when available. Since variations in fuel costs, capacity factor, and financing structure (required return on equity) can influence the economics of production, sensitivities of the first year cost of production (FYCOP) are evaluated and presented later in this Executive Summary.

**Exhibit ES-1 Case Descriptions**

| Case | Product(s) | Steam Cycle, psig/°F/°F[1] | Coal Type | Coal Drying Process | Gasifier Technology | Oxidant | $H_2S$ Separation/ Removal | Sulfur Removal/ Recovery | $CO_2$ Separation | $CO_2$ Compression/ Sequestration |
|---|---|---|---|---|---|---|---|---|---|---|
| 1 | SNG | 1800/1050/1000 | Illinois No. 6 | Standard | Siemens | 99 mol% $O_2$ | Selexol | Claus Plant | Selexol 2nd stage | N |
| 2 | SNG | 1800/1050/1000 | Illinois No. 6 | Standard | Siemens | 99 mol% $O_2$ | Selexol | Claus Plant | Selexol 2nd stage | Y |
| 3 | SNG & Ammonia | 1800/1050/1000 | Illinois No. 6 | Standard | Siemens | 99 mol% $O_2$ | Selexol | Claus Plant | Selexol 2nd stage | N |
| 4 | SNG & Ammonia | 1800/1050/1000 | Illinois No. 6 | Standard | Siemens | 99 mol% $O_2$ | Selexol | Claus Plant | Selexol 2nd stage | Y |
| 5 | SNG | 1800/1050/1000 | Montana Rosebud PRB | WTA | Siemens | 99 mol% $O_2$ | Selexol | Claus Plant | Selexol 2nd stage | N |
| 6 | SNG | 1800/1050/1000 | Montana Rosebud PRB | WTA | Siemens | 99 mol% $O_2$ | Selexol | Claus Plant | Selexol 2nd stage | Y |
| 7 | SNG | 1800/1050/1000 | North Dakota Lignite | WTA | Siemens | 99 mol% $O_2$ | Selexol | Claus Plant | Selexol 2nd stage | N |
| 8 | SNG | 1800/1050/1000 | North Dakota Lignite | WTA | Siemens | 99 mol% $O_2$ | Selexol | Claus Plant | Selexol 2nd stage | Y |

[1] Steam cycle reheat temperatures are nominal.

**Exhibit ES-2 Cost and Performance Summary and Environmental Profile**

| | Illinois No. 6 Coal | | | | Rosebud PRB | | North Dakota Lignite | |
|---|---|---|---|---|---|---|---|---|
| | SNG Only | | SNG and Ammonia | | SNG Only | | SNG Only | |
| **PERFORMANCE** | Case 1 | Case 2 | Case 3 | Case 4 | Case 5 | Case 6 | Case 7 | Case 8 |
| $CO_2$ Sequestration | No | Yes | No | Yes | No | Yes | No | Yes |
| SNG Production (Bscf/year)[1] | 57 | 56 | 42 | 42 | 55 | 55 | 52 | 52 |
| Ammonia Production (TPD)[1] | N/A | N/A | 2,204 | 2,204 | N/A | N/A | N/A | N/A |
| HHV Conversion Efficiency, % | 61.4% | 61.3% | 61.5% | 61.4% | 63.1% | 63.1% | 61.5% | 61.5% |
| Gross Power Output ($kW_e$) | 308,000 | 310,600 | 292,300 | 292,300 | 302,000 | 302,000 | 305,800 | 310,500 |
| Auxiliary Power Requirement ($kW_e$) | 216,350 | 262,090 | 263,600 | 316,030 | 247,570 | 300,190 | 259,740 | 313,930 |
| Net Power Output ($kW_e$) | 91,650 | 48,510 | 28,700 | -23,730 | 54,430 | 1,810 | 46,060 | -3,430 |
| Coal Flowrate (lb/hr) | 964,752 | 964,752 | 964,000 | 964,000 | 1,259,331 | 1,259,331 | 1,564,932 | 1,564,932 |
| HHV Thermal Input ($kW_{th}$) | 3,298,455 | 3,298,455 | 3,295,885 | 3,295,885 | 3,160,745 | 3,160,745 | 3,034,796 | 3,034,796 |
| Raw Water Withdrawal (gpm) | 7,169 | 7,123 | 7,434 | 7,458 | 4,853 | 5,509 | 4,421 | 5,131 |
| Process Water Discharge (gpm) | 1,462 | 1,451 | 1,455 | 1,461 | 1,071 | 1,218 | 1,071 | 1,230 |
| Raw Water Consumption (gpm) | 5,708 | 5,672 | 5,979 | 5,997 | 3,783 | 4,291 | 3,350 | 3,901 |
| $CO_2$ Emissions (lb/MMBtu)[2] | 128 | 5 | 147 | 10 | 140 | 0.7 | 147 | 0.9 |
| $SO_2$ Emissions (lb/MMBtu)[2] | 0.0274 | 0.0003 | 0.0229 | 0.0006 | 0.0218 | 0.0000 | 0.0239 | 0.0000 |
| NOx Emissions (lb/MMBtu)[2] | Negligible | Negligible | Negligible | Negligible | Negligible | Negligible | Negligible | Negligible |
| PM Emissions (lb/MMBtu)[2] | 0.007 | 0.007 | 0.007 | 0.007 | 0.007 | 0.007 | 0.007 | 0.007 |
| Hg Emissions (lb/MMBtu)[2] | 5.71E-07 | 5.71E-07 | 5.71E-07 | 5.71E-07 | 3.51E-07 | 3.51E-07 | 5.60E-07 | 5.60E-07 |
| **COST** | | | | | | | | |
| Total Plant Cost ($ x 1,000) | 2,628,754 | 2,692,997 | 3,048,463 | 3,119,611 | 2,741,044 | 2,813,258 | 2,864,478 | 2,941,335 |
| Total Overnight Cost ($ x 1,000) | 3,235,262 | 3,312,740 | 3,742,411 | 3,829,817 | 3,354,442 | 3,442,310 | 3,504,262 | 3,595,468 |
| Total As-spent Capital ($ x 1,000) | 3,885,549 | 3,978,601 | 4,494,636 | 4,599,610 | 4,028,684 | 4,134,215 | 4,208,619 | 4,318,158 |
| SNG FYCOP ($/MMBtu)[1] | 19.27 | 20.95 | 15.82 | 17.85 | 19.15 | 21.01 | 21.27 | 23.24 |
| *$CO_2$ TS&M Costs* | *0.00* | *0.91* | *0.00* | *1.29* | *0.00* | *0.96* | *0.00* | *1.03* |
| *Fuel Costs* | *2.67* | *2.67* | *1.89* | *1.89* | *1.41* | *1.41* | *1.34* | *1.35* |
| *Variable Costs* | *1.03* | *1.04* | *0.79* | *0.82* | *1.02* | *1.03* | *1.18* | *1.18* |
| *Fixed Costs* | *1.76* | *1.79* | *1.41* | *1.45* | *1.84* | *1.86* | *2.03* | *2.06* |
| *Electricity Costs* | *-0.77* | *-0.41* | *-0.17* | *0.15* | *-0.46* | *-0.02* | *-0.42* | *0.03* |
| *Capital Costs* | *14.58* | *14.94* | *11.90* | *12.25* | *15.34* | *15.76* | *17.13* | *17.59* |
| Ammonia FYCOP[3] ($/ton) | - | - | 799 | 828 | - | - | - | - |

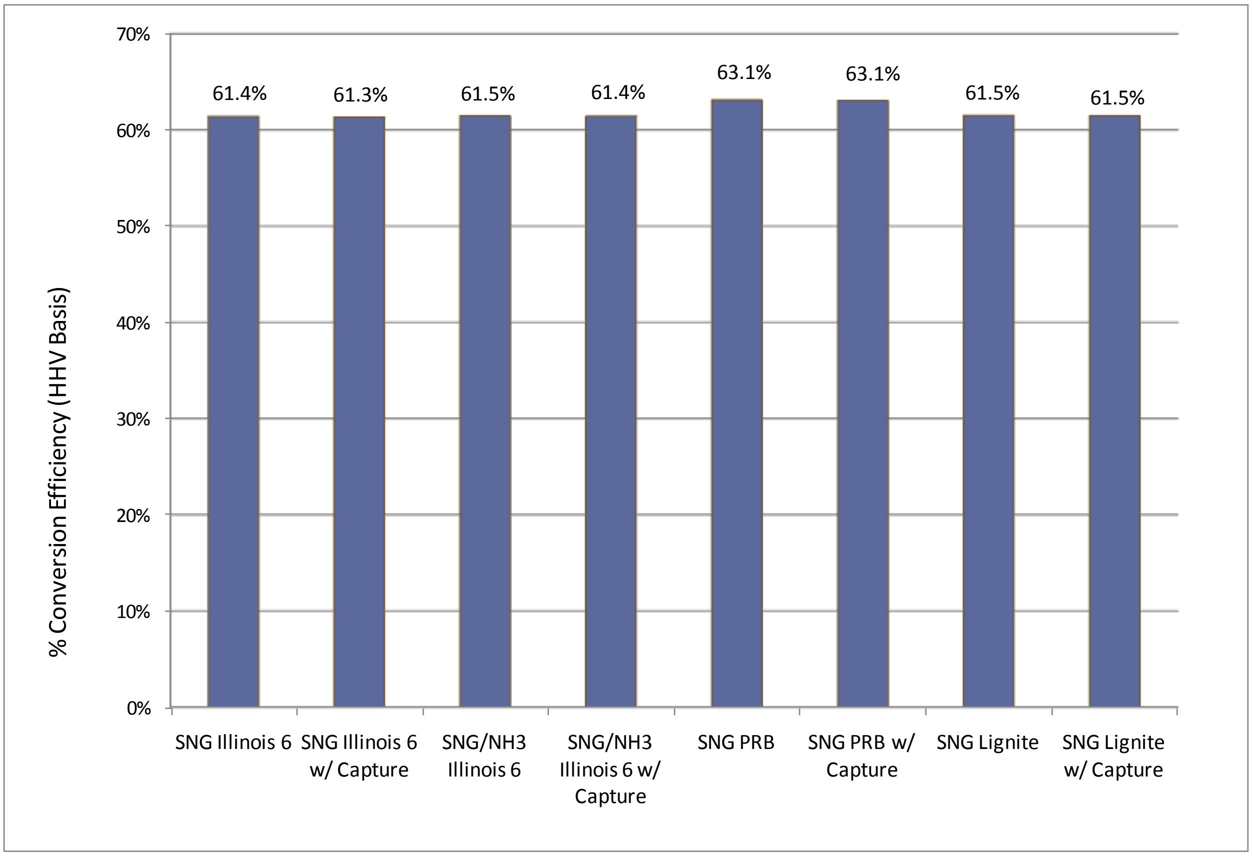
1 Based on a capacity factor of 90 percent for all cases
2 Based on coal thermal input
3 Ammonia price is correlated to historic natural gas costs

## CONVERSION EFFICIENCY

The net plant conversion efficiency (HHV basis) for all eight cases is shown in Exhibit ES-3. The conversion efficiency is the HHV of the products divided by the HHV of the coal. The primary conclusions that can be drawn are:

- The SNG production facilities feeding PRB coal have the highest conversion efficiencies of the coal types modeled in this study with a conversion efficiency of 63.1 percent for both the non-sequestration and carbon sequestration cases. This is primarily due to the low nitrogen content and high oxygen content in the design fuel, which enables the SNG produced to have a relatively lower concentration of inerts.
- The remaining six cases have essentially constant conversion efficiency (61.3 to 61.5 percent). The addition of carbon sequestration has only a minimal impact on conversion efficiency. The carbon sequestration impact is seen in the reduced available export power relative to the non-sequestration case.

**Exhibit ES-3 Net Plant Conversion Efficiency (HHV)**




## TOTAL OVERNIGHT COST

The Total Overnight Cost (TOC) for each plant was calculated by adding owner's costs to the Total Plant Cost (TPC). The TPC was determined through a combination of vendor quotes and scaled estimates from previous design/build projects. TPC includes all equipment (complete with initial chemical and catalyst loadings), materials, labor (direct and indirect), engineering and construction management, and contingencies (process and project).

The capital costs have an estimated accuracy of -15%/+30%, consistent with the screening study level of design engineering applied to the various cases in the study. The value of the study lies not in the absolute accuracy of the individual cases, but in the fact that all cases were evaluated under the same set of technical and economic assumptions. The consistency of approach allows meaningful comparisons among the cases evaluated.

Project contingencies were added to the Engineering/Procurement/Construction Management (EPCM) capital accounts to cover project uncertainty and the cost of any additional equipment that would result from a detailed design. The contingencies represent costs that are expected to occur. Each bare erected cost (BEC) account was evaluated against the level of estimate detail and field experience to determine project contingency. Process contingency was added to cost account items that were deemed to pose significant risk due to lack of operating experience. The cost accounts that received a process contingency include:

- Gasifiers and Syngas Coolers – 15 percent on all cases – next-generation commercial offering and integration with the steam generation island
- Two Stage Selexol – 20 percent on all cases - unproven technology at commercial scale
- Mercury Removal – 5 percent on all cases – minimal commercial scale experience in gasification applications
- Methanation – 10 percent on all cases – unproven technology at commercial scale
- Instrumentation and Controls – 5 percent on all accounts for integration issues

The TOC is shown for each plant configuration in Exhibit ES-4. For reference, Total As-spent Capital (TASC) is also shown. TASC is the sum of TOC, escalation during the capital expenditure period, and interest on debt during the capital expenditure period. The following observations can be made:

- The SNG only case without carbon sequestration using Illinois No. 6 coal has the lowest TOC of $3.24 billion.
- For the SNG only cases without carbon sequestration, the TOC is higher for PRB coal by 3.7 percent and lignite coal by 8.3 percent relative to the Illinois No. 6 coal cases.
- The TOC increase to add carbon sequestration for the SNG only cases is approximately 2.4 percent for the Illinois No. 6 cases, 2.6 percent for the PRB coal cases and 2.6 percent for the lignite coal case. For the SNG and ammonia co-production case, the TOC increases by 2.4 percent to add carbon sequestration.
- The SNG and ammonia co-production cases increase the TOC by approximately 16 percent when compared to the SNG only cases using Illinois No. 6 coal.

## FIRST YEAR COST OF PRODUCTION

The cost metric used in this study is the first year cost of production (FYCOP). The capital charge factor (CCF) shown in Exhibit ES-5, which was derived using the National Energy Technology Laboratory (NETL) Power Systems Financial Model (PSFM) [9], was used to calculate the FYCOP. The project financial structure is representative of a high-risk fuels project with no loan guarantees or other government subsidies. The parameters were chosen to reflect the next commercial offering in which the technical and economic risks have not been reduced by the wide-scale demonstration of the technology.

**Exhibit ES-4 Total Overnight Cost and Total As-spent Capital**

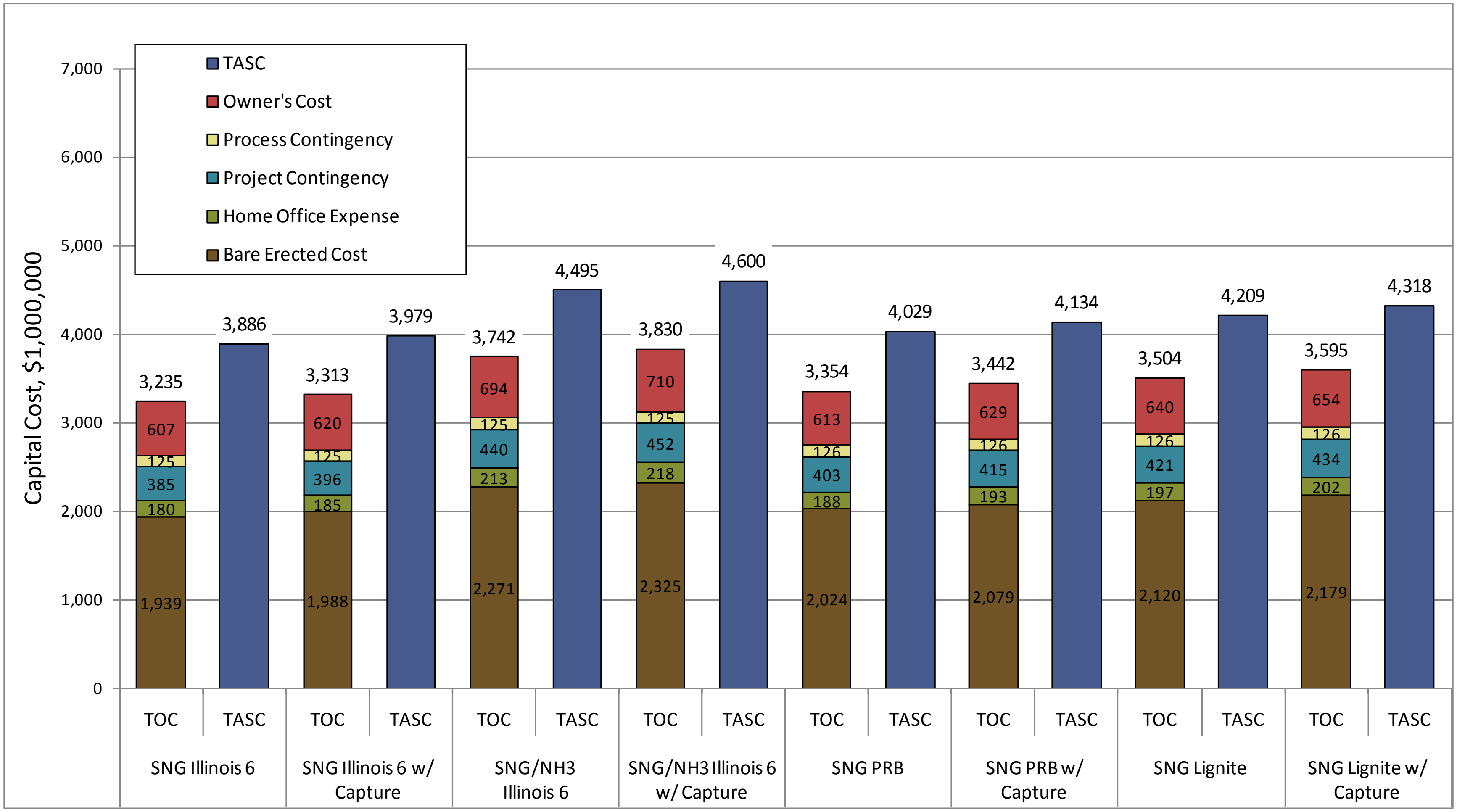


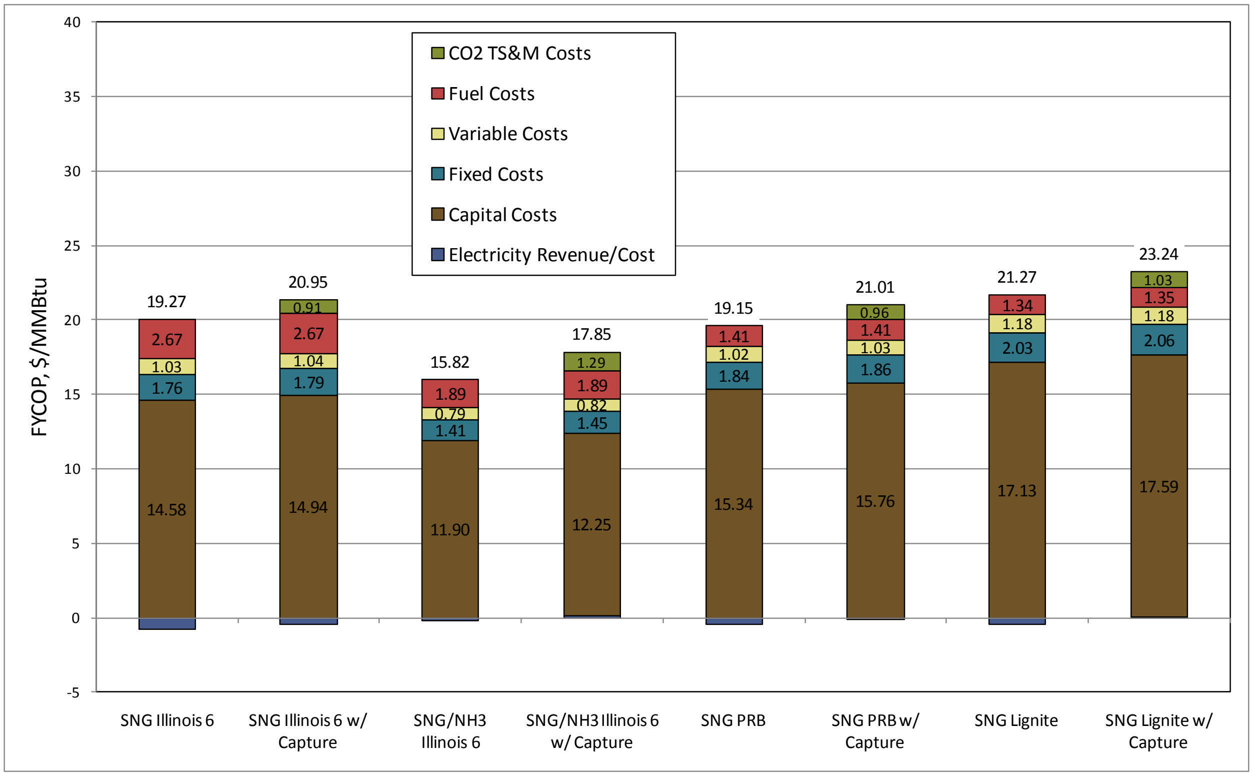
**Exhibit ES-5 Economic Parameters Used to Calculate FYCOP**

| | High Risk Fuels Financial Structure |
|---|---|
| First Year Capital Charge Factor | 0.2449 |

The FYCOP was calculated as detailed in Section 2.8. The FYCOP is shown in Exhibit ES-6 with the capital cost, fixed operating cost, variable operating cost, electricity cost (or revenue), and fuel costs shown separately. In the carbon sequestration cases, the $CO_2$ transport, storage, and monitoring (TS&M) costs are also shown as a separate bar segment. The following conclusions can be drawn, using first year cost as the basis for comparison:

- The co-production case using Illinois No. 6 coal without carbon sequestration has the lowest cost of the cases modeled in this study with an FYCOP of $15.82/MMBtu.
- The FYCOP is dominated by capital charges in all cases. The capital cost component of FYCOP, which includes owner's costs, comprises 71 to 81 percent for the SNG only cases.
- The fuel cost component in the SNG only cases represents 13 percent for Illinois No. 6 coal, 7 percent for the PRB coal, and 6 percent for the lignite coal.
- The TS&M component of FYCOP in the carbon sequestration cases is 4 percent, 5 percent, and 4 percent for the Illinois No. 6 coal, PRB coal, and lignite coal respectively. For the SNG and ammonia co-production case, the TS&M account for 7 percent of the total FYCOP.
- The SNG FYCOP increases by an average amount of 9 percent with the addition of carbon sequestration for the SNG only cases and 13 percent for the co-production case. The cost of TS&M is shared between the two products in the co-production case based on the SNG-$NH_3$ cost relationship.

**Exhibit ES-6 FYCOP by Cost Component**




The EIA's AEO 2009 shows a 2007 residential natural gas delivered price of $12.69/MMBtu. The EIA reference case projections show a price of $12.09/MMBtu in 2010, $12.50/MMBtu in 2020, and $14.31/MMBtu in 2030. The price of SNG determined in this study is generally higher than current or projected natural gas market prices. However, the assumed financing structure (addressed in Section 2.8 of this report) plays a significant role in the elevated prices. The assumed return on equity (ROE) is 20 percent. A sensitivity analysis on ROE was performed and is shown in Exhibit ES-7. As expected, as the ROE requirement increases, the cost of SNG and ammonia increases. At the assumed high-risk fuels scenario first year capital charge factor (CCF) and ROE, SNG prices range from $15.82 to $23.24/MMBtu and ammonia costs range from $799 to $828/ton. Adjusting the first year CCF to 0.1344, representative of a high-risk fuels project with a corresponding ROE of 12 percent, results in SNG FYCOP cost ranging from $9.85 to $14.85/MMBtu and ammonia cost ranging from $560 to $585/ton.

**Exhibit ES-7 FYCOP Sensitivity to ROE**

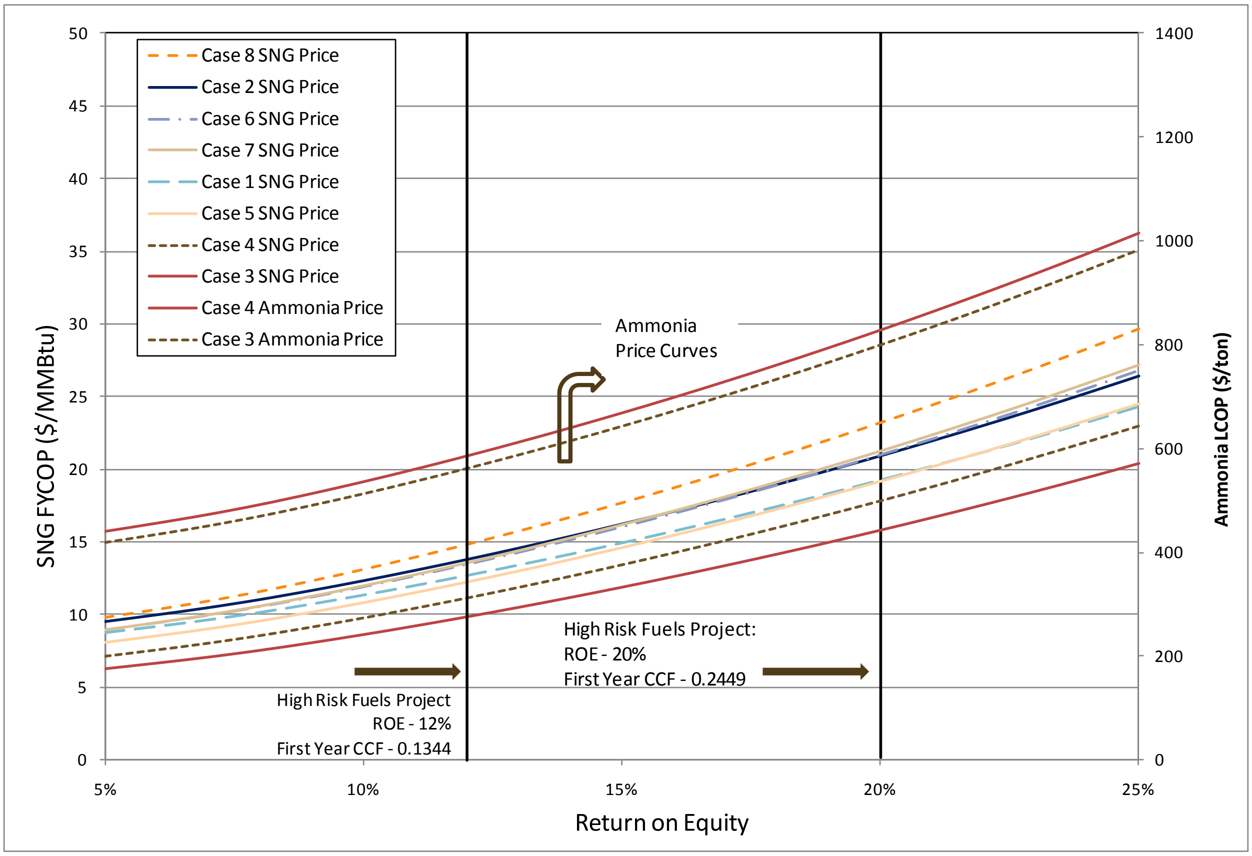



Exhibit ES-8 shows the FYCOP sensitivity to fuel costs. As expected, all cases show a linear increase in FYCOP with the increase in coal price. Fuel cost represents a relatively minor fraction of the FYCOP so that even at a fuel price of zero, the FYCOP of SNG ranges from $13.73 to $21.89/MMBtu.

The sensitivity of FYCOP to CF is shown in Exhibit ES-9. All cases show a decrease in FYCOP with an increase in CF. The CF used in this study was 90 percent. The rate of decrease is relatively small above 90 percent with the maximum differential between 90 percent and 100 percent capacity of $1.96/MMBtu for SNG only cases and $1.52/MMBtu for the co-production cases.

**Exhibit ES-8 FYCOP Sensitivity to Fuel Costs**

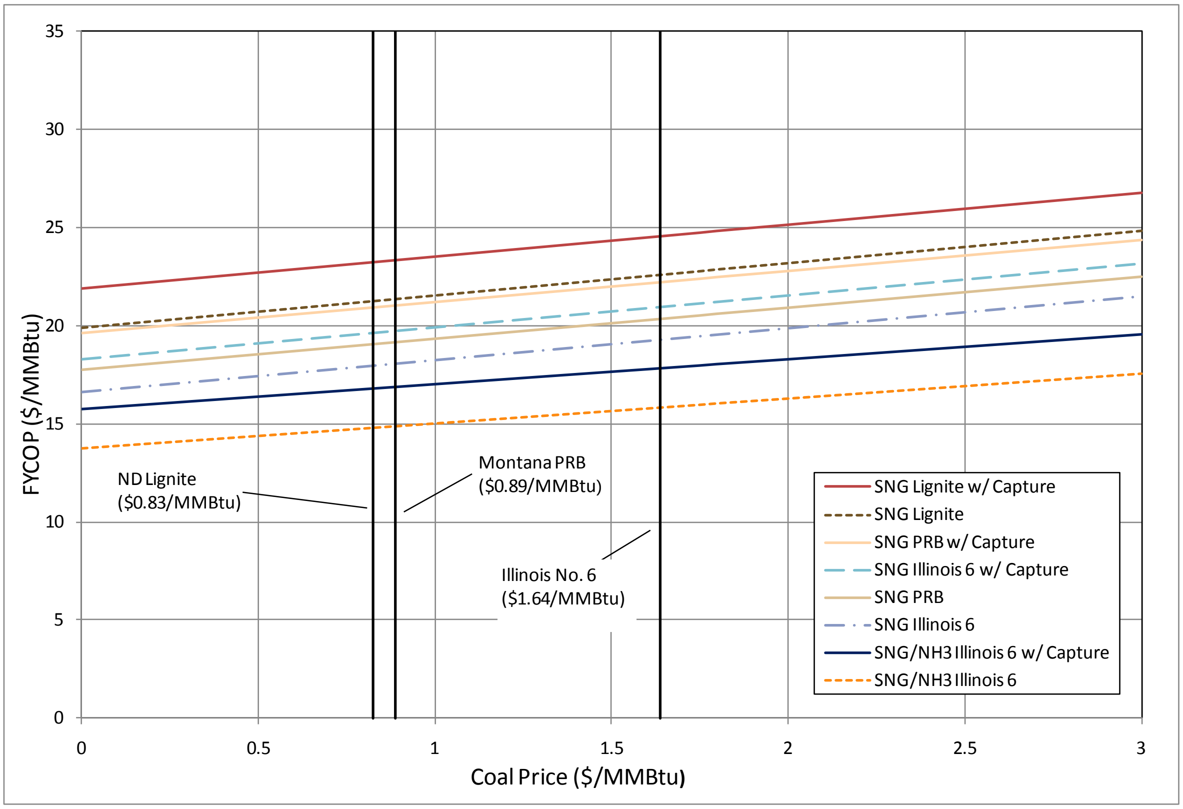
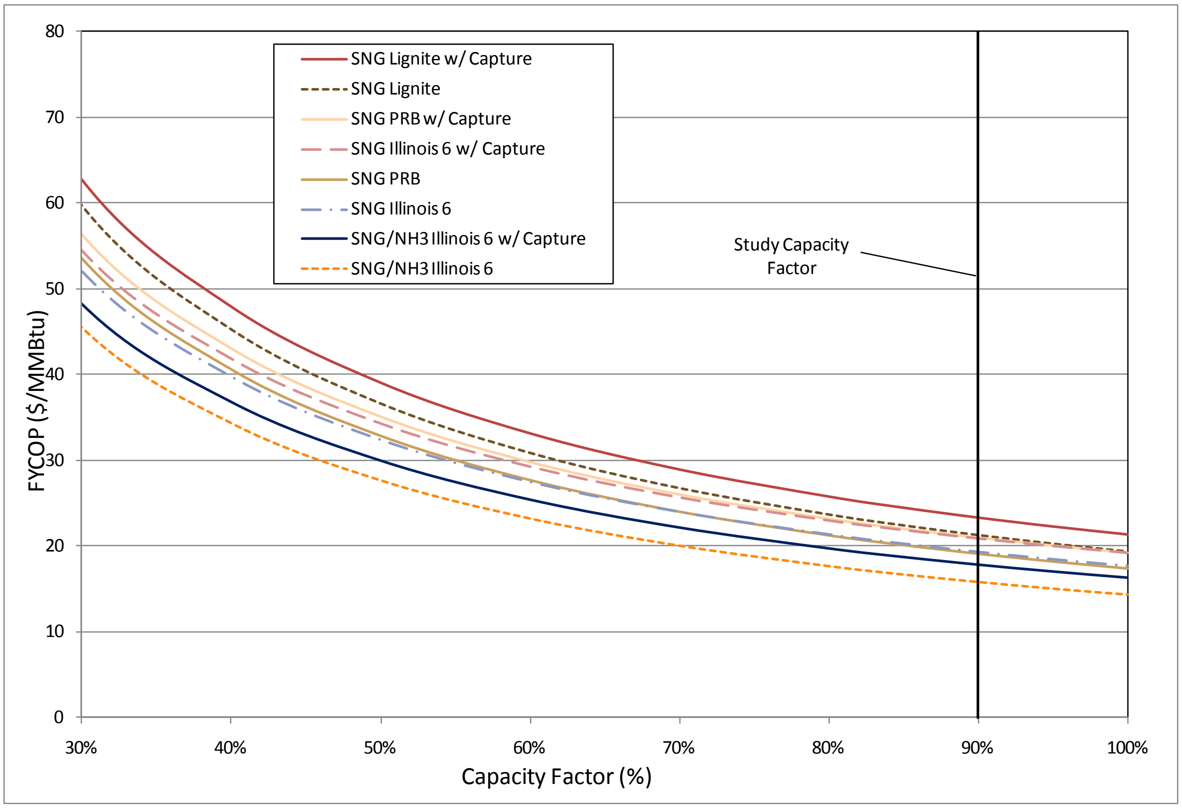



**Exhibit ES-9 FYCOP Sensitivity to Capacity Factor**



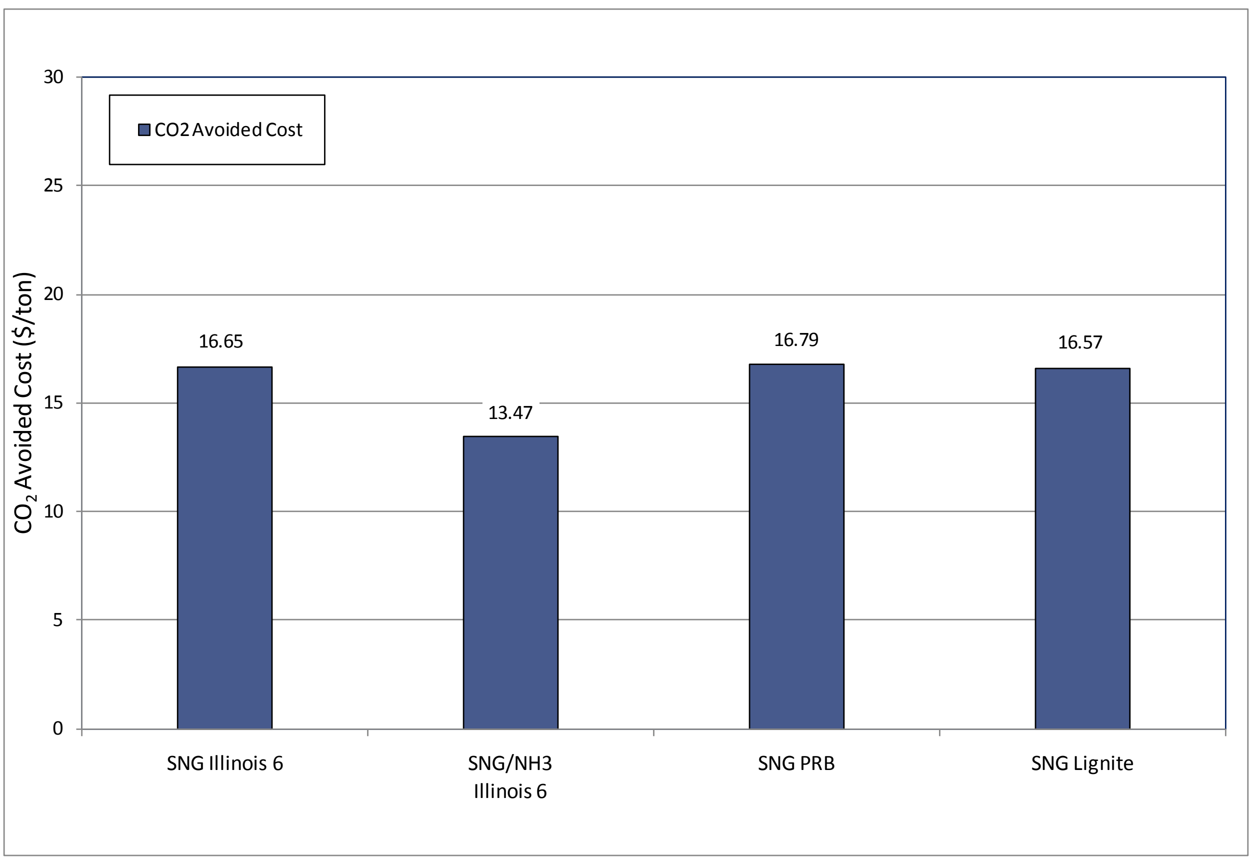
## FIRST YEAR COST OF $CO_2$ AVOIDED

In cases where the $CO_2$ was sequestered, the associated costs of $CO_2$ compression and TS&M were calculated as an avoided cost, relative to the analogous non-sequestration case, as illustrated in Equation ES-1:

$$Avoided\ Cost = \frac{\{FYCOP_{with\ sequestration} - FYCOP_{w/o\ sequestration}\}\ \$/MMBtu}{\{Emissions_{w/o\ sequestration} - Emissions_{with\ sequestration}\}\ tons/MMBtu} \qquad \text{(ES-1)}$$

The resulting first year avoided costs are shown in Exhibit ES-10 for each of the four carbon sequestration cases. The cost of $CO_2$ avoided for the SNG only cases averages $18.38/tonne ($16.67/ton) within a very narrow range. The $CO_2$ avoided costs are lower for the co-production cases because the increase in the selling price of ammonia for the carbon sequestration case absorbs some of the costs related to the addition of compression and TS&M. For integrated gasification combined cycle (IGCC) cases in Volume 1 of this series, $CO_2$ avoided costs range from $43-61/tonne ($39-56/ton) avoided. The avoided costs for the SNG plants are significantly lower than IGCC plants because the capital cost increase is lower (there is capture in all cases; the incremental cost is for compression and carbon sequestration only). The TOC increase from non-sequestration to sequestration cases ranges from 2.4 to 2.6 percent for the cases in this study. The capital cost increase for the IGCC cases in Volume 1 is from 13.5 to 21.1 percent.

**Exhibit ES-10 $CO_2$ Avoided Cost**



## ECONOMIC PATHWAYS

As stated above, the project financial structure used for the FYCOP calculations presented in this report is representative of a high-risk fuels project with no loan guarantees or other government subsidies. The parameters were chosen to reflect the next commercial offering in which the technical and economic risks have not been reduced by the wide-scale demonstration of the technology. The cost of producing SNG from coal can be substantially lower given more favorable financial and economic conditions. A more favorable structure could involve government loan guarantees, tax incentives, and shorter construction periods. Projects can also include different technologies and subsequently different capital costs, different O&M costs, production of other co-products, etc. that will impact the cost of SNG.

Exhibit ES-11 and Exhibit ES-12 show the results of adjusting selected economic assumptions for Cases 2 and 4 respectively to be representative of some of the parameters associated with a commercial coal-to-SNG project recently proposed. The modified parameters are highlighted in blue for each step. The combined modifications reduce the FYCOP by 48.6 percent for Case 2 and by 50.5 percent for Case 4. The resulting values are comparable to those quoted in the proposed project.

Loan guarantees are a near term incentive to commercialize technologies that involve high risk. Ultimately the technology must stand on its own merits. A second economic pathway was investigated to demonstrate a financial structure that would not require loan guarantees, but still achieve a similar cost of SNG. Increased commercial application of technologies reduces risks and lowers capital cost requirements. The gasification production plant capital cost in this study is based on commercial offerings; however, there have been very limited sales of these units. The costs listed reflect the “next commercial offering” level of cost rather than mature $n^{th}$-of-a-kind (NOAK) cost. As technologies mature, overall costs decrease due to multiple plant completions and lessons learned.

Exhibit ES-13 and Exhibit ES-14 show the results of adjusting the financial structure to values typically assumed for a low risk power project [1]. Reductions in TOC of 21.5 percent for Case 2 and 22.6 percent for Case 4 reduce the FYCOP values to match those of the examples shown in Exhibit ES-11 and Exhibit ES-12 respectively. While these percentages may be considered aggressive for typical first-of-a-kind (FOAK) to NOAK reductions, substantial improvements in the FYCOP are shown even when only the financial structure is modified to reflect the lower risk of mature technologies.

Additionally, after the technology matures, it may be possible for the project developer to obtain financing through the bond market and obtain longer debt payment terms than shown in Exhibit ES-13 and Exhibit ES-14 and subsequently require a lower cost of product to satisfy the equity investors. The specific financing structure options of a project developer will strongly depend on the financial wherewithal of the developer.

The FYCOP values reported in this study reflect the typical assumptions made for high risk fuels projects, which were developed for the specific purpose of comparing the relative cost of coal conversion technologies. They are not intended to represent definitive point costs; however these sensitivities to the assumed financial parameters indicate that the results are comparable to currently proposed projects.

**Exhibit ES-11 Case 2 Economic Pathway with Favorable Financial Structure**
(SNG from Illinois 6 with $CO_2$ Capture)

| Parameter | NETL Study | Step 1a | Step 1b | Step 1c | Step 2 | Step 3 | Step 4 | Step 5 |
|---|---|---|---|---|---|---|---|---|
| % Debt | 50% | 70% | 70% | 70% | 70% | 70% | 70% | 70% |
| % Equity | 50% | 30% | 30% | 30% | 30% | 30% | 30% | 30% |
| Interest Rate | 9.5% | 4.5% | 4.5% | 4.5% | 4.5% | 4.5% | 4.5% | 4.5% |
| Required IRROE | 20.0% | 20.0% | 20.0% | 20.0% | 20.0% | 20.0% | 20.0% | 20.0% |
| Repayment Term (years) | 15 | 15 | 30 | 30 | 30 | 30 | 30 | 30 |
| Capital Expenditure Period (Months)* | 60 | 60 | 60 | 48 | 48 | 48 | 48 | 48 |
| Capital Cost Escalation (%/yr) | 3.6% | 3.6% | 3.6% | 3.6% | 0.0% | 0.0% | 0.0% | 0.0% |
| Owner's Cost (%TPC) | 23% | 23% | 23% | 23% | 23% | 18% | 18% | 18% |
| Property Taxes and Insurance (%TPC) | 2.00% | 2.00% | 2.00% | 2.00% | 2.00% | 2.00% | 0.42% | 0.42% |
| $CO_2$ Credit ($/Tonne) | $0.00 | $0.00 | $0.00 | $0.00 | $0.00 | $0.00 | $0.00 | $10.00 |
| $CO_2$ Captured (Tonne/Day) | 15,134 | 15,134 | 15,134 | 15,134 | 15,134 | 15,134 | 15,134 | 15,134 |
| TOC (1000$) | 3,312,740 | 3,312,740 | 3,312,740 | 3,312,740 | 3,312,740 | 3,180,577 | 3,180,577 | 3,180,577 |
| **SNG COP $/MMBtu (excluding TS&M costs)** | **20.04** | **15.17** | **13.47** | **12.68** | **12.29** | **12.01** | **11.23** | **10.31** |
| Incremental Step Reduction | | 24.3% | 11.2% | 5.9% | 3.0% | 2.3% | 6.5% | 8.2% |
| **Cumulative Reduction** | | **24.3%** | **32.8%** | **36.7%** | **38.7%** | **40.1%** | **44.0%** | **48.6%** |

* Capital Expenditure Period (Year-by-Year Distribution)
4 years (48 months) 10 % - 40 % - 30 % - 20 %
5 years (60 months) 10 % - 30 % - 25 % - 20 % - 15 %

Sensitivity Analysis Steps
1. Modify financing structure (debt/equity & interest rate)
   1a Increase percentage of debt and decrease interest on debt
   1b Increase loan repayment term
   1c Decrease capital expenditure period
2. Reduce capital cost escalation during the capital expenditure period
3. Reduce owner's cost
4. Reduce taxes and insurance in fixed O&M costs
5. Assume $CO_2$ revenue value for enhanced oil recovery

**Exhibit ES-12 Case 4 Economic Pathway with Favorable Financial Structure**
(SNG and Ammonia from Illinois 6 with $CO_2$ Capture)

| Parameter | NETL Study | Step 1a | Step 1b | Step 1c | Step 2 | Step 3 | Step 4 | Step 5 |
|---|---|---|---|---|---|---|---|---|
| % Debt | 50% | 70% | 70% | 70% | 70% | 70% | 70% | 70% |
| % Equity | 50% | 30% | 30% | 30% | 30% | 30% | 30% | 30% |
| Interest Rate | 9.5% | 4.5% | 4.5% | 4.5% | 4.5% | 4.5% | 4.5% | 4.5% |
| Required IRROE | 20.0% | 20.0% | 20.0% | 20.0% | 20.0% | 20.0% | 20.0% | 20.0% |
| Repayment Term (years) | 15 | 15 | 30 | 30 | 30 | 30 | 30 | 30 |
| Capital Expenditure Period (Months)* | 60 | 60 | 60 | 48 | 48 | 48 | 48 | 48 |
| Capital Cost Escalation (%/yr) | 3.6% | 3.6% | 3.6% | 3.6% | 0.0% | 0.0% | 0.0% | 0.0% |
| Owner's Cost (%TPC) | 23% | 23% | 23% | 23% | 23% | 18% | 18% | 18% |
| Property Taxes and Insurance (%TPC) | 2.00% | 2.00% | 2.00% | 2.00% | 2.00% | 2.00% | 0.42% | 0.42% |
| $CO_2$ Credit ($/Tonne) | $0.00 | $0.00 | $0.00 | $0.00 | $0.00 | $0.00 | $0.00 | $10.00 |
| $CO_2$ Captured (Tonne/Day) | 16,763 | 16,763 | 16,763 | 16,763 | 16,763 | 16,763 | 16,763 | 16,763 |
| TOC (1000$) | 3,829,817 | 3,829,817 | 3,829,817 | 3,829,817 | 3,829,817 | 3,684,431 | 3,684,431 | 3,684,431 |
| **SNG COP $/MMBtu (excluding TS&M costs)** | **16.56** | **12.60** | **11.06** | **10.30** | **9.96** | **9.71** | **9.00** | **8.20** |
| Incremental Step Reduction | | 23.9% | 12.3% | 6.8% | 3.4% | 2.5% | 7.3% | 8.8% |
| **Cumulative Reduction** | | **23.9%** | **33.2%** | **37.8%** | **39.9%** | **41.4%** | **45.7%** | **50.5%** |

* Capital Expenditure Period (Year-by-Year Distribution)
4 years (48 months) 10 % - 40 % - 30 % - 20 %
5 years (60 months) 10 % - 30 % - 25 % - 20 % - 15 %

Sensitivity Analysis Steps
1. Modify financing structure (debt/equity & interest rate)
   1a Increase percentage of debt and decrease interest on debt
   1b Increase loan repayment term
   1c Decrease capital expenditure period
2. Reduce capital cost escalation during the capital expenditure period
3. Reduce owner's cost
4. Reduce taxes and insurance in fixed O&M costs
5. Assume $CO_2$ revenue value for enhanced oil recovery

**Exhibit ES-13 Case 2 Economic Pathway with Mature Technology**
(SNG from Illinois 6 with $CO_2$ Capture)

| Parameter | NETL Study | Step 1a | Step 1b | Step 2 |
|---|---|---|---|---|
| % Debt | 50% | 50% | 50% | 50% |
| % Equity | 50% | 50% | 50% | 50% |
| Interest Rate | 9.5% | 4.5% | 4.5% | 4.5% |
| Required IRROE | 20.0% | 20.0% | 12.0% | 12.0% |
| Repayment Term (years) | 15 | 15 | 15 | 15 |
| Capital Expenditure Period (Months)* | 60 | 60 | 60 | 60 |
| Capital Cost %Escalation/yr | 3.6% | 3.6% | 3.6% | 3.6% |
| Owner's Cost %TPC | 23% | 23% | 23% | 23% |
| Property Taxes and Insurance %TPC | 2.00% | 2.00% | 2.00% | 2.00% |
| $CO_2$ Credit ($/Tonne) | $0.00 | $0.00 | $0.00 | $0.00 |
| $CO_2$ Tonne/Day Captured | 15,134 | 15,134 | 15,134 | 15,134 |
| TOC (1000$) | 3,312,740 | 3,312,740 | 3,312,740 | 2,602,157 |
| Percent capital (TOC) reduction | 0.0% | 0.0% | 0.0% | 21.5% |
| **SNG COP $/MMBtu (excluding TS&M costs)** | **20.04** | **18.54** | **12.21** | **10.31** |
| Incremental Step Reduction | | 7.5% | 34.1% | 15.6% |
| **Cumulative Reduction** | | **7.5%** | **39.1%** | **48.6%** |

* Capital Expenditure Period (Year-by-Year Distribution)
5 years (60 months) 10 % - 30 % - 25 % - 20 % - 15 %

Sensitivity Analysis Steps
1. Modify financing structure (debt/equity & interest rate)
   1a Decrease interest on debt
   1b Decrease required internal rate of return on equity (IRROE)
2. Reduce capital cost to reflect FOAK to NOAK improvements

**Exhibit ES-14 Case 4 Economic Pathway with Mature Technology**
(SNG and Ammonia from Illinois 6 with $CO_2$ Capture)

| Parameter | NETL Study | Step 1a | Step 1b | Step 2 |
|---|---|---|---|---|
| % Debt | 50% | 50% | 50% | 50% |
| % Equity | 50% | 50% | 50% | 50% |
| Interest Rate | 9.5% | 4.5% | 4.5% | 4.5% |
| Required IRROE | 20.0% | 20.0% | 12.0% | 12.0% |
| Repayment Term (years) | 15 | 15 | 15 | 15 |
| Capital Expenditure Period (Months)* | 60 | 60 | 60 | 60 |
| Capital Cost %Escalation/yr | 3.6% | 3.6% | 3.6% | 3.6% |
| Owner's Cost %TPC | 23% | 23% | 23% | 23% |
| Property Taxes and Insurance %TPC | 2.00% | 2.00% | 2.00% | 2.00% |
| $CO_2$ Credit ($/Tonne) | $0.00 | $0.00 | $0.00 | $0.00 |
| $CO_2$ Tonne/Day Captured | 16,763 | 16,763 | 16,763 | 16,763 |
| TOC (1000$) | 3,829,817 | 3,829,817 | 3,829,817 | 2,962,746 |
| Percent capital (TOC) reduction | 0.0% | 0.0% | 0.0% | 22.6% |
| **SNG COP $/MMBtu (excluding TS&M costs)** | **16.56** | **15.67** | **10.02** | **8.20** |
| Incremental Step Reduction | | 5.4% | 36.0% | 18.2% |
| **Cumulative Reduction** | | **5.4%** | **39.5%** | **50.5%** |

* Capital Expenditure Period (Year-by-Year Distribution)
5 years (60 months) 10 % - 30 % - 25 % - 20 % - 15 %

Sensitivity Analysis Steps
1. Modify financing structure (debt/equity & interest rate)
   1a Decrease interest on debt
   1b Decrease required internal rate of return on equity (IRROE)
2. Reduce capital cost to reflect FOAK to NOAK improvements

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# 1. INTRODUCTION

The objective of this report is to present an accurate, independent assessment of the cost and performance of fossil energy fuel systems, specifically the Siemens gasifier using bituminous and low rank coals to produce SNG and ammonia as a co-product, in a consistent technical and economic manner that accurately reflects current market conditions for plants starting operation in the near term. This report is Volume 2 of a four volume series. The four volumes are defined below with Volume 1 having been originally published in May 2007 and the most recent revision issued in November 2010. Volume 3 was released in March and May, 2011 and Volume 4 is expected to be released later in 2011.

- Volume 1: Bituminous Coal and Natural Gas to Electricity
- Volume 2: Coal to Synthetic Natural Gas and Ammonia (Various Coal Ranks)
- Volume 3: Low Rank Coal and Natural Gas to Electricity
- Volume 4: Bituminous Coal to Liquid Fuels with Carbon Capture

Eight cases are modeled and analyzed. Since all cases include $CO_2$ capture the normal designation of capture and non-capture is not used in this study. Rather the cases are differentiated by the designation carbon sequestration or non-sequestration. The carbon sequestration cases assume that compression and drying are the only steps required to produce $CO_2$ suitable for transportation and sequestration. Two cases use Illinois No. 6 bituminous coal for the production of SNG only, with and without carbon sequestration; two cases use Illinois No. 6 coal for the co-production of SNG and ammonia, with and without carbon sequestration; two cases use PRB subbituminous coal for the production of SNG only, with and without carbon sequestration; and two cases use NDL coal for the production of SNG only, also with and without carbon sequestration.

The gasification scheme used in this study includes a partial quench followed by a syngas cooler. Heat recovery was necessary to produce enough power to satisfy internal auxiliary loads with some excess power for export in most cases. This configuration is not currently a commercial offering by Siemens, but based on personal communications, it is a configuration they are considering for coal-to-SNG projects.

This Final Report covers the following eight cases:

**Case 1 –** This case consists of a coal to SNG plant without $CO_2$ compression and sequestration based on a Siemens gasifier. Illinois No. 6 coal is the fuel and the plant is located at a generic site in Midwestern, U.S.

**Case 2 –** This case is the same as Case 1 except it includes $CO_2$ compression and sequestration.

**Case 3 –** This case consists of a coal to SNG and ammonia co-production plant without $CO_2$ compression and sequestration based on a Siemens gasifier. Illinois No. 6 coal is the fuel and the plant is located at a generic site in Midwestern, U.S.

**Case 4 –** This case is the same as Case 3 except it includes $CO_2$ compression and sequestration.

**Case 5 –** This case includes a coal to SNG plant without $CO_2$ compression and sequestration based on a Siemens gasifier. Rosebud PRB coal is the fuel and the plant is located at a generic site in Western, U.S.

**Case 6 –** This case is the same as Case 5 except it includes $CO_2$ compression and sequestration.

**Case 7 –** This case consists of a coal to SNG plant without $CO_2$ compression and sequestration based on a Siemens gasifier. NDL coal is the fuel and the plant is located at a minemouth site in North Dakota, U.S.

**Case 8 –** This case is the same as Case 7 except it includes $CO_2$ compression and sequestration.

While input was sought from various technology vendors, the final assessment of performance and cost was determined independently, and may not represent the views of the technology vendors.

**Generating Unit Configurations**

The six SNG only cases have a nominal gasifier thermal input of 500 MW. The actual capacity varies from 506 to 549 MWth depending on coal type. This translates into a production of 52 to 57 billion standard cubic feet (Bscf)/year of SNG at 90 percent CF. This translates into a daily capacity of 158-172 million standard cubic feet (MMScf)/day of SNG. The two SNG and ammonia co-production cases are based on a standard-size 2,000 tonnes/day (2,204 tons/day) ammonia facility with the remainder of the syngas used for SNG production.

The balance of this report is organized as follows:

- Chapter 2 provides the basis for technical, environmental, and cost evaluations.
- Chapter 3 describes the major process systems for all eight cases
- Chapter 4 provides the detailed results of the four bituminous coal cases.
- Chapter 5 provides the detailed results of the four low rank coal cases.
- Chapter 0 summarizes the results of the eight cases.
- Chapter 7 contains the reference list.

# 2. GENERAL EVALUATION BASIS

For each of the plant configurations in this study an Aspen model was developed and used to generate material and energy balances, which in turn were used to provide a design basis for items in the major equipment list. The equipment list and material balances were used as the basis for generating the capital and operating cost estimates. Performance and process limits were based upon published reports, information obtained from vendors of the technology, performance data from design/build utility projects, and/or best engineering judgment. Capital and operating costs were estimated by WorleyParsons Group, Inc. (WorleyParsons) based on simulation results and through a combination of vendor quotes and scaled estimates from previous design/build projects. Ultimately a first year cost of production (FYCOP) of SNG was calculated for each of the cases and is reported as the economic figure-of-merit. Due to the process requirements for both SNG and ammonia production, all cases utilize an AGR system that removes or captures the $CO_2$ from the syngas. Some cases simply release this removed $CO_2$ to the atmosphere and are labeled “non-sequestration.” The cases that compress the removed $CO_2$ and include TS&M are labeled “sequestration.”

The balance of this chapter documents the design basis, environmental targets, and cost assumptions used in the study.

## 2.1 SITE CHARACTERISTICS

The plants are located at three different generic plant sites. Plants using Illinois No. 6 coal are assumed to be located at a generic site in Midwestern, U.S. Plants using PRB coal are assumed to be located at a generic site in Montana. Plants using lignite coal are assumed to be located at a minemouth site in North Dakota. The ambient conditions for the three sites are shown in Exhibit 2-1.

**Exhibit 2-1 Ambient Conditions**

| | Location | | |
|---|---|---|---|
| | Midwest | Montana | North Dakota |
| Average Elevation, ft | 0 | 3,400 | 1,900 |
| Barometric Pressure, psia | 14.7 | 13.0 | 13.8 |
| Design Ambient Temperature, Dry Bulb, °F | 59 | 42 | 40 |
| Design Ambient Temperature, Wet Bulb, °F | 51.5 | 37 | 36 |
| Design Ambient Relative Humidity, % | 60 | 62 | 68 |

The site characteristics are assumed to be the same for all plant locations, except where noted as shown in Exhibit 2-2.

**Exhibit 2-2 Site Characteristics**

| | |
|---|---|
| Location | Greenfield, Midwest (Illinois No. 6 coal), Montana (PRB coal) or North Dakota (lignite coal) |
| Topography | Level |
| Size, acres | 300 |
| Transportation | Rail |
| Ash/Slag Disposal | Off Site |
| Water | Municipal (50%) / Groundwater (50%) |
| Access | Land locked, having access by railway and highway |
| $CO_2$ Storage | Compressed to 15.3 MPa (2,215 psia), transported 80 kilometers (50 miles) and sequestered in a saline formation at a depth of 1,239 meters (4,055 feet) |

The land area for all the production plant cases assumes 30 acres are required for the plant proper and the balance provides a buffer of approximately 0.25 miles to the fence line. The extra land could also provide for a rail loop if required.

In all cases it was assumed that the steam turbine (ST) is enclosed in a turbine building, but the gasifier is not enclosed. The following design parameters are considered site-specific, and are not quantified for this study. Allowances for normal conditions and construction are included in the cost estimates.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Fire protection
- Local code height requirements
- Noise regulations – Impact on site and surrounding area

## 2.2 PLANT PRODUCTION CAPACITY

SNG Only Plant

Currently, the only operating SNG plant (and other co-products) within the U.S. is the Dakota Gasification Company's Great Plains Synfuels Plant. This plant has an annual production of approximately 54 Bscf [2]. In addition, there are several proposed SNG plants in the United States as listed in Exhibit 2-3 [3]. The gasifier capacity is coal-type dependent and the resulting SNG capacity varies accordingly. The six gasifier systems in this study result in average SNG outputs of 56 Bscf/yr for bituminous coal, 55 Bscf/yr for PRB coal and 52 Bscf/yr for lignite coal at 90 percent CF. These capacities are consistent with the existing plant and announced projects shown in Exhibit 2-3.

**Exhibit 2-3 Proposed Capacity for SNG Plants in the United States**

| Project | Location | Feedstock | Gasifier | Capacity | Units |
|---|---|---|---|---|---|
| Kentucky NewGas | Kentucky | Kentucky Bituminous | CoP E-Gas | 1,700 – 1,980 (60-70) | MNm$^3$/year (Bscf/year) |
| Hunton Energy Freeport Plant (Lockwood) | Texas | petcoke | GEE | 1,870 (66) | MNm$^3$/year (Bscf/year) |
| South Heart | North Dakota | ND Lignite | Lurgi | 1,050 (37) | MNm$^3$/year (Bscf/year) |
| Taylorville Energy Center | Illinois | Illinois Bituminous | GEE | 74 - 450 (2.6-16) | MNm$^3$/year (Bscf/year) |
| Southern Illinois Coal-to-SNG facility | Illinois | Illinois Bituminous | GEE | 1,420 (50) | MNm$^3$/year (Bscf/year) |
| Secure Energy Decatur Gasification | Illinois | Illinois Bituminous | Siemens | 570 (20) | MNm$^3$/year (Bscf/year) |
| Indiana SNG Project | Indiana | Illinois Bituminous | GEE | 1,130 (40) | MNm$^3$/year (Bscf/year) |
| Lake Charles Cogeneration LLC | Louisiana | petcoke | GEE | unknown | |
| Scriba Coal Gasification Plant | New York | coal | unknown | 2,830 (100) | MNm$^3$/year (Bscf/year) |
| Mississippi SNG Project | Mississippi | petcoke | CoP E-Gas | 990 (35) | MNm$^3$/year (Bscf/year) |
| Cash Creek SNG Project | Kentucky | Kentucky Bituminous | GEE | unknown | |

<u>SNG and Ammonia Co-Production Plant</u>

The co-production facility cases use the same coal type and coal feed rate as the Illinois No. 6 bituminous coal non-sequestration and sequestration SNG only cases (Cases 1 and 2) with a portion of the syngas utilized to feed a large commercial size ammonia plant and the remaining syngas utilized for SNG production. Some currently operating ammonia plants in the United States and their associated feedstock and capacity are shown in Exhibit 2-4.

**Exhibit 2-4 U.S. Ammonia Plants**

| Company | Location | Feedstock | Capacity | Units |
|---|---|---|---|---|
| Agrium Inc. | Texas | natural gas | 1,342 (1,479) | tonnes/day (tons/day) |
| CF Industries, Inc. | Louisiana | natural gas | 5,589 (6,156) | tonnes/day (tons/day) |
| Coffeyville Resources, Nitrogen Fertilizers, LLC | Louisiana | coal | 1,027 (1,132) | tonnes/day (tons/day) |
| Dakota Gasification Co. | North Dakota | coal | 995 (1,095) | tonnes/day (tons/day) |

| Company | Location | Feedstock | Capacity | Units |
|---|---|---|---|---|
| Honeywell International Inc. | Virginia | natural gas | 1,452 (1,599) | tonnes/day (tons/day) |
| Koch Nitrogen Co. | Nebraska | natural gas | 726 (800) | tonnes/day (tons/day) |
| Mosaic Co. | Louisiana | coal/petcoke | 1,392 (1,533) | tonnes/day (tons/day) |
| PCS Nitrogen, Inc., LP | Georgia | natural gas | 1,885 (2,076) | tonnes/day (tons/day) |
| Rentech Energy Midwest Corp. | Illinois | coal | 762 (839) | tonnes/day (tons/day) |

The standard capacity for new ammonia plants being built today is approximately 2,000 tonnes/day (2,204 tons/day) and is used as the basis for the ammonia portion of the co-production facilities [4]. Advances in ammonia production technology have enabled even higher production rates. At higher operating pressures, higher ammonia conversion per pass is achieved. This allows ammonia plant production capacity to increase while maintaining reasonable reactor and equipment sizes [5]. The remaining portion of the syngas is used to produce approximately 42 Bscf/year of SNG at 90 percent CF.

## 2.3 COAL CHARACTERISTICS

There are three design coals, bituminous Illinois No. 6 coal, subbituminous PRB coal from Montana, and lignite coal from North Dakota. The coal properties are from NETL's Coal Quality Guidelines and are shown in Exhibit 2-5, Exhibit 2-6, and Exhibit 2-7 [6].

The coal mercury (Hg) content was not available from reference 6 and was obtained through alternate means. Thirty-four samples of Illinois No. 6 coal have an arithmetic mean value of 0.09 parts per million (ppm) (dry basis) with standard deviation of 0.06 based on coal samples shipped by Illinois mines [7]. The coal mercury content for Illinois No. 6 coal shown in Exhibit 2-5 was assumed to be 0.15 ppm (dry) for all Illinois No. 6 cases, which corresponds to the mean plus one standard deviation and encompasses about 84 percent of the samples.

The mercury concentration for the low rank coals used in this study was determined from the Environmental Protection Agency's (EPA) Information Collection Request (ICR) database. The ICR database has 137 records of Montana Rosebud subbituminous coal with an average Hg concentration of 0.056 ppm (dry) and a standard deviation of 0.025 ppm. There are 266 records for NDL from the Beulah seam with an average Hg concentration of 0.081 ppm (dry) and a standard deviation of 0.035 ppm. The mercury values in Exhibit 2-6 and Exhibit 2-7 are the mean plus one standard deviation, or 0.081 ppm (dry) for PRB coal and 0.116 ppm (dry) for NDL [8]. It was further assumed that all of the coal Hg enters the gas phase and none leaves with the bottom ash or slag.

The Power Systems Financial Model (PSFM) was used to derive the capital charge factor (CCF) for this study [9]. The PSFM requires that all cost inputs have a consistent cost year basis. Because the capital and operating cost estimates are in June 2007 dollars, the fuel costs must also be in June 2007 dollars

The coal cost used in this study is $1.55/GJ ($1.64/MMBtu) for Illinois No. 6 coal, $0.84/GJ ($0.89/MMBtu) for PRB coal and $0.78/GJ ($0.83/MMBtu) for lignite coal (2007 cost of coal in

June 2007 dollars). These costs were determined using the following information from the EIA's 2008 AEO:

- The 2007 minemouth cost of Illinois No. 6 in 2006 dollars, $32.66/tonne ($29.63/ton), was obtained from Supplemental Table 112 of the EIA's 2008 AEO for eastern interior high-sulfur bituminous coal. From the same table the 2007 minemouth cost of PRB coal in 2006 dollars is $13.02/tonne ($11.81/ton) and of NDL coal in 2006 dollars is $11.68/tonne ($10.59/ton).
- The cost of Illinois No. 6 coal was escalated to 2007 dollars using the gross domestic product (GDP) chain-type price index from AEO 2008, resulting in a price of $33.67/tonne ($30.55/ton) [10]. Similarly, the cost of PRB coal in 2007 dollars is $13.42/tonne ($12.17/ton) and the cost of NDL in 2007 dollars is $12.04/tonne ($10.92/ton).
- Transportation costs for Illinois No. 6 and PRB coal were estimated to be 25 percent of the minemouth cost based on the average transportation rate of the respective coals to the surrounding regions [11]. The final delivered costs for Illinois No. 6 coal shown in the report is $42.09/tonne ($38.18/ton) or $1.55/GJ ($1.64/MMBtu). PRB coal has a final delivered cost of $16.77/tonne ($15.22/ton) or $0.84/GJ ($0.89/MMBtu). The NDL plant is assumed to be located at the minemouth, so transportation costs are zero. (Note: The Illinois No. 6 coal cost of $1.6366/MMBtu was used in calculations, but only two decimal places are shown in the report. Similarly, the PRB and lignite fuel costs convert to $0.8884/MMBtu and $0.8251/MMBtu, respectively.)

**Exhibit 2-5 Illinois No. 6 Bituminous, Old Ben Mine Bituminous Design Coal Analysis**

| **Proximate Analysis** | **Dry Basis, %** | **As Received, %** |
|---|---|---|
| Moisture | 0.0 | 11.12 |
| Ash | 10.91 | 9.70 |
| Volatile Matter | 39.37 | 34.99 |
| Fixed Carbon | 49.72 | 44.19 |
| Total | 100.0 | 100.0 |
| **Ultimate Analysis** | **Dry Basis, %** | **As Received, %** |
| Carbon | 71.72 | 63.75 |
| Hydrogen | 5.06 | 4.50 |
| Nitrogen | 1.41 | 1.25 |
| Sulfur | 2.82 | 2.51 |
| Chlorine | 0.33 | 0.29 |
| Ash | 10.91 | 9.70 |
| Moisture | 0.00 | 11.12 |
| Oxygen (Note A) | 7.75 | 6.88 |
| Total | 100.0 | 100.0 |
| **Heating Value** | **Dry Basis** | **As Received** |
| HHV, kJ/kg | 30,506 | 27,113 |
| HHV, Btu/lb | 13,126 | 11,666 |
| LHV, kJ/kg | 29,544 | 26,151 |
| LHV, Btu/lb | 12,712 | 11,252 |
| **Hardgrove Grindability Index** | 60 | |
| **Ash Mineral Analysis** | | **%** |
| Silica | $SiO_2$ | 45.0 |
| Aluminum Oxide | $Al_2O_3$ | 18.0 |
| Iron Oxide | $Fe_2O_3$ | 20.0 |
| Titanium Dioxide | $TiO_2$ | 1.0 |
| Calcium Oxide | CaO | 7.0 |
| Magnesium Oxide | MgO | 1.0 |
| Sodium Oxide | $Na_2O$ | 0.6 |
| Potassium Oxide | $K_2O$ | 1.9 |
| Sulfur Trioxide | $SO_3$ | 3.5 |
| Phosphorous Pentoxide | $P_2O_5$ | 0.2 |
| Barium Oxide | $Ba_2O$ | 0.00 |
| Strontium Oxide | SrO | 0.00 |
| Unknown | --- | 1.8 |
| | Total | 100.0 |
| **Trace Components** | | **ppmd** |
| Mercury (Note B) | Hg | 0.15 |

Notes: A. By Difference

B. Mercury value is the mean plus one standard deviation using 2004 Keystone Coal Industry Manual

**Exhibit 2-6 Montana Rosebud PRB, Area D, Western Energy Co. Mine Subbituminous Design Coal Analysis**

| **Proximate Analysis** | **Dry Basis, %** | **As Received, %** |
|---|---|---|
| Moisture | 0.0 | 25.77 |
| Ash | 11.04 | 8.19 |
| Volatile Matter | 40.87 | 30.34 |
| Fixed Carbon | 48.09 | 35.70 |
| Total | 100.0 | 100.0 |
| **Ultimate Analysis** | **Dry Basis, %** | **As Received, %** |
| Carbon | 67.45 | 50.07 |
| Hydrogen | 4.56 | 3.38 |
| Nitrogen | 0.96 | 0.71 |
| Sulfur | 0.98 | 0.73 |
| Chlorine | 0.01 | 0.01 |
| Ash | 11.03 | 8.19 |
| Moisture | 0.00 | 25.77 |
| Oxygen (Note A) | 15.01 | 11.14 |
| Total | 100.0 | 100.0 |
| **Heating Value** | **Dry Basis** | **As Received** |
| HHV, kJ/kg | 26,787 | 19,920 |
| HHV, Btu/lb | 11,516 | 8,564 |
| LHV, kJ/kg | 25,810 | 19,195 |
| LHV, Btu/lb | 11,096 | 8,252 |
| **Hardgrove Grindability Index** | 57 | |
| **Ash Mineral Analysis** | | **%** |
| Silica | $SiO_2$ | 38.09 |
| Aluminum Oxide | $Al_2O_3$ | 16.73 |
| Iron Oxide | $Fe_2O_3$ | 6.46 |
| Titanium Dioxide | $TiO_2$ | 0.72 |
| Calcium Oxide | CaO | 16.56 |
| Magnesium Oxide | MgO | 4.25 |
| Sodium Oxide | $Na_2O$ | 0.54 |
| Potassium Oxide | $K_2O$ | 0.38 |
| Sulfur Trioxide | $SO_3$ | 15.08 |
| Phosphorous Pentoxide | $P_2O_5$ | 0.35 |
| Barium Oxide | $Ba_2O$ | 0.00 |
| Strontium Oxide | SrO | 0.00 |
| Unknown | --- | 0.84 |
| | Total | 100.0 |
| **Trace Components** | | **ppmd** |
| Mercury (Note B) | Hg | 0.081 |

Notes: A. By Difference
B. Mercury value is the mean plus one standard deviation using EPA's ICR data

**Exhibit 2-7 North Dakota Beulah-Zap Lignite, Freedom, ND Mine Lignite Design Coal Analysis**

| **Proximate Analysis** | **Dry Basis, %** | **As Received, %** |
|---|---|---|
| Moisture | 0.0 | 36.08 |
| Ash | 15.43 | 9.86 |
| Volatile Matter | 41.49 | 26.52 |
| Fixed Carbon | 43.09 | 27.54 |
| Total | 100.0 | 100.0 |
| **Ultimate Analysis** | **Dry Basis, %** | **As Received, %** |
| Carbon | 61.88 | 39.55 |
| Hydrogen | 4.29 | 2.74 |
| Nitrogen | 0.98 | 0.63 |
| Sulfur | 0.98 | 0.63 |
| Chlorine | 0.00 | 0.00 |
| Ash | 15.43 | 9.86 |
| Moisture | 0.00 | 36.08 |
| Oxygen (Note A) | 16.44 | 10.51 |
| Total | 100.0 | 100.0 |
| **Heating Value** | **Dry Basis** | **As Received** |
| HHV, kJ/kg | 24,254 | 15,391 |
| HHV, Btu/lb | 10,427 | 6,617 |
| LHV, kJ/kg | 23,335 | 14,804 |
| LHV, Btu/lb | 10,032 | 6,364 |
| **Hardgrove Grindability Index** | Not applicable | |
| **Ash Mineral Analysis** | | **%** |
| Silica | $SiO_2$ | 35.06 |
| Aluminum Oxide | $Al_2O_3$ | 12.29 |
| Iron Oxide | $Fe_2O_3$ | 5.12 |
| Titanium Dioxide | $TiO_2$ | 0.58 |
| Calcium Oxide | CaO | 14.39 |
| Magnesium Oxide | MgO | 6.61 |
| Sodium Oxide | $Na_2O$ | 5.18 |
| Potassium Oxide | $K_2O$ | 0.64 |
| Sulfur Trioxide | $SO_3$ | 16.27 |
| Barium Oxide | $Ba_2O$ | 0.56 |
| Strontium Oxide | SrO | 0.27 |
| Manganese Dioxide | $MnO_2$ | 0.02 |
| Unknown | --- | 3.00 |
| | Total | 100.0 |
| **Trace Components** | | **ppmd** |
| Mercury (Note B) | Hg | 0.116 |

Notes: A. By Difference
B. Mercury value is the mean plus one standard deviation using EPA's ICR data

## 2.4 PRODUCT SPECIFICATIONS

Synthetic Natural Gas

To be a viable pipeline replacement fuel for natural gas, the SNG must meet certain specifications. Exhibit 2-8 lists generic specifications for domestic and commercial natural gas from a pipeline handbook [12].

**Exhibit 2-8 SNG Pipeline Requirements**

| Parameter | Value |
|---|---|
| Heating Value, kJ/Nm$^3$ (Btu/scf) | > 35,367 (950) |
| Sulfur Content, kg/100 Nm$^3$ (lb/100 scf) | < 0.0023 (0.00014) |
| $CO_2$ Content, vol % | < 2 |
| Moisture Content, kg/1000 Nm$^3$ (lb/1000 scf) | < 62.5 (3.9) |
| Temperature, °C (°F) | < 49 (120) |

As liquefied natural gas (LNG) and other non-conventional gas sources become increasingly prevalent, more attention is being placed on natural gas specifications and its interchangeability. An administrative law judge at the Federal Energy Regulatory Commission (FERC) recently ruled in favor of Florida Gas Transmission (FGT) setting stringent specifications on LNG imported through the gulf coast [13]. The ruling set an acceptable Wobbe Index range of 1,340-1,396 and gross heating value range of 1,025-1,110 British thermal unit per standard cubic foot (Btu/scf). The Wobbe Index is a measure of the energy carrying capacity of a given fuel and is found by dividing the fuel heating value by the square root of the specific gravity (SG). Notably the ruling was not applied to domestic gas received into FGT's system.

There is a wide range of fuel quality that is acceptable to most combustion equipment, but the same equipment often will not handle high variability in the fuel quality. The SNG produced in this study falls at the low end of the acceptable quality range with HHVs ranging from 965-975 Btu/scf, Wobbe Indices in the range of 1,265-1,275, and inert concentrations ranging from 3-4.5 percent. The primary reason for the lower quality product is the absence of higher hydrocarbons. The methanation process produces only methane, and the balance of the SNG is necessarily inerts. Pure methane has a HHV of 1,012 Btu/scf and a Wobbe Index of 1,357, which represent the upper limits of the methanation process.

The quality of the SNG produced in this study could be enhanced in a number of ways:

- Increase the oxygen purity from 99 percent to 99.5 percent (or greater) with a corresponding increase in air separation unit (ASU) cost.
- Upgrade the SNG purification system (pressure swing adsorption [PSA]) by increasing the bed depth or adding beds in series to remove additional $CO_2$ (again at increased cost).

Ultimately the SNG purity requirements will be dictated by project location and fuel end use. The quality of the product is consistent across the cases modeled in this study.

Ammonia

The two general quality specifications for commercial grade ammonia are shown in Exhibit 2-9 [14]. For the co-production cases, the commercial grade specification was used as the refrigeration grade specification requires further purification that was deemed beyond the scope of these baseline cases.

**Exhibit 2-9 Ammonia Quality Requirements**

| Parameter | Commercial Grade | Refrigeration Grade |
|---|---|---|
| Purity (min wt%) | 99.5 | 99.98 |
| Water (max wt%) | 0.5 | 0.015 |
| Inerts (max mL/g) | not specified | 0.1 |
| Oil (ppm by wt) | 5.0 | 3.0 |

## 2.5 ENVIRONMENTAL TARGETS

There are relatively few emission sources associated with the SNG and ammonia production plants. The primary emissions occur when the SNG is consumed, presumably in a combustion-type process, which is outside of the battery limit of this study. The plants in this study have several point source emissions, but consist mostly of fugitive emissions, which are also beyond the scope of this study. Typical emission sources include particulate matter (PM) from the coal pile, coal transfer operations, slag handling, and the cooling tower; fugitive emissions from valves and other process equipment; and emissions from emergency generators and flare stacks. None of these operations are included in the Aspen model, and emissions from these sources are not quantified.

There are some point source emissions that are modeled and accounted for in the various cases:

- Drying in the bituminous coal cases includes combustion of a slipstream of clean syngas and Claus plant tail gas. The hot flue gases are used to dry the coal and then exhausted to atmosphere.
- In the $NH_3$ co-production cases there is a waste heat boiler that is fueled by the PSA purification unit purge gas. Heat is recovered from the boiler exhaust gases prior to venting it to atmosphere.
- In all cases a portion of the lockhopper transport gas is vented to atmosphere.
- In all non-sequestration cases the $CO_2$-rich gas from the two-stage Selexol process is vented to atmosphere. This stream is compressed and stored in a saline aquifer in the carbon sequestration cases.
- In all non-sequestration cases the off gas from the PSA SNG purification unit is vented to atmosphere. This stream is compressed and stored in a saline aquifer in the carbon sequestration cases.

The only emissions reported for each case are sulfur dioxide ($SO_2$) and $CO_2$. In the PRB and lignite coal cases, there are no combustion operations so no $SO_2$ emissions are reported. PM in the SNG is not released until combustion occurs and fugitive dust sources are not quantified. Similarly, mercury remaining in the SNG is not released to the atmosphere until combustion occurs. Since the fuel in the two combustion operations (coal dryer in bituminous coal cases and waste heat boiler in $NH_3$ co-production cases) is void of nitrogen, and thermal oxides of nitrogen (NOx) formation is minimal at the temperatures of combustion, NOx is not quantified as an emission.

Since mercury is a methanation catalyst poison, essentially all of the mercury must be removed from the gas stream. The environmental target for mercury capture is greater than 90 percent. Based on experience at the Eastman Chemical plant, where synthesis gas (syngas) from a General Electric Energy (GEE) gasifier is treated, the actual mercury removal efficiency used is 95 percent. Sulfur-impregnated activated carbon is used by Eastman as the adsorbent in the packed beds operated at 30°C (86°F) and 6.2 megapascals (MPa) (900 pounds per square inch gage [psig]). Mercury removal between 90 and 95 percent has been reported with a bed life of 18 to 24 months. Removal efficiencies may be even higher, but at 95 percent the measurement precision limit was reached. Eastman has yet to experience any mercury contamination in its product [15]. Mercury removals of greater than 99 percent can be achieved by the use of dual beds, i.e., two beds in series. However, this study assumes that the use of sulfur-impregnated carbon in a single carbon bed achieves 95 percent reduction of mercury emissions, which reduces the Hg content in the clean syngas to less than 0.45 ppb in all cases.

Since momentum for carbon regulations continues to build, $CO_2$ emissions are also reported for each case. Because the form of emission limits, should they be imposed, is not known, $CO_2$ emissions are reported on a pounds per million British thermal units (lb/MMBtu) heat input to the gasifier basis as well as total tons per year. The emission values were used to quantify the impact of a carbon tax later in the report.

Ultimately, the environmental regulations for an SNG or SNG and ammonia co-production plant will be established by state and local jurisdictions. However, since the emissions profiles of these plants are relatively benign, it is assumed that environmental permits will be attainable.

## 2.6 CAPACITY FACTOR

Electric Power Research Institute (EPRI) examined the historical forced and scheduled outage times for IGCCs and concluded that the reliability factor (which looks at forced or unscheduled outage time only) for a single train IGCC (no spares) would be about 90 percent [16]. To get the AF, one has to deduct the scheduled outage time. In reality the scheduled outage time differs from gasifier technology-to-gasifier technology, but the differences are relatively small and would have minimal impact on the CF.

There are four operating IGCC's worldwide that use a solid feedstock and are primarily power producers (Polk, Wabash, Buggenum, and Puertollano). A 2006 report by Higman et al. examined the reliability of these IGCC power generation units and concluded that typical annual on-stream times are around 80 percent [17]. The CF would be somewhat less than the on-stream time since most plants operate at less than full load for some portion of the operating year.

Previous NETL studies utilizing IGCC technology have assumed a CF of 80 percent based on EPRI's studies and Higman's report. However, a 90 percent CF was assumed for the six gasifier gasification island used in this study, which has been supported by Siemens Energy, Inc. [18]. It was assumed that if one of the six gasifiers stopped operation for maintenance or repair, reducing the CF to 83 percent, the system could operate at 100 percent capacity long enough to compensate for the down time.

The addition of $CO_2$ capture or an ammonia co-production facility was assumed not to impact the AF, nor was the addition of compression and sequestration. This assumption was made to enable a comparison based on the impact of capital and variable operating costs only. Any reduction in assumed CF would further increase the FYCOP for the carbon sequestration cases and ammonia co-production cases as shown in the sensitivity analysis in Exhibit ES-9.

## 2.7 RAW WATER WITHDRAWAL AND CONSUMPTION

A water balance was performed for each case on the major water consumers in the process. The total water demand for each subsystem was determined and internal recycle water available from various sources like boiler feedwater (BFW) blowdown and condensate from syngas was applied to offset the water demand. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is the water removed from the ground or diverted from a surface-water source for use in the plant. Raw water consumption is also accounted for as the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source it was withdrawn from.

Raw water makeup was assumed to be provided 50 percent by a publicly owned treatment works (POTW) and 50 percent from groundwater. Raw water withdrawal is defined as the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, slag handling makeup and quench system makeup. The difference between withdrawal and process water returned to the source is consumption. Consumption represents the net impact of the process on the water source.

BFW blowdown and a portion of the sour water stripper blowdown were assumed to be treated and recycled to the cooling tower. The cooling tower blowdown and the balance of the SWS blowdown streams were assumed to be treated and 90 percent returned to the water source with the balance sent to the ash ponds for evaporation.

The largest consumer of raw water in all cases is cooling tower makeup. The bituminous coal cases utilize a mechanical draft, evaporative cooling tower, and all process blowdown streams were assumed to be treated and recycled to the cooling tower. It was assumed that the PRB and NDL coal cases utilized a parallel cooling system with half of the turbine exhaust steam condensed in an air-cooled condenser and half in a water-cooled condenser.

The design ambient wet bulb (WB) temperature of 11°C (51.5°F) at the Midwestern site was used to achieve a cooling water temperature of 16°C (60°F) using an approach of 5°C (8.5°F). The conditions of 3°C (37°F) at the Montana site and 2°C (36°F) at the North Dakota site (Exhibit 2-1) were used to achieve a cooling water temperature of 9°C (48°F) and 8°C (47°F), respectively at the two sites using an approach of 6°C (11°F). The cooling water range was assumed to be 11°C (20°F) for all cases. The cooling tower makeup rate was determined using the following:[19]

- Evaporative losses of 0.8 percent of the circulating water flow rate per 10°F of range
- Drift losses of 0.001 percent of the circulating water flow rate

- Blowdown losses were calculated as follows:
  - Blowdown Losses = Evaporative Losses / (Cycles of Concentration - 1)

    Where cycles of concentration are a measure of water quality and a mid-range value of 4 was chosen for this study.

The water balances presented in subsequent sections include the water demand of the major water consumers within the process, the amount provided by internal recycle, the amount of raw water withdrawal by difference, the amount of process water returned to the source and the raw water consumption, again by difference.

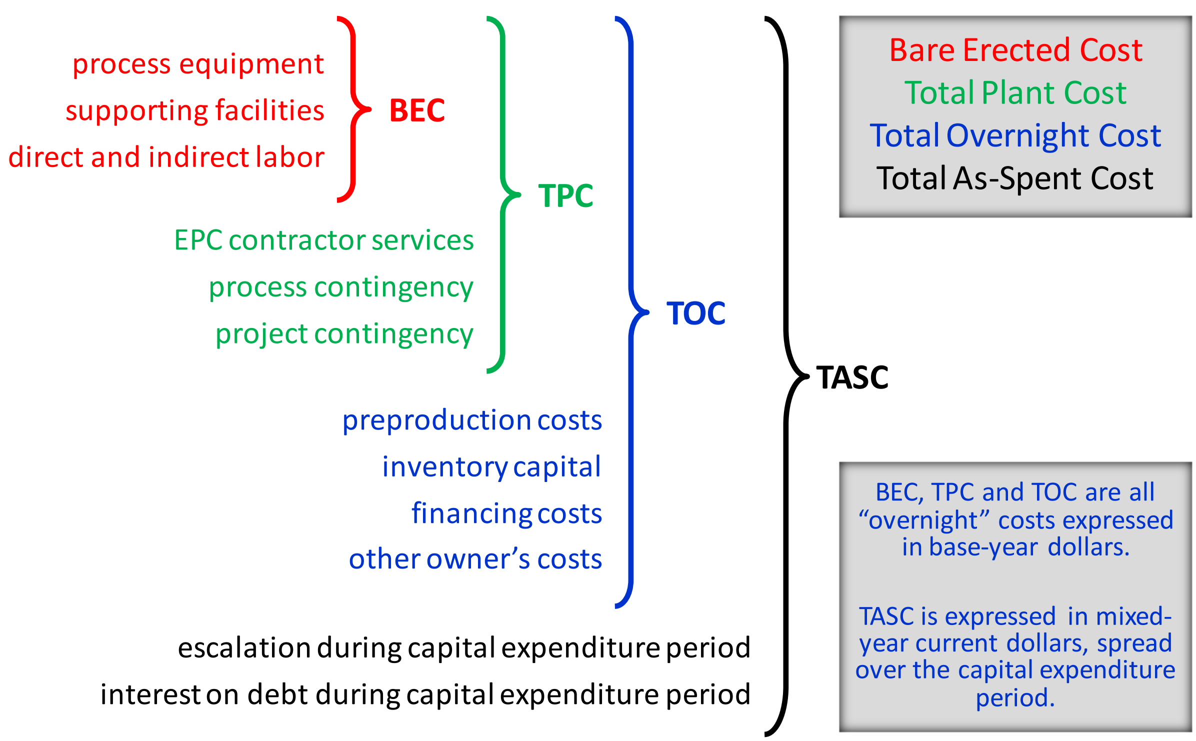
## 2.8 COST ESTIMATING METHODOLOGY

The estimating methodology for capital costs, operations and maintenance costs, and $CO_2$ TS&M costs are described below. The finance structure, basis for the discounted cash flow analysis, and first-year SNG cost calculations are also described.

### 2.8.1 Capital Costs

As illustrated in Exhibit 2-10, this study reports capital cost at four levels: Bare Erected Cost (BEC), Total Plant Cost (TPC), Total Overnight Cost (TOC) and Total As-spent Capital (TASC). BEC, TPC and TOC are "overnight" costs and are expressed in "base-year" dollars. The base year is the first year of capital expenditure, which for this study is assumed to be 2007. TASC is expressed in mixed-year, current-year dollars over the entire capital expenditure period, which is assumed to last five years (2007 to 2011).

**Exhibit 2-10 Capital Cost Levels and their Elements**



The BEC comprises the cost of process equipment, on-site facilities and infrastructure that support the plant (e.g., shops, offices, labs, road), and the direct and indirect labor required for its construction and/or installation. The cost of EPC services and contingencies is not included in BEC. BEC is an overnight cost expressed in base-year (2007) dollars.

The TPC comprises the BEC plus the cost of services provided by the engineering, procurement and construction (EPC) contractor and project and process contingencies. EPC services include: detailed design, contractor permitting (i.e., those permits that individual contractors must obtain to perform their scopes of work, as opposed to project permitting, which is not included here), and project/construction management costs. TPC is an overnight cost expressed in base-year (2007) dollars.

The TOC comprises the TPC plus owner's costs. TOC is an "overnight" cost, expressed in base-year (2007) dollars and as such does not include escalation during construction or interest during construction. TOC is an overnight cost expressed in base-year (2007) dollars.

The TASC is the sum of all capital expenditures as they are incurred during the capital expenditure period including their escalation. TASC also includes interest during construction. Accordingly, TASC is expressed in mixed, current-year dollars over the capital expenditure period.

**Cost Estimate Basis and Classification**

The TPC and Operation and Maintenance (O&M) costs for each of the cases in the study were estimated by WorleyParsons using an in-house database and conceptual estimating models. Costs were further calibrated using a combination of adjusted vendor-furnished and actual cost data from recent design projects.

The capital costs for each cost account were reviewed by comparing individual accounts across each of the cases to ensure an accurate representation of the relative cost differences between the cases and accounts.

Recommended Practice 18R-97 of the Association for the Advancement of Cost Engineering International (AACE) describes a Cost Estimate Classification System as applied in Engineering, Procurement and Construction for the process industries [20].

Most techno-economic studies, including this coal-to-SNG study, completed by NETL feature cost estimates intended for the purpose of a "Feasibility Study" (AACE Class 4). Exhibit 2-11 describes the characteristics of an AACE Class 4 Cost Estimate. Cost estimates in this study have an expected accuracy range of -15%/+30%.

**System Code-of-Accounts**

The costs are grouped according to a process/system oriented code of accounts. This type of code-of-account structure has the advantage of grouping all reasonably allocable components of a system or process so they are included in the specific system account. (This would not be the case had a facility, area, or commodity account structure been chosen instead).

**Exhibit 2-11 Features of an AACE Class 4 Cost Estimate**

| Project Definition | Typical Engineering Completed | Expected Accuracy |
|---|---|---|
| 1 to 15% | plant capacity, block schematics, indicated layout, process flow diagrams for main process systems, and preliminary engineered process and utility equipment lists | -15% to -30% on the low side, and +20% to +50% on the high side |

### Gasification Plant Maturity

The gasification production plant is based on commercial offerings; however, there have been very limited sales of these units. The costs listed reflect the "next commercial offering" level of cost rather than mature n$^{th}$-of-a-kind (NOAK) cost. Thus, each of these cases reflects the expected cost for the next commercial sale of the gasification technology.

### $CO_2$ Removal Maturity

Pre-combustion $CO_2$ removal from syngas streams has been proven in chemical processes with similar conditions to that in gasification plants in this study, and has been demonstrated at the Great Plains Synfuels plan and in hundreds of syngas processing plants such as ammonia, methanol, hydrogen, and other petrochemical plants worldwide.

### Contracting Strategy

The estimates are based on an EPCM approach utilizing multiple subcontracts. This approach provides the Owner with greater control of the project, while minimizing, if not eliminating most of the risk premiums typically included in an Engineer/Procure/Construct (EPC) contract price.

In a traditional lump sum EPC contract, the Contractor assumes significant financial risk for performance, schedule, and cost up to stipulated contractual levels. However, as a result of current market conditions, EPC contractors appear more unwilling to accept significant project completion and performance risk. Rather, the current trend appears to be a modified EPC approach where much of the risk remains with the Owner. Where Contractors are willing to accept the risk in EPC type lump-sum arrangements, it is reflected in the project cost. In today's market, Contractor premiums for accepting these risks, particularly performance risk, can be substantial and increase the overall project costs dramatically.

The EPCM approach used as the basis for the estimates here is anticipated to be the most likely contracting model for SNG/ammonia projects today. While the Owner retains the risks, the risks become reduced with time, as there is better scope definition at the time of contract award(s).

### Estimate Scope

The estimates represent a complete synfuels production facility, and in the co-production cases, a complete $NH_3$ production facility on a generic site.

The plant boundary limit is defined as the total plant facility within the "fence line" including coal receiving and water supply system, but terminating at the high voltage side of the main power transformers and the compression of the SNG. The single exception to the fence line limit is in the carbon sequestration cases where costs are included for TS&M of the $CO_2$.

## Capital Costs

WorleyParsons developed the capital cost estimates for each plant using the company's in-house database and conceptual estimating models for each of the specific technologies. This database and the respective models are maintained by WorleyParsons as part of a commercial gasification design base of experience for similar equipment in the company's range of power and process projects. A reference bottoms-up estimate for each major system provides the basis for the estimating models.

Other key estimate considerations include the following:

- Labor costs are based on Midwest, Merit Shop. The estimating models are based on U.S. Gulf Coast and the labor has been factored to Midwest. The basis for the factors is the PAS, Inc. (PAS) "Merit Shop Wage & Benefit Survey," which is published annually. Based on the data provided in PAS, WorleyParsons used the weighted average payroll plus fringe rate for a standard craft distribution as developed for the estimating models. PAS presents information for eight separate regions. For this study, Region 5 (IL, IN, MI, MN, OH, and WI) was selected. The weighted average rate for Region 8 (CO, MT, ND, SD, UT, and WY) is within one-half of one percent of that for Region 5. Given the accuracy limits of the studies, the differential is insignificant.
- The estimates are based on a competitive bidding environment, with adequate skilled craft labor available locally.
- Labor is based on a 50-hour work-week (5-10s). No additional incentives such as per-diems or bonuses have been included to attract craft labor.
- While not included at this time, labor incentives may ultimately be required to attract and retain skilled labor depending on the amount of competing work in the region, and the availability of skilled craft in the area at the time the projects proceed to construction.
- The estimates are based on a greenfield sites.
- The sites are considered to be Seismic Zone 1, relatively level, and free from hazardous materials, archeological artifacts, or excessive rock. Soil conditions are considered adequate for spread footing foundations. The soil bearing capability is assumed adequate such that piling is not needed to support the foundation loads.
- Costs are limited to within the "fence line," terminating at the high voltage side of the main power transformers with the exception of costs included for TS&M of $CO_2$ in all carbon sequestration cases.
- Engineering and Construction Management are estimated at 10 percent of BEC. These costs consist of all home office engineering and procurement services as well as field construction management costs. Site staffing generally includes a construction manager, resident engineer, scheduler, and personnel for project controls, document control, materials management, site safety, and field inspection.
- All capital costs are presented as "Overnight Costs" in June 2007 dollars. Escalation to period-of-performance is specifically excluded in the TPC. However, the CCF used to calculate FYCOP includes capital cost escalation during construction of 3.6 percent.

### Price Fluctuations

During the course of this study, the prices of equipment and bulk materials fluctuated quite substantially. Some reference quotes pre-dated the 2007 year cost basis while others were received post-2007. All vendor quotes used to develop these estimates were adjusted to June 2007 dollars accounting for the price fluctuations. Adjustments of costs pre-dating 2007 benefitted from a vendor survey of actual and projected pricing increases from 2004 through mid-2007 that WorleyParsons conducted for another project. The results of that survey were used to validate/recalibrate the corresponding escalation factors used in the conceptual estimating models. The more recent economic down turn has resulted in a reduction of commodity prices such that many price indices have similar values in January 2010 compared to June 2007. For example, the Chemical Engineering Plant Cost Index was 532.7 in June 2007 and 532.9 in January 2010, and the Gross Domestic Product Chain-type Price Index was 106.7 on July 1, 2007 and 110.0 on January 1, 2010. While these overall indices are nearly constant, it should be noted that the cost of individual equipment types may still deviate from the June 2007 reference point.

### Exclusions

The capital cost estimate includes all anticipated costs for equipment and materials, installation labor, professional services (Engineering and Construction Management), and contingency. The following items are excluded from the capital costs:

- Escalation to period-of-performance
- Owner's costs – these are accounted for separately and are described below.
- All taxes, with the exception of payroll and property taxes (property taxes are included with the fixed O&M costs)
- Site specific considerations – including, but not limited to, seismic zone, accessibility, local regulatory requirements, excessive rock, piles, laydown space, etc.
- Labor incentives in excess of a 5-10s
- Additional premiums associated with an EPC contracting approach

### Contingency

Process and project contingencies are included in estimates to account for unknown costs that are omitted or unforeseen due to a lack of complete project definition and engineering. Contingencies are added because experience has shown that such costs are likely, and expected, to be incurred even though they cannot be explicitly determined at the time the estimate is prepared.

Capital cost contingencies do not cover uncertainties or risks associated with

- scope changes
- changes in labor availability or productivity
- delays in equipment deliveries
- changes in regulatory requirements
- unexpected cost escalation
- performance of the plant after startup (e.g., availability, efficiency)

**Process Contingency**

Process contingency is intended to compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of a technology. Process contingencies are applied to each plant section based on its current technology status.

As shown in Exhibit 2-12, AACE International Recommended Practice 16R-90 provides guidelines for estimating process contingency based on EPRI philosophy [21].

Process contingencies have been applied to the estimates in this study as follows:

- Gasifiers and Syngas Coolers – 15 percent on all cases – next-generation commercial offering
- Two Stage Selexol – 20 percent on all cases - unproven technology at commercial scale
- Methanation Process – 10 percent on all cases – unproven technology at commercial scale
- Mercury Removal – 5 percent on all cases – minimal commercial scale experience in gasification applications
- Instrumentation and Controls – 5 percent for integration issues

**Exhibit 2-12 AACE Guidelines for Process Contingency**

| Technology Status | Process Contingency (% of Associated Process Capital) |
|---|---|
| New concept with limited data | 40+ |
| Concept with bench-scale data | 30-70 |
| Small pilot plant data | 20-35 |
| Full-sized modules have been operated | 5-20 |
| Process is used commercially | 0-10 |

Process contingency is typically not applied to costs that are set equal to a research goal or programmatic target since these values presume to reflect the total cost.

**Project Contingency**

AACE 16R-90 states that project contingency for a “budget-type” estimate (AACE Class 4 or 5) should be 15 to 30 percent of the sum of BEC, EPC fees and process contingency. This was used as a general guideline, but some project contingency values outside of this range occur based on WorleyParsons’ in-house experience.

**Owner's Costs**

Exhibit 2-13 explains the estimation method for owner's costs. With some exceptions, the estimation method follows guidelines in Sections 12.4.7 to 12.4.12 of AACE International Recommended Practice No. 16R-90 [21]. The Electric Power Research Institute's "Technical Assessment Guide (TAG®) – Power Generation and Storage Technology Options" also has guidelines for estimating owner's costs. The EPRI and AACE guidelines are very similar. In instances where they differ, this study has sometimes adopted the EPRI approach.

**Exhibit 2-13 Owner's Costs Included in TOC**

| Owner's Cost | Estimate Basis |
|---|---|
| **Prepaid Royalties** | Any technology royalties are assumed to be included in the associated equipment cost, and thus are not included as an owner's cost. |
| **Preproduction (Start-Up) Costs** | • 6 months operating labor<br>• 1 month maintenance materials at full capacity<br>• 1 month non-fuel consumables at full capacity<br>• 1 month waste disposal<br>• 25% of one month's fuel cost at full capacity<br>• 2% of TPC<br>Compared to AACE 16R-90, this includes additional costs for operating labor (6 months versus 1 month) to cover the cost of training the plant operators, including their participation in startup, and involving them occasionally during the design and construction. AACE 16R-90 and EPRI TAG® differ on the amount of fuel cost to include; this estimate follows EPRI. |
| **Working Capital** | Although inventory capital (see below) is accounted for, no additional costs are included for working capital. |
| **Inventory Capital** | • 0.5% of TPC for spare parts<br>• 60 day supply (at full capacity) of fuel. Not applicable for natural gas.<br>• 60 day supply (at full capacity) of non-fuel consumables (e.g., chemicals and catalysts) that are stored on site. Does not include catalysts and adsorbents that are batch replacements such as WGS, COS, and SCR catalysts and activated carbon.<br>AACE 16R-90 does not include an inventory cost for fuel, but EPRI TAG® does. |
| **Land** | • $3,000/acre (300 acres for IGCC and PC, 100 acres for NGCC) |
| **Financing Cost** | • 2.7% of TPC<br>This financing cost (not included by AACE 16R-90) covers the cost of securing financing, including fees and closing costs but not including interest during construction (or AFUDC). The "rule of thumb" estimate (2.7% of TPC) is based on a 2008 private communication with a capital services firm. |

| Owner's Cost | Estimate Basis |
|---|---|
| **Other Owner's Costs** | • 15% of TPC<br><br>This additional lumped cost is not included by AACE 16R-90 or EPRI TAG®. The "rule of thumb" estimate (15% of TPC) is based on a 2009 private communication with WorleyParsons. Significant deviation from this value is possible as it is very site and owner specific. The lumped cost includes:<br><br>- Preliminary feasibility studies, including a Front-End Engineering Design (FEED) study<br>- Economic development (costs for incentivizing local collaboration and support)<br>- Construction and/or improvement of roads and/or railroad spurs outside of site boundary<br>- Legal fees<br>- Permitting costs<br>- Owner's engineering (staff paid by owner to give third-party advice and to help the owner oversee/evaluate the work of the EPC contractor and other contractors)<br>- Owner's contingency (Sometimes called "management reserve", these are funds to cover costs relating to delayed startup, fluctuations in equipment costs, unplanned labor incentives in excess of a five-day/ten-hour-per-day work week. Owner's contingency is NOT a part of project contingency.)<br><br>This lumped cost does NOT include:<br><br>- EPC Risk Premiums (Costs estimates are based on an Engineering Procurement Construction Management approach utilizing multiple subcontracts, in which the owner assumes project risks for performance, schedule and cost)<br>- Transmission interconnection: the cost of interconnecting with power transmission infrastructure beyond the plant busbar.<br>- Taxes on capital costs: all capital costs are assumed to be exempt from state and local taxes.<br>- Unusual site improvements: normal costs associated with improvements to the plant site are included in the bare erected cost, assuming that the site is level and requires no environmental remediation. Unusual costs associated with the following design parameters are excluded: flood plain considerations, existing soil/site conditions, water discharges and reuse, rainfall/snowfall criteria, seismic design, buildings/enclosures, fire protection, local code height requirements, noise regulations. |

### 2.8.2 Operations and Maintenance Costs

The production costs or operating costs and related maintenance expenses (O&M) pertain to those charges associated with operating and maintaining the plants over their expected life. These costs include:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel
- Waste disposal
- Co-product or by-product credit (that is, a negative cost for any by-products sold)

There are two components of O&M costs; fixed O&M, which is independent of product production, and variable O&M, which is proportional to product production.

**Operating Labor**

Operating labor cost was determined based on of the number of operators required for each specific case. The average base labor rate used to determine annual cost is $34.65/hour (hr) in 2007 dollars. The associated labor burden is estimated at 30 percent of the base labor rate.

All cases have two skilled operators, one foreman, and three laboratory technician positions. The co-production plants have 13 operator positions, whereas the SNG only plants have 12 operator positions. This differs from Volumes 1 and 3 of the Baseline Series, which use nine operator positions for the non-capture cases and ten for the capture cases. The number of the other labor personnel remains the same.

**Maintenance Material and Labor**

Maintenance cost was evaluated on the basis of relationships of maintenance cost to initial capital cost. This represents a weighted analysis in which the individual cost relationships were considered for each major plant component or section.

**Administrative and Support Labor**

Labor administration and overhead charges are assessed at a rate of 25 percent of the burdened O&M labor.

**Consumables**

The cost of consumables, including fuel, was determined on the basis of individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours.

Quantities for major consumables such as fuel were taken from technology-specific heat and mass balance diagrams developed for each plant application. Other consumables were evaluated on the basis of the quantity required using reference data.

The quantities for initial fills and daily consumables were calculated on a 100 percent operating capacity basis. The annual cost for the daily consumables was then adjusted to incorporate the annual plant operating basis, or CF.

Initial fills of the consumables, fuels and chemicals, are different from the initial chemical loadings, which are included with the equipment pricing in the capital cost.

**Waste Disposal**

Waste quantities and disposal costs were determined similarly to the consumables. The slag and spent ZnO sorbent from each case are considered wastes with disposal costs of $17.89/tonne ($16.23/ton). The carbon used for mercury control is considered a hazardous waste with disposal cost of $919/tonne ($834/ton).

**Co-Products and By-Products**

The primary products in this study are SNG in all cases and ammonia in the co-production cases. The by-products include sulfur, electricity, and potentially $CO_2$. Due to the variable marketability of the sulfur, no credit or disposal cost was used. The electricity produced in excess of internal requirements was assumed sold on the open market at a price of $58/megawatt-hour (MWh). $CO_2$ was assumed to be a waste product that is either vented to atmosphere or compressed and sequestered with the associated costs. A sensitivity analysis was done to determine the economic impact of selling the $CO_2$ for enhanced oil recovery (EOR) at one extreme and debiting the project for $CO_2$ emissions assuming a carbon tax at the other extreme.

It should be noted that by-product credits and/or disposal costs could potentially be an additional determining factor in the choice of technology for some companies and in selecting some sites. A high local value of the product can establish whether or not added capital should be included in the plant costs to produce a particular co-product. Slag is a potential by-product in certain markets and would have potential marketability. However, as stated above, the slag is considered waste in this study with a concomitant disposal cost.

### 2.8.3 $CO_2$ Transport, Storage, and Monitoring

For those cases that feature carbon sequestration, the capital and operating costs for $CO_2$ TS&M were independently estimated by NETL. Those costs were converted to a FYCOP and combined with the plant capital and operating costs to produce an overall FYCOP. The TS&M costs were initially levelized over a 30-year period using the methodology described in the next subsection of this report, and subsequently converted to first year costs by dividing by the levelization factor.

$CO_2$ TS&M costs were estimated based on the following assumptions:

- $CO_2$ is supplied to the pipeline at the plant fence line at a pressure of 15.3 MPa (2,215 psia). The $CO_2$ product gas composition varies in the cases presented, but is expected to meet the specification described in Exhibit 2-14. A glycol dryer located near the mid-point of the compression train is used to meet the moisture specification.

**Exhibit 2-14 $CO_2$ Pipeline Specification**

| Parameter | Units | Parameter Value |
|---|---|---|
| Inlet Pressure | MPa (psia) | 15.3 (2,215) |
| Outlet Pressure | MPa (psia) | 10.4 (1,515) |
| Inlet Temperature | °C (°F) | 35 (95) |
| $N_2$ Concentration | ppmv | < 300 |
| $O_2$ Concentration | ppmv | < 40 |
| Ar Concentration | ppmv | < 10 |
| $H_2O$ Concentration | ppmv | < 150 |

- The $CO_2$ is transported 80 kilometers (km) (50 miles) via pipeline to a geologic sequestration field for injection into a saline formation.
- The $CO_2$ is transported and injected as a supercritical (SC) fluid in order to avoid two-phase flow and achieve maximum efficiency [22]. The pipeline is assumed to have an outlet pressure (above the SC pressure) of 8.3 MPa (1,200 psia) with no recompression along the way. Accordingly, $CO_2$ flow in the pipeline was modeled to determine the pipe diameter that results in a pressure drop of 6.9 MPa (1,000 psi) over an 80 km (50 mile) pipeline length [23]. (Although not explored in this study, the use of boost compressors and a smaller pipeline diameter could possibly reduce capital costs for sufficiently long pipelines.) The diameter of the injection pipe will be of sufficient size that frictional losses during injection are minimal and no booster compression is required at the well-head in order to achieve an appropriate down-hole pressure, with hydrostatic head making up the difference between the injection and reservoir pressure.
- The saline formation is at a depth of 1,239 m (4,055 ft) and has a permeability of 22 millidarcy (md) (22 $\mu m^2$) and formation pressure of 8.4 MPa (1,220 psig) [24]. This is considered an average storage site and requires roughly one injection well for each 9,360 tonnes (10,320 tons) of $CO_2$ injected per day [24]. The assumed aquifer characteristics are tabulated in Exhibit 2-15.

The cost metrics utilized in this study provide a best estimate of TS&M costs for a "typical" sequestration project, and may vary significantly based on variables such as terrain to be crossed by the pipeline, reservoir characteristics, and number of land owners from which sub-surface rights must be acquired. Raw capital and operating costs are derived from detailed cost metrics found in the literature, escalated to June 2007-year dollars using appropriate price indices. These costs were then verified against values quoted by industrial sources where possible. Where regulatory uncertainty exists or costs are undefined, such as liability costs and the acquisition of underground pore volume, analogous existing policies were used for representative cost scenarios.

**Exhibit 2-15 Deep Saline Aquifer Specification**

| Parameter | Units | Base Case |
|---|---|---|
| Pressure | MPa (psi) | 8.4 (1,220) |
| Thickness | m (ft) | 161 (530) |
| Depth | m (ft) | 1,236 (4,055) |
| Permeability | md ($\mu m^2$) | 22 (22) |
| Pipeline Distance | Km (miles) | 80 (50) |
| Injection Rate per Well | Tonne (ton) $CO_2$/day | 9,360 (10,320) |

The following subsections describe the sources and methodology used for each metric.

**Cost Levelization and Conversion to First Year Costs**

Capital and operating costs were levelized over a 35-year period (30-year operational period plus 5-year capital expenditure period) and include both a 20 percent process contingency and 30 percent project contingency.

In several areas, such as Pore Volume Acquisition, Monitoring, and Liability, cost outlays occur over a longer time period, up to 100 years. In these cases a capital fund is established based on the net present value of the cost outlay, and this fund is then levelized similar to the other costs.

First year costs are calculated by dividing the levelized cost by the levelization factor. It is the first year cost that is used in this study.

**Transport Costs**

$CO_2$ transport costs are divided into three categories: pipeline costs, related capital expenditures, and O&M costs.

Pipeline costs are derived from data published in the Oil and Gas Journal's (O&GJ) annual Pipeline Economics Report for existing natural gas, oil, and petroleum pipeline project costs from 1991 to 2003. These costs are expected to be analogous to the cost of building a $CO_2$ pipeline, as noted in various studies [22,24,25]. The University of California performed a regression analysis to generate cost curves from the O&GJ data for: (1) Pipeline Materials, (2) Direct Labor, (3) Indirect Costs, and (4) Right-of-way acquisition, with each represented as a function of pipeline length and diameter [25]. These cost curves were escalated to the June 2007 year dollars used in this study.

Related capital expenditures were based on the findings of a previous study funded by NETL, Carbon Dioxide Sequestration in Saline Formations – Engineering and Economic Assessment [24]. This study utilized a similar basis for pipeline costs (O&GJ Pipeline cost data up to the year 2000), but added a $CO_2$ surge tank and pipeline control system to the project.

Transport O&M costs were assessed using metrics published in a second NETL sponsored report entitled Economic Evaluation of $CO_2$ Storage and Sink Enhancement Options [22]. This study was chosen due to the reporting of O&M costs in terms of pipeline length, whereas the other studies mentioned above either (a) do not report operating costs, or (b) report them in absolute terms for one pipeline, as opposed to a length- or diameter-based metric.

**Storage Costs**

Storage costs were divided into five categories: (1) Site Screening and Evaluation, (2) Injection Wells, (3) Injection Equipment, (4) O&M Costs, and (5) Pore Volume Acquisition. With the exception of Pore Volume Acquisition, all of the costs were obtained from Economic Evaluation of $CO_2$ Storage and Sink Enhancement Options [22]. These costs include all of the costs associated with determining, developing, and maintaining a $CO_2$ storage location, including site evaluation, well drilling, and the capital equipment required for distributing and injecting $CO_2$.

Pore Volume Acquisition costs are the costs associated with acquiring rights to use the sub-surface volume where the $CO_2$ will be stored, i.e. the pore space in the geologic formation. These costs were based on recent research by Carnegie Mellon University, which examined existing sub-surface rights acquisition as it pertains to natural gas storage [26]. The regulatory uncertainty in this area combined with unknowns regarding the number and type (private or government) of property owners, require a number of "best engineering judgment" decisions to be made. In this study it was assumed that long-term lease rights were acquired from the property owners in the projected $CO_2$ plume growth region for a nominal fee, and that an annual "rent" was paid when the plume reached each individual acre of their property for a period of up to 100 years from the injection start date. The present value of the life cycle pore volume costs are assessed at a 10 percent discount rate and a capital fund is set up to pay for these costs over the 100 year rent scenario.

**Liability Protection**

Liability Protection addresses the fact that if damages are caused by injection and long-term storage of $CO_2$, the injecting party may bear financial liability. Several types of liability protection schemes have been suggested for $CO_2$ storage, including Bonding, Insurance, and Federal Compensation Systems combined with either tort law (as with the Trans-Alaska Pipeline Fund), or with damage caps and preemption, as is used for nuclear energy under the Price Anderson Act [27]. However, at present, a specific liability regime has yet to be dictated either at a Federal or (to our knowledge) State level. However, certain state governments have enacted legislation, which assigns liability to the injecting party, either in perpetuity (Wyoming) or until ten years after the cessation of injection operations, pending reservoir integrity certification, at which time liability is turned over to the state (North Dakota and Louisiana) [28,29,30]. In the case of Louisiana, a trust fund totaling five million dollars is established over the first ten years (120 months) of injection operations for each injector. This fund is then used by the state for $CO_2$ monitoring and, in the event of an at-fault incident, damage payments.

Liability costs assume that a bond must be purchased before injection operations are permitted in order to establish the ability and good will of an injector to address damages where they are deemed liable. A figure of five million dollars was used for the bond based on the Louisiana fund level. This bond level may be conservatively high, in that the Louisiana fund covers both liability and monitoring, but that fund also pertains to a certified reservoir where injection operations have ceased, having a reduced risk compared to active operations. The bond cost was not escalated.

**Monitoring Costs**

Monitoring costs were evaluated based on the methodology set forth in the IEA Greenhouse Gas (GHG) R&D Programme's Overview of Monitoring Projects for Geologic Storage Projects report [31]. In this scenario, operational monitoring of the $CO_2$ plume occurs over thirty years (during plant operation) and closure monitoring occurs for the following fifty years (for a total of eighty years). Monitoring is via electromagnetic (EM) survey, gravity survey, and periodic seismic survey; EM and gravity surveys are ongoing while seismic survey occurs in years 1, 2, 5, 10, 15, 20, 25, and 30 during the operational period, then in years 40, 50, 60, 70, and 80 after injection ceases.

### 2.8.4 Finance Structure, Discounted Cash Flow Analysis, and First-Year Cost of SNG

The global economic assumptions are listed in Exhibit 2-16.

The finance structure was chosen based on the assumption of a high-risk fuels project. Exhibit 2-17 describes the finance structure that was assumed. This finance structure was based on information presented in a 2008 NETL report, "Recommended Project Finance Structures for the Economic Analysis of Fossil-Based Energy Projects", based on interviews with project developers/owners, financial organizations and law firms.

**Exhibit 2-16 Global Economic Assumptions**

| Parameter | Value |
|---|---|
| **TAXES** | |
| Income Tax Rate | 38% (Effective 34% Federal, 6% State) |
| Capital Depreciation | 20 years, 150% declining balance |
| Investment Tax Credit | 0% |
| Tax Holiday | 0 years |
| **CONTRACTING AND FINANCING TERMS** | |
| Contracting Strategy | Engineering Procurement Construction Management (owner assumes project risks for performance, schedule and cost) |
| Type of Debt Financing | Non-Recourse (collateral that secures debt is limited to the real assets of the project) |
| Repayment Term of Debt | 15 years |
| Grace Period on Debt Repayment | 0 years |
| Debt Reserve Fund | None |
| **ANALYSIS TIME PERIODS** | |
| Capital Expenditure Period | 5 Years |
| Operational Period | 30 years |
| Economic Analysis Period (used for IRROE) | 35 Years (capital expenditure period plus operational period) |

| Parameter | Value |
|---|---|
| **TREATMENT OF CAPITAL COSTS** | |
| Capital Cost Escalation During Capital Expenditure Period (nominal annual rate) | 3.6%[1] |
| Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation) | 10%, 30%, 25%, 20%, 15% |
| Working Capital | zero for all parameters |
| % of Total Overnight Capital that is Depreciated | 100% *(this assumption introduces a very small error even if a substantial amount of TOC is actually non-depreciable)* |
| **ESCALATION OF OPERATING REVENUES AND COSTS** | |
| Escalation of COP (revenue), O&M Costs, and Fuel Costs (nominal annual rate) | 3.0%[2] |

**Exhibit 2-17 Financial Structure for High Risk Fuels Projects**

| Type of Security | % of Total | Current (Nominal) Dollar Cost | Weighted Current (Nominal) Cost | After Tax Weighted Cost of Capital |
|---|---|---|---|---|
| Debt | 50 | 9.5% | 4.75% | |
| Equity | 50 | 20% | 10% | |
| Total | 100 | N/A | 14.75% | 12.94% |

The NETL Power Systems Financial Model (PSFM) is a nominal-dollar[3] (current dollar) discounted cash flow (DCF) analysis tool [9]. As explained below, the PSFM was used to calculate first-year cost of production (FYCOP)[4].

- The FYCOP is the revenue received by the generator per net unit of production (bbl, MMBtu, kg, etc.) *during the plant's first year of operation*. To calculate the first-year COP, the PSFM was used to determine a "base-year" (2007) COP that, when escalated at

[1] A nominal average annual rate of 3.6% is assumed for escalation of capital costs during construction. This rate is equivalent to the nominal average annual escalation rate for process plant construction costs between 1947 and 2008 according to the *Chemical Engineering* Plant Cost Index.

[2] An average annual inflation rate of 3.0% is assumed. This rate is equivalent to the average annual escalation rate between 1947 and 2008 for the U.S. Department of Labor's Producer Price Index for Finished Goods, the so-called "headline" index of the various Producer Price Indices. (The Producer Price Index for the Electric Power Generation Industry may be more applicable, but that data does not provide a long-term historical perspective since it only dates back to December 2003.)

[3] Since the analysis takes into account taxes and depreciation, a nominal dollar basis is preferred to properly reflect the interplay between depreciation and inflation.

[4] For this calculation, "cost of product" is somewhat of a misnomer because from the production plant's perspective it is actually the "price" received for the product generated. However, since the price paid to the plant for the product is ultimately charged to the end user, from the customer's perspective it is part of the cost of product.

an assumed nominal annual general inflation rate of 3 percent[5], provided the stipulated internal rate of return on equity over the entire economic analysis period (capital expenditure period plus thirty years of operation). Since this analysis assumes that COP increases over the economic analysis period at the nominal annual general inflation rate, it remains constant in real terms and the first-year COP is equivalent to the base-year COP when expressed in base-year (2007) dollars.

Since 2007 is the first year of the capital expenditure period it is also the base year for the economic analysis. Accordingly, all costs are expressed in base-year (June 2007) dollars except for TASC, which is expressed in mixed-year, current dollars.

All capital costs included in this analysis, including project development and construction costs, are assumed to be incurred during the capital expenditure period. The plants are assumed to have a capital expenditure period of five years. Since the plants begin expending capital in the base year (2007), this means that the analysis assumes that they begin operating in 2012. In addition to the capital expenditure period, the economic analysis considers thirty years of operation.

**Estimating First-Year COE with Capital Charge Factors**

For scenarios that adhere to the global economic assumptions listed in Exhibit 2-16 and utilize the finance structure listed in Exhibit 2-17, the following simplified equation can be used to estimate first-year COP as a function of TOC, fixed O&M, variable O&M (including fuel), capacity factor and net production. The equation requires the application of the capital charge factor (CCF) listed in Exhibit 2-18. This first year CCF is valid only for the global economic assumptions listed in Exhibit 2-16, the stated finance structure, and the stated capital expenditure period.

**Exhibit 2-18 Economic Parameters for FYCOP Calculation**

| | **High Risk Fuels Financial Structure** |
|---|---|
| First Year Capital Charge Factor | 0.2449 |

All factors in the first-year COP equation are expressed in base-year dollars. The base year is the first year of capital expenditure, which for this study is assumed to be 2007. As shown in Exhibit 2-16, all factors (COP, O&M and fuel) are assumed to escalate at a nominal annual

[5] This nominal escalation rate is equal to the average annual inflation rate between 1947 and 2008 for the U.S. Department of Labor's Producer Price Index for Finished Goods. This index was used instead of the Producer Price Index for the Electric Power Generation Industry because the Electric Power Index only dates back to December 2003 and the Producer Price Index is considered the "headline" index for all of the various Producer Price Indices.

general inflation rate of 3.0 percent. Accordingly, all first-year costs (COP and O&M) are equivalent to base-year costs when expressed in base-year (2007) dollars.

$$First\ Year\ COP = \frac{\begin{matrix} first\ year \\ capital\ charge \end{matrix} + \begin{matrix} first\ year \\ fixed\ operating \\ costs \end{matrix} + \begin{matrix} first\ year \\ variable\ operating \\ costs \end{matrix}}{\begin{matrix} annual\ quantity \\ of\ product\ generated \end{matrix}}$$

$$First\ Year\ COP = \frac{(CCF)(TOC) + OC_{FIX} + (CF)(OC_{VAR})}{(CF)(Qty)}$$

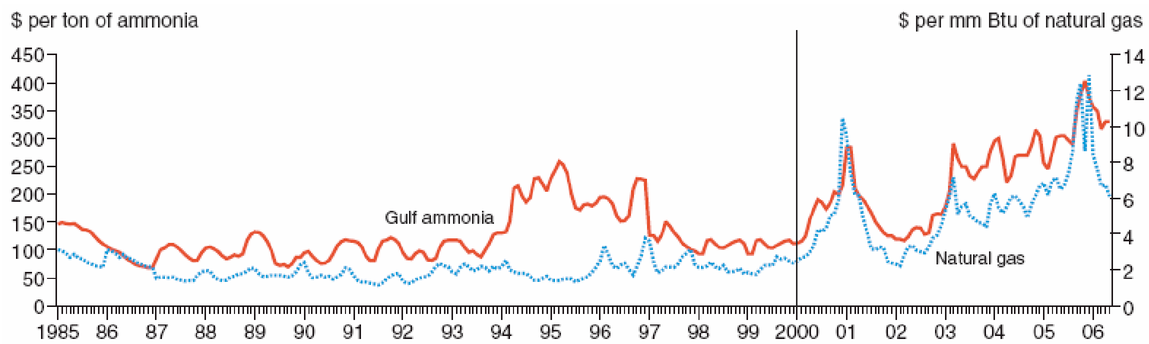
where:

| | |
|---|---|
| First Year COP= | current-dollar first-year cost of product *expressed in base-year dollars per unit of product* ($/bbl, $/kg, $/MMBtu, etc.)) |
| CCF = | capital charge factor taken from Exhibit 2-18 that matches the applicable finance structure and capital expenditure period |
| TOC = | total overnight capital*, expressed in base-year dollars* |
| $OC_{FIX}$ = | the sum of all fixed annual operating costs*, expressed in base-year dollars* |
| $OC_{VAR}$ = | the sum of all variable annual operating costs, including fuel at 100 percent capacity factor*, expressed in base-year dollars* |
| CF = | plant capacity factor, assumed to be constant over the operational period |
| Qty = | annual quantity of product generated at 100 percent capacity factor |

### 2.8.5 Ammonia Price

Historically, the price of ammonia has been predominantly dependent on ammonia demand. However, recent volatility of natural gas prices has caused the price of ammonia to become increasingly dependent on the cost of natural gas. A report issued by the USDA found that the price correlation between natural gas and ammonia was weak at 0.17 to 0.07 prior to 2000, but became strong at 0.8 to 0.7 as natural gas prices rose after 2000 [32]. The same report states that from 2000 to 2006, rising natural gas prices resulted in a decrease of U.S. ammonia production by 44 percent while ammonia prices paid by farmers increased by 130 percent. Exhibit 2-19 shows historic prices of natural gas and ammonia [32]. This trend is expected to continue as natural gas prices are expected to rise. Exhibit 2-20 shows AEO 2009 price projections to 2030 for residential, industrial and electric power natural gas users along with high and low economic growth scenarios shown as range bars [33].

**Exhibit 2-19  Monthly U.S. Prices of Natural Gas and Ammonia**




A cost correlation between the historic prices of U.S. industrial use natural gas ($/MMBtu) and the price of ammonia paid by U.S. farmers (dollars per ton [$/ton]) was developed to estimate the selling price of ammonia in the co-production facilities [33,34]. The correlation was made to the ammonia price paid by U.S. farmers because 84 percent of global ammonia production is utilized by the fertilizer industry. Exhibit 2-21 illustrates ammonia end use [35]. Non-fertilizer applications account for only 16 percent of worldwide ammonia. Direct application of ammonia for fertilization accounts for 3 percent of the total production.

The selling price of ammonia serves as a basis to adequately compare if the synergies created from the co-production facility enhance or serve to detriment the FYCOP of SNG. Exhibit 2-22 shows the data trend for U.S. farmer ammonia prices and industrial natural gas prices. Because other secondary factors such as ammonia supply and demand and cost differences across regions influence the cost of ammonia, the trend is not entirely linear. However, as natural gas prices rise, so does the cost of ammonia.

**Exhibit 2-20 Price Projections for Natural Gas Consumers**


**Exhibit 2-21 Global Ammonia Utilization**

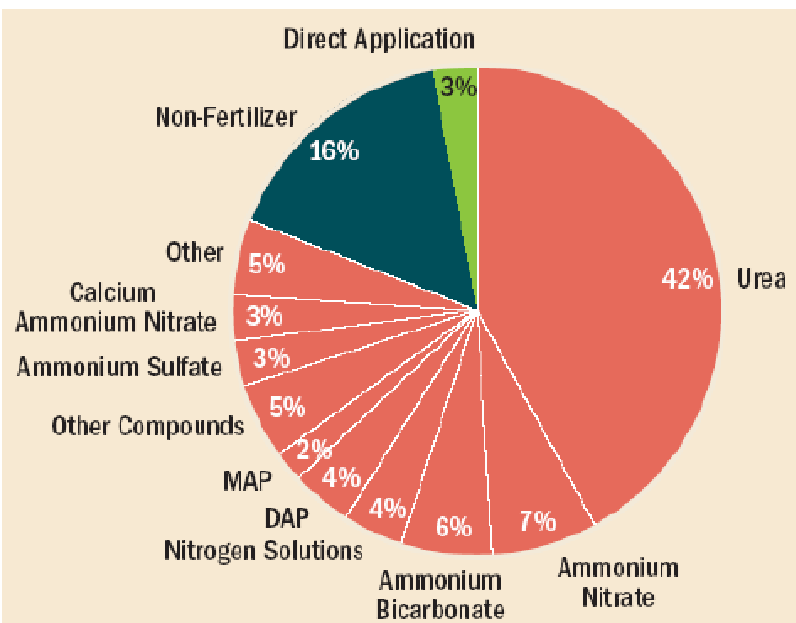
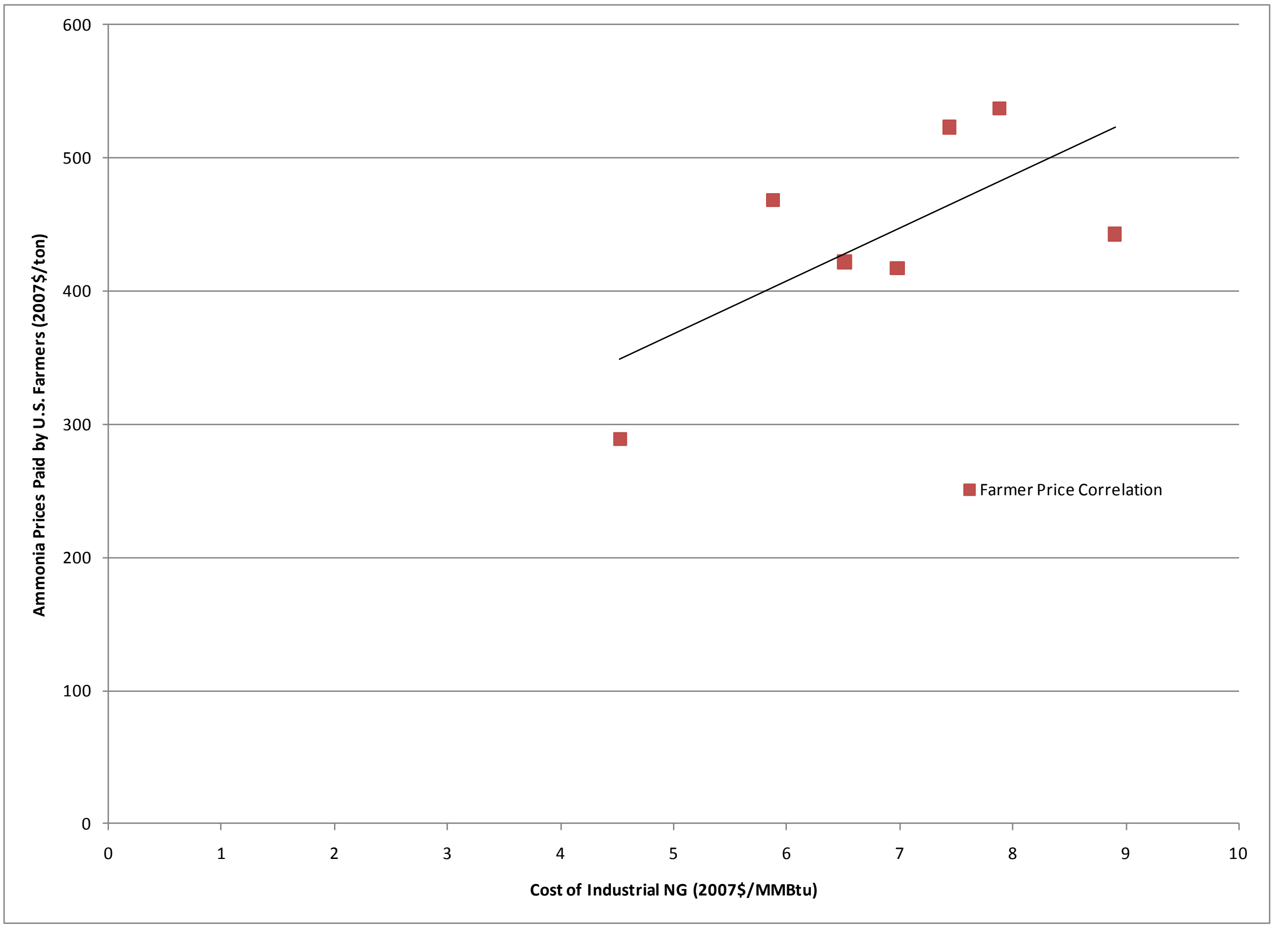


**Exhibit 2-22 U.S. Farmer Ammonia and Industrial Natural Gas Prices**




The following correlation was derived from Exhibit 2-22:

$$\text{Ammonia Price Paid by Farmers}\left(\frac{\$}{\text{ton}}\right) = 39.78 * \text{Cost of Industrial NG}\left(\frac{\$}{\text{MMBtu}}\right) + 169.29$$

This correlation was used to relate the first year price of ammonia to the first year cost of synthetic natural gas.

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# 3. SNG AND SNG AND AMMONIA CO-PRODUCTION FACILITIES

Six SNG only and two SNG and ammonia co-production configurations were evaluated and the major process area descriptions are presented in this section. Each design is based on a market-ready technology that is assumed to be commercially available to support startup in the near term.

All eight cases are based on the Siemens gasifier using either bituminous, PRB, or lignite coal, with and without carbon sequestration. As discussed in Section 2, the thermal input for the six SNG only cases varies by coal type based on vendor input [18]. The output of SNG for the co-production cases is reduced because a portion of the syngas is used to produce 2,000 tonnes/day (2,204 tons/day) of ammonia.

The evaluation scope included developing heat and mass balances and estimating plant performance. Equipment lists were developed for each design to support plant capital and operating cost estimates. The evaluation basis details, including site ambient conditions, fuel composition and environmental targets, were provided in Section 2. Section 3.1 covers general information for the major process areas of the SNG only and co-production plants.

## 3.1 MAJOR PROCESS AREAS

All eight cases have process areas, which are common to each plant configuration such as coal receiving and storage, coal drying, oxygen supply, gas cleanup, power generation, etc. As detailed descriptions of these process areas for each case would be burdensome and repetitious, they are presented in this section for general background information, where information is present that is specific to a particular case or set of cases it is explicitly stated.

### 3.1.1 Coal Receiving and Storage

The function of the Coal Receiving and Storage system is to provide the equipment required for unloading, conveying, preparing, and storing the fuel delivered to the plant. The scope of the system is from the coal conveyor receiving hoppers up to the slide gate valves at the outlet of the coal storage silos.

**Operation Description –** The lignite plants are located at the minemouth and coal is delivered to the plant from the mine by conveyor to the plant storage and reclaim piles. Coal is delivered to the Illinois No. 6 and PRB plants by 100-car unit trains comprised of 91 tonne (100 ton) rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal into two receiving hoppers.

Coal from the reclaim pile is fed directly into a vibratory feeder. The 8 centimenter (cm) x 0 (3" x 0) coal from the feeder is discharged onto a belt conveyor. Two conveyors with an intermediate transfer tower are assumed to convey the coal to the coal stacker, which transfer the coal to either the long-term storage pile or to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

The reclaimer loads the coal into two vibratory feeders located in the reclaim hopper under the pile. The feeders transfer the coal onto a belt conveyor that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3 cm x 0 (1¼" x 0) by the crusher. A conveyor then transfers the coal to a transfer tower. In the transfer tower the coal is routed to the tripper, which loads the coal into one of three silos. Two sampling systems are supplied: the as-received sampling system and the as-fired sampling system. Data from the analyses are used to support the reliable and efficient operation of the plant.

### 3.1.2 Coal Drying

Reduction in coal moisture content improves the efficiency of dry-feed gasifiers, but there is in addition a materials handling requirement. It is necessary to reduce most, if not all, of the surface moisture for coal transport properties to be acceptable. Coal moisture consists of two components, surface moisture and inherent moisture. Low rank coals have higher inherent moisture content and total moisture content than bituminous and other high rank coals and two different drying schemes are used due to this difference. After drying the coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense phase pneumatic conveyor, which uses $CO_2$ from the Selexol unit that is further pressurized and used to convey the coal to the gasifiers.

Bituminous Coal

The coal is simultaneously crushed and dried in the coal mill then delivered to a surge hopper with an approximate 2-hour capacity. The drying medium is provided by combining off-gas from the Claus plant tail gas treatment unit (TGTU) and a slipstream of clean syngas and passing them through an incinerator. The incinerator flue gas, with an oxygen content of 6 volume percent (vol%), is then used to dry the coal in the mill. The removed moisture is vented to atmosphere through the plant stack along with the incinerator flue gas. The bituminous coal in this study is dried to 6 percent from 11.12 percent [18].

Low Rank Coals

No specific coal drying technique is recommended by Siemens either in publications or their response to modeling questions. Therefore, the WTA drying process was chosen primarily based on a previous analysis done for the Shell gasification system that is equally applicable to a Siemens gasification system. In a recent GTC paper, Shell examined the WTA process for drying low rank coals and considered two cases [36]:

1) Case 1: Lignite coal dried from 53 to 12 percent
2) Case 2: Subbituminous coal dried from 30 to 6 percent

For the Siemens gasification cases it is assumed that the subbituminous coal is dried to 6 percent moisture and the lignite to 12 percent moisture [18]. Coal drying options that were evaluated prior to selecting the WTA process included:

- Option 1: Use conventional IGCC coal drying methods, which consist of deriving heat from the combustion of natural gas and/or syngas and using the flue gas either directly or indirectly, by heating a nitrogen stream, for use in drying the coal. Some examples include the following:
    - At the Buggenum facility bituminous coal is ground in a conventional roller mill and simultaneously dried using a heated inert recycle gas stream that carries the evaporated water from the system as it sweeps the PC through an internal, rotating classifier. The inert gas generator is fueled by the in-line combustion of treated syngas. Excess gas is vented on pressure control [37].

  - At the Puertollano facility, coal is ground in mills using nitrogen for drying from 10 percent to less than 2 percent moisture. The drying circuit is heated to about 250°C (482°F) by intermediate-pressure (IP)-steam and additional burning of natural gas. The report notes that a dual drying circuit using only natural gas and the produced syngas could be studied [38].

- Option 2: Dry the coal using a scheme similar to the Great River Energy (GRE) Clean Coal Power Initiative (CCPI) project approach. In the case of gasification where the drying requirements are greater than in PC plants, higher grade heat than is available from the cooling water exiting the condenser will be required. However, there are many sources of low-level waste heat in a gasification-to-fuels plant and integration of this heat into the drying circuit is possible. Because of the tendency for low rank coals to spontaneously combust after the moisture is removed, it may be necessary to fluidize the coal bed with something other than air. Depending on the source of the fluidization gas, the process could become cost prohibitive.
- Option 3: Use the WTA process (German acronym for "fluidized bed dryer with integrated waste heat recovery" [39]) proposed by RWE, which consists of a fluid bed dryer utilizing a heat pump-type cycle with recovered coal moisture used for fluidizing the coal bed and used as the working fluid [40].
- Option 4: Use a process similar to the Integrated Drying and Gasification Combined Cycle (IDGCC) being developed in Australia.

Conventional coal drying methods would be suitable for the low rank coals used in this study, but there is an efficiency penalty involved with burning either syngas or natural gas as a heat source and there are additional $CO_2$ emissions generated by the process. Using steam as the heat source also impacts the production of electricity.

The GRE process has potential because of the many sources of low-grade heat in the gasification process. However, the fluidization gas source would have to be identified assuming ambient air cannot be used because of the combustion tendencies of dried, low rank coals.

The RWE process has the advantage of using steam and coal moisture as the fluidizing medium thus eliminating the spontaneous combustion concerns. However, the vapor compressor imposes a significant auxiliary load.

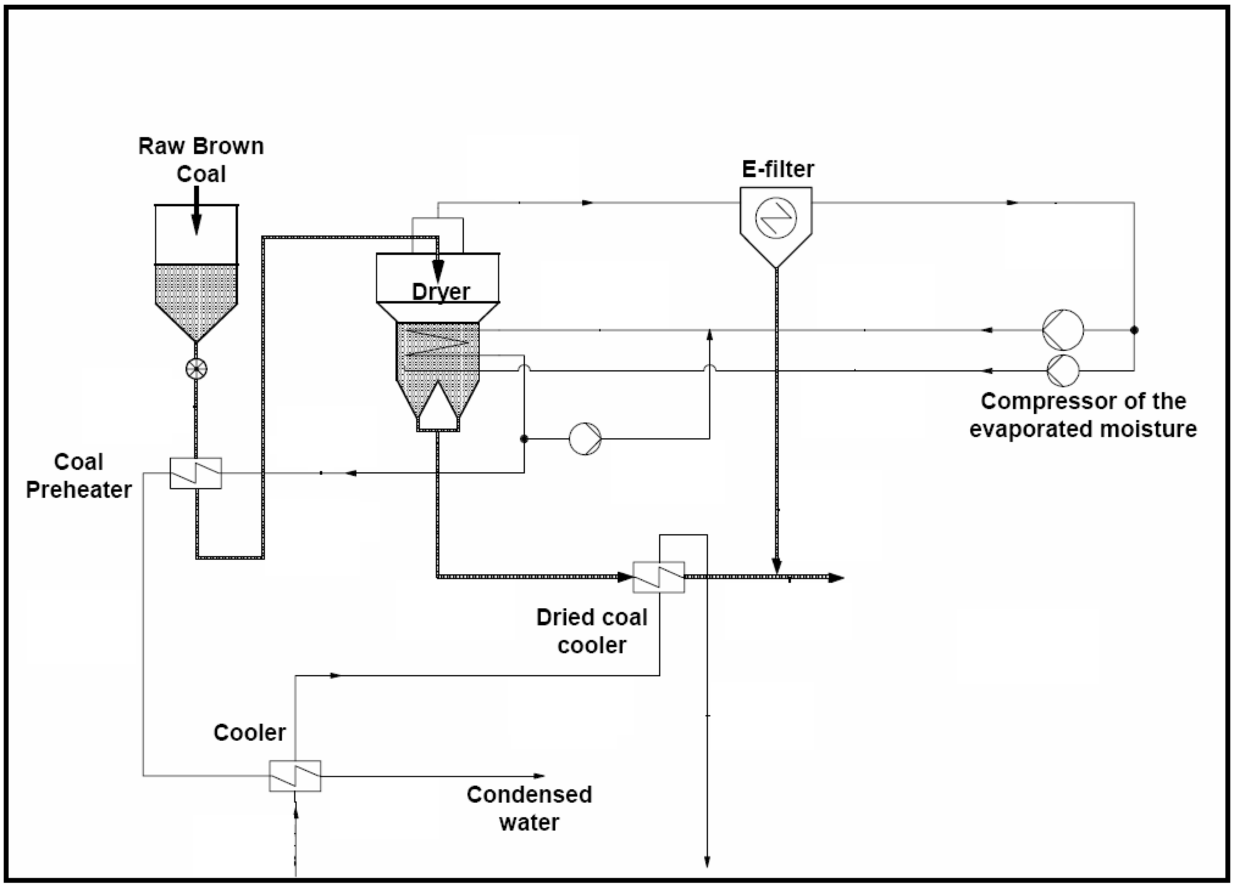
The IDGCC process is being commercialized by HRL Ltd. (Australia) and Harbin Power (China). They recently announced plans for a 400 MW demonstration plant in Australia [41]. The process utilizes hot syngas from an air blown fluidized bed gasifier (FBG) to dry brown coal prior to entering the gasifier. Sulfur is captured in the fluid bed and particulates are removed in a downstream candle filter. The warm, humidified syngas is then burned in a combustion turbine (GT) [42].

The technology is being developed for an air-blown FBG and has features that make it less attractive in the case of an entrained-flow, oxygen-blown gasifier like the Siemens process. The coal is crushed to 10 mm top size and pressurized to 2.5 MPa (363 psi) through a lock hopper system. The FBG can accept 10 mm coal, and at that size and high moisture content the coal will still flow through the lock hoppers. However, in the case of the Siemens gasifier, the coal top size is significantly smaller (Siemens states that conventional coal pulverizers are used, which likely results in an average coal size of ≤ 500 μm for bituminous coal) and the coal must be dried prior to pressurizing to avoid lock hopper flow problems.

The WTA coal drying system was ultimately chosen for this evaluation largely because of its ability to recover the water from the coal in liquid state for use in the process (recycled to the cooling tower) and the fact that syngas is not used to provide heat for drying. In conventional dryers, the water is mixed with the heating gas and discharged to atmosphere as vapor. Recovery of the coal moisture in a liquid state results in a sizable auxiliary load.

The 'closed' WTA process has been demonstrated at pilot scale. Plans for a commercial demonstration of an 'open' version of the process have been delayed. In spite of the uncertainty of the commercial demonstration, the potential benefit of the technology was viewed to be significant enough to use the 'closed version' of the process in this study. A process schematic is shown in Exhibit 3-1 [43].

**Exhibit 3-1 WTA Process Schematic**




### 3.1.3 Air Separation Unit Choice

In order to economically and efficiently support gasification projects, air separation equipment has been modified and improved in response to production requirements and the consistent need to increase single train output. "Elevated pressure" air separation designs have been implemented that result in distillation column operating pressures that are about twice as high as traditional plants. In this study, the main air compressor discharge pressure was set at 1.3 MPa

(190 psia) compared to a traditional ASU plant operating pressure of about 0.7 MPa (105 psia) [44]. For gasification applications the elevated pressure ASU process minimizes power consumption and decreases the size of some of the equipment items.

**Elevated Pressure ASU Experience in Gasification**

The Buggenum, Netherlands unit built for Demkolec was the first elevated-pressure, fully integrated ASU to be constructed. It was designed to produce up to 1,796 tonnes/day (1,980 tons per day [TPD]) of 95 percent purity oxygen for a Shell coal-based gasification unit that fuels a Siemens V94.2 GT. In normal operation at the Buggenum plant the ASU receives all of its air supply from and sends all residual nitrogen to the GT.

The Polk County, Florida ASU for the Tampa Electric IGCC is also an elevated-pressure, 95 percent purity oxygen design that provides 1,832 tonnes/day (2,020 TPD) of oxygen to a GEE coal-based gasification unit, which fuels a General Electric 7FA GT. All of the nitrogen produced in the ASU is used in the GT. The original design did not allow for air extraction from the combustion turbine (CT). After a CT air compressor failure in January, 2005, a modification was made to allow air extraction, which in turn eliminated a bottleneck in ASU capacity and increased overall power output [45].

**Air Separation Plant Process Description**

The air separation plant is designed to produce 99 mole percent (mol%) oxygen ($O_2$) for use in the gasifier [46]. Conventional IGCC plants typically use oxygen purities of 95 mol%. However, the SNG product specification requires higher purity oxygen be used in the gasifier. The plant is designed with four production trains. The air compressor is powered by an electric motor. Nitrogen is also recovered, compressed, and used in the PSA process and the ammonia synthesis loop in the co-production cases only. An additional nitrogen compressor is required in the co-production cases to deliver the nitrogen to the ammonia process at a pressure of 13.6 MPa (1,975 psia). In the SNG only cases, most of the nitrogen is vented to the atmosphere without compression. A process schematic of a typical ASU is shown in Exhibit 3-2.

The air is supplied from a stand-alone compressor. Air to the stand-alone compressor is first filtered in a suction filter upstream of the compressor. This air filter removes particulate, which may tend to cause compressor wheel erosion and foul intercoolers. The filtered air is then compressed in the centrifugal compressor, with intercooling between each stage. The power requirement of the stand-alone compressor is affected by elevation. This effect is accounted for in the plant performance.

Air from the stand-alone compressor is cooled and fed to an adsorbent-based pre-purifier system. The adsorbent removes water, $CO_2$, butane + higher paraffins ($C_4$+), and saturated hydrocarbons from the air. After passing through the adsorption beds, the air is filtered with a dust filter to remove any adsorbent fines that may be present. Downstream of the dust filter a small stream of air is withdrawn to supply the instrument air requirements of the ASU.

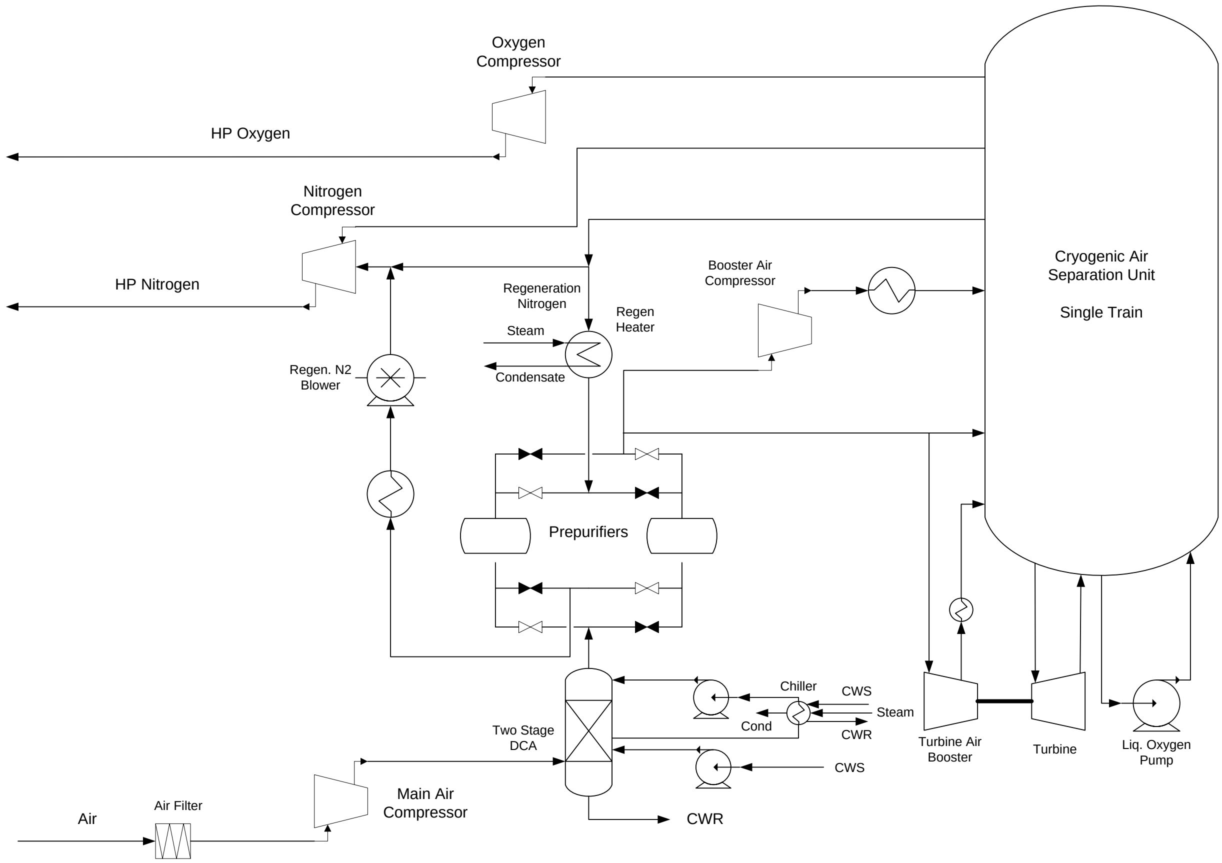
Regeneration of the adsorbent in the pre-purifiers is accomplished by passing a hot nitrogen stream through the off-stream bed(s) in a direction countercurrent to the normal airflow. The nitrogen is heated against extraction steam (1.7 MPa [250 psia]) in a shell and tube heat exchanger. The regeneration nitrogen drives off the adsorbed contaminants. Following regeneration, the heated bed is cooled to near normal operating temperature by passing a cool nitrogen stream through the adsorbent beds. The bed is re-pressurized with air and placed on stream so that the current on-stream bed(s) can be regenerated.

The air from the pre-purifier is then split into three streams. About 70 percent of the air is fed directly to the cold box. About 25 percent of the air is compressed in an air booster compressor. This boosted air is then cooled in an aftercooler against cooling water in the first stage and against chilled water in the second stage before it is fed to the cold box. The chiller utilizes low-pressure (LP) process steam at 0.45 MPa (65 psia). The remaining 5 percent of the air is fed to a turbine-driven, single-stage, centrifugal booster compressor. This stream is cooled in a shell and tube aftercooler against cooling water before it is fed to the cold box.

All three air feeds are cooled in the cold box to cryogenic temperatures against returning product oxygen and nitrogen streams in plate-and-fin heat exchangers. The large air stream is fed directly to the first distillation column to begin the separation process. The second largest air stream is liquefied against boiling liquid oxygen before it is fed to the distillation columns. The third, smallest air stream is fed to the cryogenic expander to produce refrigeration to sustain the cryogenic separation process.

Inside the cold box the air is separated into oxygen and nitrogen products. The oxygen product is withdrawn from the distillation columns as a liquid and is pressurized by a cryogenic pump. The pressurized liquid oxygen is then vaporized against the high-pressure (HP) air feed before being warmed to ambient temperature. The gaseous oxygen exits the cold box and is fed to the centrifugal compressor with intercooling between each stage of compression. The compressed oxygen is then fed to the gasification unit. Nitrogen is produced from the cold box at two pressure levels and is compressed to fulfill the needs of the PSA systems and the ammonia synthesis loop in the co-production cases only. In the SNG only cases, the nitrogen is vented to the atmosphere.

**Exhibit 3-2 Typical ASU Process Schematic**




### 3.1.4 Gasifier

The Siemens gasifier was chosen for this study to allow the use of a single gasifier across all coal types and the vendor was cooperative in providing process information for this and other studies that could be applied to this study. The configuration of processing equipment downstream of the gasifier, such as syngas cooling and degree of quench, may vary with coal type and may not be representative of the latest commercial technology offering.

**Development and Current Status –** The Siemens Fuel Gasification (SFG) process is based on the Noell process, also known under the name GSP, developed by Deutsches Brennstoffinstitut Freiberg in 1975 for the gasification of domestic brown coal and other solid fuels. The Noell Group acquired the technology in 1991 and did further development to gasify waste materials and liquid residues. The gasifier was marketed under the Future Energy GmbH Company and was sold to Siemens in 2006. The first gasifier of this type was a 200 MW thermal input unit built at Schwarze Pumpe, Germany in 1984, firing high sodium lignite. It was converted to gasify natural gas and waste liquids in 1991. Other installations include a 5 MW test plant in Freiberg, Germany where alternative feed testing is ongoing and a 130 MW autothermal oil conversion plant in the Czech Republic, commissioned in 2008. Current gasifier projects include the JinCheng 1000 MW Chinese anthracite coal-to-ammonia plant (under development)[18], the Ningxia Coal Based Polypropylene Project with capacity of 2500 MW of coal to polypropylene production (scheduled for startup in April/May 2010), and the Secure Energy Decatur Coal-to-

SNG project. Scale-up risks should be minimized with start-up of the Ningxia project, although a process contingency is still included in the cost estimates for this project.

The Siemens gasifier, which is a single-stage, entrained-flow, dry-feed gasifier, is modeled as an equilibrium reactor. Many literature references support this modeling strategy [39,47]. Steam injection is based on published data for other single-stage, entrained-flow, dry-feed gasifiers, and the oxygen injection is controlled to maintain estimated heat losses for the gasifier. It is assumed that the gasifier cooling screen that insulates the reactor walls generates LP steam.

**Gasifier Capacity –** The largest current commercial offering from Siemens is a nominal 500 MW thermal input gasifier, with two trains used in the JinCheng project, five trains for the Ningxia project and two trains used in the Secure Energy project. For this study, all cases use six gasifiers. Actual thermal input to each gasifier is coal-type dependent and was based on input from Siemens. The thermal input ranges from 549 MW for bituminous coal to 506 MW for lignite coal with PRB intermediate at 527 MW [18].

**Distinguishing Characteristics –** The key advantage of the Siemens coal gasification technology is its high carbon conversion with a variety of coals and reliability. Similar to other dry feed systems, the Siemens gasifier allows for minimal diluent to be used to inject the feedstock into the gasifier, which has undergone long term testing to characterize a wide range of fuels. The cooling screen is used to control the gasifier shell temperature and creates a solid and liquid slag lining along the gasifier walls instead of a refractory lining, which improves availability, startup and turndown time, and reduces the concerns surrounding coal, and specifically, ash properties. The raw syngas is tar free due to the high gasification temperatures, which improves the downstream gas cleanup. The water quench simplifies the high temperature syngas cooling, improving reliability, resulting in syngas with a high water content, which contributes to the water gas shift (WGS) reaction, benefitting $CO_2$ capture and chemical synthesis applications where hydrogen ($H_2$) and $CO_2$ is preferable to carbon monoxide (CO) production.

The main disadvantage to this, and all dry feed systems, is the need to dry the coal for transport and injection into the gasifier. Water content is generally inversely proportional to coal rank, so, while this gasifier can handle a wide range of fuels, the coal processing and drying can impact performance with different rank coals.

**Important Coal Characteristics –** The Siemens gasifier is generally able to fire any type of coal, petcoke, and mixed feeds.

### 3.1.5 Water Gas Shift Reactors

**Selection of Technology -** In all cases, the gasifier product must be converted to satisfy the requirements for the SNG methanation process and/or the ammonia synthesis loop. The methanation process requires a 3:1 $H_2$ to CO ratio, while the ammonia synthesis loop requires that the maximum amount of hydrogen is produced that is technically and economically feasible.

The concept is to convert the syngas carbon monoxide to hydrogen and $CO_2$ by reacting the CO with water over a bed of catalyst.

*Water Gas Shift:* $CO + H_2O \leftrightarrow CO_2 + H_2$

The CO shift converter can be located either upstream of the acid gas removal (AGR) step (SGS [sour gas shift]) or immediately downstream (sweet gas shift). If the CO converter is located downstream of the AGR, then the metallurgy of the unit is less stringent, but additional equipment must be added to the process. Products from the gasifier are quenched with water and contain a portion of the water vapor necessary to meet the water-to-gas criteria at the reactor inlet. If the CO converter is located downstream of the AGR, then the gasifier product would first have to be cooled and the free water separated and treated. Then additional steam would have to be generated and re-injected into the CO converter feed to meet the required water-to-gas ratio. If the CO converter is located upstream of the AGR step, no additional equipment is required. This is because the CO converter promotes carbonyl sulfide (COS) hydrolysis without a separate catalyst bed. Therefore, for this study the CO converter was located upstream of the AGR unit and is referred to as SGS.

**Process Description –** The two methods to control the amount of shift achieved are the $H_2O$:CO molar ratio and/or the amount of bypass around the shift reactor.

The individual process descriptions for the SNG and co-production portions of the SGS reactors, including steam requirements and bypass ratios, are presented in Sections 4.1 through 5.2.

### 3.1.6 Mercury Removal

A gasification facility has the potential of removing mercury in a more simple and cost-effective manner than conventional PC plants. This is because mercury can be removed from the syngas at elevated pressure and prior to product synthesis so that syngas volumes are much smaller than flue gas volumes in comparable PC cases. A conceptual design for a carbon bed adsorption system was developed for mercury control in the gasification plants being studied. Data on the performance of carbon bed systems were obtained from the Eastman Chemical Company, which uses carbon beds at its syngas facility in Kingsport, Tennessee [15]. The coal mercury content (0.15 ppm dry for Illinois No. 6, 0.081 ppm dry for PRB, and 0.116 ppm dry for lignite) and carbon bed removal efficiency (95 percent) were discussed previously in Section 2.5. Gasification-specific design considerations are discussed below.

**Carbon Bed Location –** The packed carbon bed vessels are located upstream of the AGR process and syngas enters at a temperature near 38°C (100°F). Consideration was given to locating the beds further upstream before the COS hydrolysis/SGS unit at a temperature near 204°C (400°F). However, while the mercury removal efficiency of carbon has been found to be relatively insensitive to pressure variations, temperature adversely affects the removal efficiency [48]. Eastman Chemical also operates their beds ahead of their sulfur recovery unit (SRU) at a temperature of 30°C (86°F) [15].

Consideration was also given to locating the beds downstream of the AGR. However, it was felt that removing the mercury and other contaminants before the AGR unit would enhance the performance of both the AGR and SRU and increase the life of the various solvents.

**Process Parameters –** An empty vessel basis gas residence time of approximately 20 seconds was used based on Eastman Chemical's experience [15]. Allowable gas velocities are limited by considerations of particle entrainment, bed agitation, and pressure drop. One-foot-per-second superficial velocity is in the middle of the range normally encountered [48] and was selected for this application.

The bed density of 30 lb/ft$^3$ was based on the Calgon Carbon Corporation HGR-P sulfur-impregnated pelletized activated carbon [49]. These parameters determined the size of the vessels and the amount of carbon required. Each gasifier train has one mercury removal bed and there are three gasifier trains in each case, resulting in three carbon beds per case.

**Carbon Replacement Time –** Eastman Chemicals replaces its bed every 18 to 24 months [15]. However, bed replacement is not because of mercury loading, but for other reasons including:

- A buildup in pressure drop
- A buildup of water in the bed
- A buildup of other contaminants

For this study a 24 month carbon replacement cycle was assumed. Under these assumptions, the mercury loading in the bed would build up to 0.64 weight percent (wt%). Mercury capacity of sulfur-impregnated carbon can be as high as 20 wt% [50]. The mercury laden carbon is considered to be a hazardous waste, and the disposal cost estimate reflects this categorization.

### 3.1.7 Acid Gas Removal (AGR) Process Selection

Gasification of coal to generate a chemical product produces a syngas that must be treated prior to further utilization. A portion of the treatment consists of AGR and sulfur recovery. To protect the methanation catalyst, the sulfur content of the syngas must be reduced to below 1 ppm to achieve reasonable catalyst life [51,52]. This includes all sulfur species, but in particular the total of COS and hydrogen sulfide ($H_2S$). A two-stage Selexol Process is capable of cleaning the syngas to these low sulfur levels [53].

**COS Hydrolysis**

The use of COS hydrolysis pretreatment in the feed to the AGR process provides a means to reduce the COS concentration. The WGS catalyst also serves to hydrolyze COS, so only the portion of the syngas bypassing the shift reactors must pass through a COS reactor. In the low rank coal cases the syngas sulfur content and WGS bypass fraction are low enough to eliminate the need for a COS reactor in the bypass stream.

This method was first commercially proven at the Buggenum plant, and was also used at both the Tampa Electric and Wabash River IGCC projects. Several catalyst manufacturers including Haldor Topsoe and Porocel offer a catalyst that promotes the COS hydrolysis reaction. The COS hydrolysis reactor designs are based on information from Porocel. The SGS reactors also reduce COS to $H_2S$ as discussed in Section 3.1.5.

The COS hydrolysis reaction is equimolar with a slightly exothermic heat of reaction. The reaction is represented as follows:

$$COS + H_2O \leftrightarrow CO_2 + H_2S$$

Since the reaction is exothermic, higher conversion is achieved at lower temperatures. However, at lower temperatures, the reaction kinetics are slower. Since the exit gas COS concentration is critical to the amount of $H_2S$ that must be removed with the AGR process, a retention time of 50-75 seconds was used to achieve 99.5 percent conversion of the COS. The Porocel activated alumina-based catalyst, designated as Hydrocel 640 catalyst, promotes the COS hydrolysis reaction without promoting reaction of $H_2S$ and CO to form COS and $H_2$.

Although the reaction is exothermic, the heat of reaction is dissipated among the large amount of non-reacting components. Therefore, the reaction is essentially isothermal. The product gas, now containing less than 4 parts per million volume (ppmv) of COS, is cooled prior to entering the mercury removal process and the AGR.

**Acid Gas Removal**

All cases in this study require the removal of hydrogen sulfide and carbon dioxide from the syngas stream. Since the two-stage Selexol process has demonstrated the ability to achieve the low sulfur levels required, and it is also an effective process to remove $CO_2$, it was used for all cases in this study. The Rectisol process would be equally effective in this application, but at greater capital and operating costs. A brief process description follows.

Untreated syngas enters the first of two absorbers where $H_2S$ is preferentially removed using loaded solvent from the $CO_2$ absorber. The gas exiting the $H_2S$ absorber passes through the second absorber where $CO_2$ is removed using first flash regenerated, chilled solvent followed by thermally regenerated solvent added near the top of the column. The treated gas exits the absorber and is sent either to the methanation reactors for SNG production or to the ammonia synthesis loop.

The amount of hydrogen remaining in the syngas stream is dependent on the Selexol process design conditions. In this study, hydrogen recovery is 99.4 percent. The minimal hydrogen slip to the carbon sequestration stream maximizes the overall plant efficiency. The Selexol plant cost estimates are based on a plant designed to recover this high percentage of hydrogen. The balance of the hydrogen is either co-sequestered with the $CO_2$, destroyed in the Claus plant burner, or recycled to the gasifier.

The $CO_2$ loaded solvent exits the $CO_2$ absorber and a portion is sent to the $H_2S$ absorber, a portion is sent to a reabsorber and the remainder is sent to a series of flash drums for regeneration. The $CO_2$ product stream is obtained from the three flash drums, and after flash regeneration the solvent is chilled and returned to the $CO_2$ absorber.

The rich solvent exiting the $H_2S$ absorber is combined with the rich solvent from the reabsorber and the combined stream is heated using the lean solvent from the stripper. The hot, rich solvent enters the $H_2S$ concentrator and partially flashes. The remaining liquid contacts nitrogen from the ASU and a portion of the $CO_2$ along with lesser amounts of $H_2S$ and COS are stripped from the rich solvent. The stripped gases from the $H_2S$ concentrator are sent to the reabsorber where the $H_2S$ and COS that were co-stripped in the concentrator are transferred to a stream of loaded solvent from the $CO_2$ absorber. The clean gas from the reabsorber is combined with the clean gas from the $H_2S$ absorber and sent to SNG and/or $NH_3$ production.

The solvent exiting the $H_2S$ concentrator is sent to the stripper where the absorbed gases are liberated by hot gases flowing up the column from the steam heated reboiler. Water in the overhead vapor from the stripper is condensed and returned as reflux to the stripper or exported as necessary to maintain the proper water content of the lean solvent. The acid gas from the stripper is sent to the Claus plant for further processing. The lean solvent exiting the stripper is first cooled by providing heat to the rich solvent, then further cooled by exchange with the product gas and finally chilled in the lean chiller before returning to the top of the $CO_2$ absorber.

### 3.1.8 Methanation, Synthetic Natural Gas Purification and Compression

Methanation is the hydrogenation of carbon monoxide to methane, typically performed over a nickel-based catalyst. There are two highly exothermic reactions:

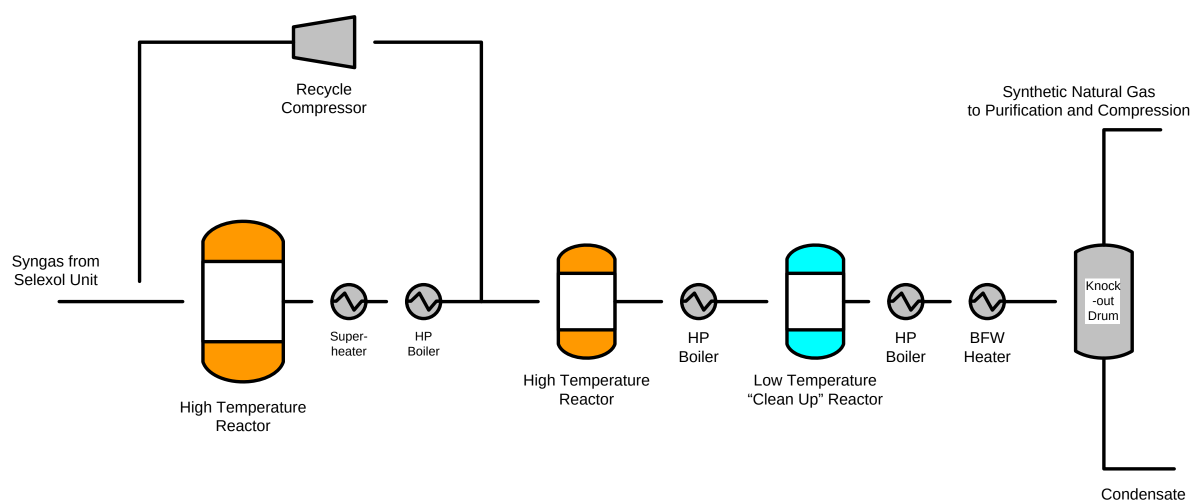
$$CO + 3H_2 \leftrightarrow CH_4 + H_2O\ (\Delta H^0_{298} = -206\ kJ/mol)$$

$$CO_2 + 4H_2 \leftrightarrow CH_4 + 2H_2O\ (\Delta H^0_{298} = -165\ kJ/mol)$$

Due to the highly exothermic nature of the reactions, temperature control and high temperature tolerant catalysts must be implemented.

The methanation system used in this study is based on the high temperature TREMP™ Process by Haldor Topsoe [54]. The TREMP™ process was chosen primarily because it operates at high-temperature and because of the availability of process data in the public domain. A general schematic of the methanation process in this study is shown in Exhibit 3-3. The process uses two different catalysts to account for the adiabatic temperature rise. The high temperature catalyst used in the first and second reactor can operate a range of 300°C (572°F) to 700°C (1290°F) and the low temperature catalyst in the "clean up" reactor can operate at inlet temperatures as low as 200°C (392°F). The process has the ability to make superheated steam at two pressure levels: 12.41 MPa/566°C (1,800 psia/1,050°F) and 4.14 MPa/288°C (600 psia/550°F). The HP steam is used to drive a subcritical turbine, and the medium pressure steam is used for process requirements like shift steam in the WGS reactors.

**Exhibit 3-3 Haldor Topsoe TREMP™ High Temperature Methanation Process**




The syngas from the Selexol Unit, with a $H_2$ to CO ratio of approximately 3:1, is combined with a recycle stream and fed to the first high temperature reactor after being preheated to the minimum catalyst temperature. The reactor temperature for the first high temperature reactor is maintained below the catalyst threshold by varying the amount of recycle mixed with the syngas. After the second high temperature reactor, the syngas is fed to a low temperature "clean up" reactor that reduces the hydrogen content of the SNG product to 0.5 to 1 vol%.

After the methanation reactors the process condensate is removed from the product stream and the SNG product is then passed through a molecular sieve where it removes nearly all of the remaining water and 98.5 vol% of the $CO_2$, which is vented through the plant stack [55]. The

SNG product is further compressed to 6.21 MPa (900 psia) to meet pipeline pressure specifications [56].

### 3.1.9 Pressure Swing Adsorber (Co-Production Cases Only)

The PSA unit is used to purify the syngas from the $NH_3$-production Selexol unit and the purge gas from the ammonia synthesis loop. The operation of a PSA unit for ammonia production based on syngas derived from gasification has been demonstrated at the Coffeyville Resources Fertilizer Complex albeit at a smaller scale.

The PSA unit produces 98.8 vol% hydrogen with a recovery efficiency of 94 vol%. Typical recovery efficiencies are from 88 to 90 vol%. The use of nitrogen as the purging medium in the desorption cycle enables the high recovery efficiency used in this study [57].

The PSA process is accomplished in a series of five steps. The gas is fed to the unit at HP, where impurities are adsorbed and the hydrogen product is produced. When the first adsorber is at capacity, it is taken off-line and the syngas is feed to another adsorber. Hydrogen remains trapped in the adsorbent space. To recover this hydrogen, the adsorber is co-currently depressurized from the product side, and the hydrogen is withdrawn. Nitrogen is used to repressurize and purge other adsorbers. After the hydrogen recovery is complete, the impurities have migrated to the top of the adsorbent bed, filling the bed to capacity. The bed is partly regenerated by depressurization towards the feed end, and the impurities are rejected to the PSA off-gas. The adsorbent is purged with more nitrogen at constant off-gas pressure to further regenerate the bed. The adsorber is repressurized with hydrogen from the co-current depressurization and with a slipstream from the hydrogen product prior to the introduction of the feed gas [14].

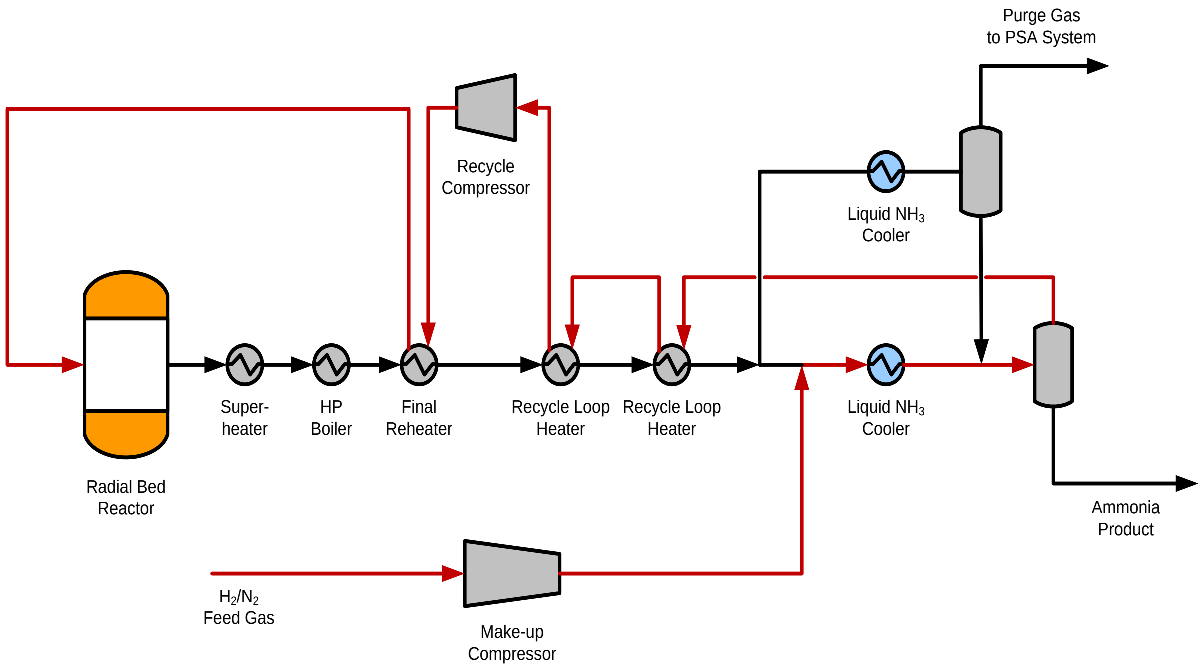
### 3.1.10 Ammonia Synthesis (Co-Production Cases Only)

Ammonia synthesis occurs through the hydrogenation of nitrogen over an iron based catalyst. The hydrogen is derived from the shifted syngas purified by a two-stage Selexol unit and the PSA unit. The nitrogen is produced from the ASU. Co-production cases include the cost and performance adjustments for nitrogen compression.

$$N_2 + 3H_2 \leftrightarrow 2NH_3 \ (\Delta H^0_{298} = -46 \text{ kJ/mol})$$

The synthesis of ammonia typically takes place at pressure in the range of 10.00 to 24.99 MPa (1,450 to 3,625 psia) at a temperature of 343°C to 550°C (650 to 1,022°F) [58]. For this study the gas enters the reactor at 14.00 MPa (2,030 psia) and 343°C (650°F). The configuration modeled in this study is based on the Haldor Topsoe S-300 Ammonia Synthesis Loop and is shown in Exhibit 3-4. The path of the feed gas through the system is shown in red.

**Exhibit 3-4 Ammonia Synthesis Loop Configuration**




The makeup gas from the PSA system is introduced into the synthesis loop before the separation of the ammonia product. Each reactor pass converts only 20 to 30 percent due to the unfavorable equilibrium conditions. Due to the exothermic nature of the reaction, superheated steam at 12.51 MPa (1,815 psia) and HP steam at 4.14 MPa (600 psia) can be generated in the synthesis loop. The ammonia is separated/condensed by a refrigeration system.

Purge gas is removed from the system to reduce the concentration of catalyst poisons and inerts such as argon and methane. Permanent poisons are sulfur containing compounds. Temporary catalyst poisons can cause deactivation during the time of exposure. Typical poisons are oxygen containing compounds, such as carbon oxides, water, and oxygen. Catalyst poisoning is a function of the total concentration of oxygen equivalents (O eqv) [59].

$$\text{ppm O eqv} = 2 \text{ x (ppm } O_2) + 2 \text{ x (ppm } CO_2) + \text{ppm CO} + \text{ppm } H_2O$$

Less than 2 ppm O eqv is considered ideal and at higher concentrations the catalyst will begin to show a loss of activity. Cases 3 and 4 achieve below 1 total ppm O equivalent.

### 3.1.11 Zinc Oxide Guard Bed

The sulfur removal requirements for the methanation and ammonia processes are at the extreme limit of Selexol's capabilities. To protect against system upsets or fluctuations in coal sulfur content, a zinc oxide guard bed was included in each case to remove trace sulfur contaminants.

A zinc oxide bed can absorb over 99.99 percent of the inlet $H_2S$ and can absorb a maximum of 39.3 pounds of sulfur per pound of pure zinc oxide (ZnO). In this study it was assumed that the bed would reach 70 percent of saturation at breakthrough.

### 3.1.12 Sulfur Recovery/Tail Gas Cleanup Process Selection

Currently, most of the world's sulfur is produced from the acid gases coming from gas treating. The Claus process remains the mainstay for sulfur recovery. Conventional three-stage Claus plants, with indirect reheat and feeds with a high $H_2S$ content, can approach 98 percent sulfur recovery efficiency. However, since environmental regulations have become more stringent, sulfur recovery plants are required to recover sulfur with over 99.8 percent efficiency. To meet these stricter regulations, the Claus process underwent various modifications and add-ons.

The add-on modification to the Claus plant selected for this study can be considered a separate option from the Claus process. In this context, it is often called a TGTU process.

**The Claus Process**

The Claus process converts $H_2S$ to elemental sulfur via the following reactions:

$$H_2S + 3/2\ O_2 \leftrightarrow H_2O + SO_2$$

$$2H_2S + SO_2 \leftrightarrow 2H_2O + 3S$$

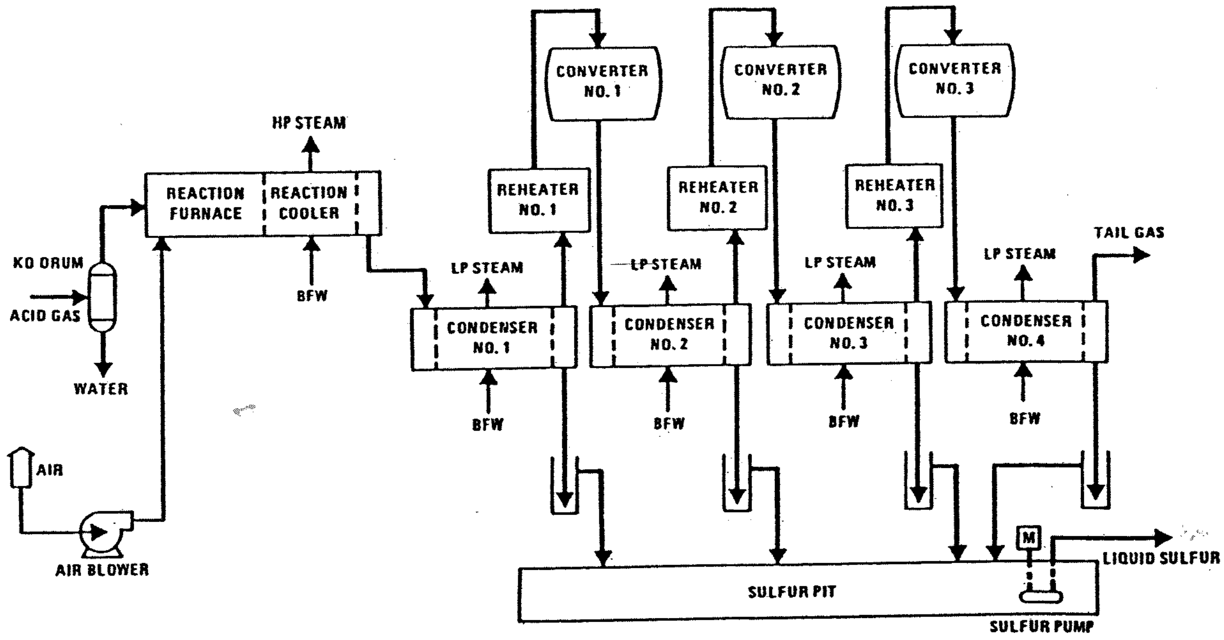
The second reaction, the Claus reaction, is equilibrium limited. The overall reaction is:

$$3H_2S + 3/2\ O_2 \leftrightarrow 3H_2O + 3S$$

The sulfur in the vapor phase exists as $S_2$, $S_6$, and $S_8$ molecular species, with the $S_2$ predominant at higher temperatures, and $S_8$ predominant at lower temperatures.

A simplified process flow diagram of a typical three-stage Claus plant is shown in Exhibit 3-5 [60]. One-third of the $H_2S$ is burned in the furnace with oxygen from the air to give sufficient $SO_2$ to react with the remaining $H_2S$. Since these reactions are highly exothermic, a waste heat boiler that recovers this heat to generate HP steam usually follows the furnace. Sulfur is condensed in a condenser that follows the HP steam recovery section. LP steam is raised in the condenser. The tail gas from the first condenser then goes to several catalytic conversion stages, usually two to three, where the remaining sulfur is recovered via the Claus reaction. Each catalytic stage consists of gas preheat, a catalytic reactor, and a sulfur condenser. The liquid sulfur goes to the sulfur pit, while the tail gas proceeds to the incinerator or for further processing in a TGTU.

**Exhibit 3-5 Typical Three-Stage Claus Sulfur Plant**




## Claus Plant Sulfur Recovery Efficiency

The Claus reaction is equilibrium limited, and sulfur conversion is sensitive to the reaction temperature. The highest sulfur conversion in the thermal zone is limited to about 75 percent. Typical furnace temperatures are in the range from 1,093 to 1,427°C (2,000 to 2,600°F), and as the temperature decreases, conversion increases dramatically.

Claus plant sulfur recovery efficiency depends on many factors:

- $H_2S$ concentration of the feed gas
- Number of catalytic stages
- Gas reheat method

In order to keep Claus plant recovery efficiencies approaching 94 to 96 percent for feed gases that contain about 20 to 50 percent $H_2S$, a split-flow design is often used. In this version of the Claus plant, part of the feed gas is bypassed around the furnace to the first catalytic stage, while the rest of the gas is oxidized in the furnace to mostly $SO_2$. This results in a more stable temperature in the furnace.

## Oxygen-Blown Claus

Large diluent streams in the feed to the Claus plant, such as $N_2$ from combustion air, or a high $CO_2$ content in the feed gas, lead to higher cost Claus processes and any add-on or tail gas units. One way to reduce diluent flows through the Claus plant and to obtain stable temperatures in the furnace for dilute $H_2S$ streams is the oxygen-blown Claus process.

The oxygen-blown Claus process was originally developed to increase capacity at existing conventional Claus plants and to increase flame temperatures of low $H_2S$ content gases. The process has also been used to provide the capacity and operating flexibility for sulfur plants where the feed gas is variable in flow and composition such as often found in refineries. The application of the process has now been extended to grass roots installations, even for rich $H_2S$ feed streams, to provide operating flexibility at lower costs than would be the case for conventional Claus units. At least four of the recently built gasification plants in Europe use oxygen enriched Claus units.

Oxygen enrichment results in higher temperatures in the front-end furnace, potentially reaching temperatures as high as 1,593 to 1,649°C (2,900 to 3,000°F) as the enrichment moves beyond 40 to 70 vol% $O_2$ in the oxidant feed stream. Although oxygen enrichment has many benefits, its primary benefit for lean $H_2S$ feeds is a stable furnace temperature. Sulfur recovery is not significantly enhanced by oxygen enrichment. Because the gasification process already requires an ASU, the oxygen-blown Claus plant was chosen for all cases.

**Tail Gas Treating**

In many refinery and other conventional Claus applications, tail gas treating involves the removal of the remaining sulfur compounds from gases exiting the SRU. Tail gas from a typical Claus process, whether a conventional Claus or one of the extended versions of the process, usually contains small, but varying quantities of COS, carbon disulfide ($CS_2$), $H_2S$, $SO_2$, and elemental sulfur vapors. In addition, there may be $H_2$, CO, and $CO_2$ in the tail gas. In order to remove the rest of the sulfur compounds from the tail gas, all of the sulfur-bearing species must first be converted to $H_2S$. Then, the resulting $H_2S$ is absorbed into a solvent and the clean gas vented or recycled for further processing. The clean gas resulting from the hydrolysis step can undergo further cleanup in a dedicated absorption unit or be integrated with an upstream AGR unit. The latter option is particularly suitable with physical absorption solvents. The approach of treating the tail gas in a dedicated amine absorption unit and recycling the resulting acid gas to the Claus plant is the one used by the SCOT process. With tail gas treatment, Claus plants can achieve overall removal efficiencies in excess of 99.9 percent.

In the case of gasification-based applications, the tail gas from the Claus plant can be catalytically hydrogenated and then recycled back into the system with the choice of location being technology dependent, or it can be treated with a SCOT-type process. The low rank coal cases in this report all use a catalytic hydrogenation step with tail gas recycle to the AGR. The bituminous coal cases utilize the SCOT process with acid gas recycled back to the AGR and the clean gas used for coal drying. The Shell Puertollano plant treats the tail gas in a similar manner, but the recycle endpoint is not specified [38].

**Flare Stack**

A self-supporting, refractory-lined, carbon steel (CS) flare stack is typically provided to combust and dispose of unreacted gas during startup, shutdown, and upset conditions. However, in all eight cases a flare stack was provided for syngas dumping during startup, shutdown, etc. This flare stack eliminates the need for a separate Claus plant flare.

### 3.1.13 Slag Handling

The slag handling system conveys, stores, and disposes of slag removed from the gasification process. Spent material drains from the gasifier bed into a water bath in the bottom of the gasifier vessel. A slag crusher receives slag from the water bath and grinds the material into peasized fragments. A slag/water slurry that is between 5 and 10 percent solids leaves the gasifier pressure boundary through the use of lockhoppers to a series of dewatering bins.

The slag is dewatered, the water is clarified and recycled and the dried slag is transferred to a storage area for disposal.

In this study the slag bins were sized for a nominal holdup capacity of 72 hours of full-load operation. At periodic intervals, a convoy of slag-hauling trucks will transit the unloading station underneath the hopper and remove a quantity of slag for disposal. Approximately 30-50 truckloads per day are required to remove the total quantity of slag produced by the plant, depending on coal type, operating at nominal rated power. While the slag is suitable for use as a component of road paving mixtures, it was assumed in this study that the slag would be landfilled at a specified cost.

### 3.1.14 Steam Generation Island

#### Steam Turbine Generator and Auxiliaries

The ST consists of an HP section, an IP section, and one double-flow LP section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing.

Superheated main steam is combined in a header, and then passes through the stop valves and control valves and enters the turbine at 12.4 MPa/566°C (1800 psig/1050°F). The steam initially enters the turbine near the middle of the HP span, flows through the turbine. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser. In the Midwestern location cases the single condenser is water-cooled. In the Montana and North Dakota location cases the steam exiting the turbine splits evenly between a water–cooled condenser and an air-cooled condenser.

The generator is a hydrogen-cooled synchronous type, generating power at 24 kilovolt (kV). A static, transformer type exciter is provided. The generator is cooled with a hydrogen gas recirculation system using fans mounted on the generator rotor shaft. The heat absorbed by the gas is removed as it passes over finned tube gas coolers mounted in the stator frame. Gas is prevented from escaping at the rotor shafts by a closed-loop (CL) oil seal system. The oil seal system consists of storage tank, pumps, filters, and pressure controls, all skid-mounted.

The STG is controlled by a triple-redundant, microprocessor-based electro-hydraulic control system. The system provides digital control of the unit in accordance with programmed control algorithms, color cathode ray tube (CRT) operator interfacing, and data link interfaces to the balance-of-plant distributed control system (DCS), and incorporates on-line repair capability.

#### Feedwater System

The function of the feedwater (FW) system is to pump the various FW streams from the deaerator storage tank to the respective vaporizers. Two 50 percent capacity boiler feed pumps (BFPs) are provided for each of three pressure levels, HP, IP, and LP. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

The FW pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing temperature. FW pump suction pressure and temperature are also monitored. In addition, the suction of each BFP is equipped with a startup strainer.

**Main and Reheat Steam Systems**

The function of the main steam system is to convey main steam generated in the raw gas coolers and the methanation and ammonia synthesis superheater (where applicable) outlets to the HP turbine stop valves.

Main steam at approximately 12.4 MPa/566°C (1800 psig/1050°F) exits through a motor-operated stop/check valve and a motor-operated gate valve, and is routed to a single line feeding the HP turbine. Cold reheat steam at 3.5 MPa/376°C (501 psia/708°F) exits the HP turbine, flows through a motor-operated isolation gate valve, to the syngas cooler. Hot reheat steam at approximately 3.5 MPa/538°C (501 psia/1000°F) exits the syngas cooler and is routed to the IP turbines.

Safety valves are installed to comply with appropriate codes and to ensure the safety of personnel and equipment.

**Circulating Water System**

The circulating water system (CWS) is a closed-cycle cooling water system that supplies cooling water to the condenser to condense the main turbine exhaust steam (all of the exhaust steam in Midwestern locations and half of the steam in the Montana and North Dakota locations). The system also supplies cooling water to the AGR plant as required, and to the auxiliary cooling system. The auxiliary cooling system is a CL process that utilizes a higher quality water to remove heat from compressor intercoolers, oil coolers and other ancillary equipment and transfers that heat to the main circulating cooling water system in plate and frame heat exchangers. The heat transferred to the circulating water in the condenser and other applications is removed by a mechanical draft cooling tower.

The system consists of two 50 percent capacity vertical CWPs, a mechanical draft evaporative cooling tower, and CS cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counter flow mechanical draft cooling tower.

<u>Midwestern Location Cases</u>

The condenser is a single-pass, horizontal type with divided water boxes. All of the steam is condensed using cooling water for the bituminous coal cases. There are two separate circulating water circuits in each box. One-half of the condenser can be removed from service for cleaning or for plugging tubes. This can be done during normal operation at reduced load.

The condenser is equipped with an air extraction system to evacuate the condenser steam space for removal of non-condensable gases during ST operation and to rapidly reduce the condenser pressure from atmospheric pressure before unit startup and admission of steam to the condenser.

<u>Montana and North Dakota Location Cases</u>

Exhaust steam from the ST is split 50/50 to a surface condenser cooled with cooling water and to an air-cooled condenser used ambient air and forced convection. A decision to use a parallel wet/dry cooling system was based primarily on the plans for the Xcel Energy Comanche 3 PC plant currently in start up, and the desire to reduce the plant water requirement in western locations where water constraints are frequently encountered. With the relatively low ambient temperature, the performance impact from the parallel cooling, as compared to wet cooling, is minor.

The major impact of parallel cooling is a significant reduction in water requirement when compared to a wet cooling system. This impact is included in the water balance presented later in this report.

With this cooling system and the specific ambient temperature, a condenser pressure of 0.005 MPa (0.698 psia) (condensing temperature of 32°C [90°F]) is used in the Western location cases as compared to 0.007 MPa (0.983 psia) (condensing temperature of 38°C [101°F]) used in the Midwestern location cases.

**Raw Water, Fire Protection, and Cycle Makeup Water Systems**

The raw water system supplies cooling tower makeup, cycle makeup, service water, and potable water requirements. The water source is 50 percent from a POTW and 50 percent from groundwater. Booster pumps within the plant boundary provide the necessary pressure.

The fire protection system provides water under pressure to the fire hydrants, hose stations, and fixed water suppression system within the buildings and structures. The system consists of pumps, underground and aboveground supply piping, distribution piping, hydrants, hose stations, spray systems, and deluge spray systems. One motor-operated booster pump is supplied on the intake structure of the cooling tower with a diesel engine backup pump installed on the water inlet line.

The cycle makeup water system provides high quality demineralized water for makeup to the steam cycle, for steam injection ahead of the WGS reactors and gasifier, and for injection steam to the auxiliary boiler for control of NOx emissions, if required.

The cycle makeup system consists of two 100 percent trains, each with a full-capacity activated carbon filter, primary cation exchanger, primary anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. The equipment is skid-mounted and includes a control panel and associated piping, valves, and instrumentation.

### 3.1.15 Accessory Electric Plant

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

### 3.1.16 Instrumentation and Control

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

percent of the time it is required (99.5 percent availability). The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are manually implemented, with operator selection of modular automation routines available. The exception to this, and an important facet of the control system for gasification, is the critical controller system, which is a part of the license package from the gasifier supplier and is a dedicated and distinct hardware segment of the DCS.

This critical controller system is used to control the gasification process. The partial oxidation of the fuel feed and oxygen feed streams to form a syngas product is a stoichiometric, temperature- and pressure-dependent reaction. The critical controller utilizes a redundant microprocessor executing calculations and dynamic controls at 100- to 200-millisecond intervals. The enhanced execution speeds as well as evolved predictive controls allow the critical controller to mitigate process upsets and maintain the reactor operation within a stable set of operating parameters.

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# 4. BITUMINOUS COAL CASES

This section contains an evaluation of plant designs for Cases 1 through 4, which are based on the Siemens gasifier using Illinois No. 6 coal as the fuel. All cases capture $CO_2$ from the syngas stream because of the necessary process conditions for chemical production, but in one case $CO_2$ is released to the atmosphere and in the other it is compressed and sequestered. Cases 1 and 2 are very similar in terms of process, equipment, scope and arrangement, except that Case 2 utilizes a $CO_2$ compression train to enable transportation and sequestration. The same is true for Cases 3 and 4.

Section 4.1 covers the two SNG cases using Illinois No. 6 coal. Section 4.2 covers the two SNG and ammonia co-production cases using Illinois No. 6 coal. Each section is structured analogously as follows:

- Process and System Description provides an overview of the technology operation as applied.
- Key Assumptions is a summary of relevant study and modeling assumptions.
- Sparing Philosophy is provided.
- Performance Results provide the main modeling results, including the performance summary, environmental performance, carbon balance, sulfur balance, water balance, mass and energy balance diagrams, and mass and energy balance tables.
- Equipment List provides an itemized list of major equipment with account codes that correspond to the cost accounts in the Cost Estimates section.
- Cost Estimates provides a summary of capital and operating costs.

If the information is identical to that presented for the non-sequestration cases, a reference is made to the earlier section rather than repeating the information.

## 4.1 SYNTHETIC NATURAL GAS PRODUCTION USING ILLINOIS NO. 6 COAL

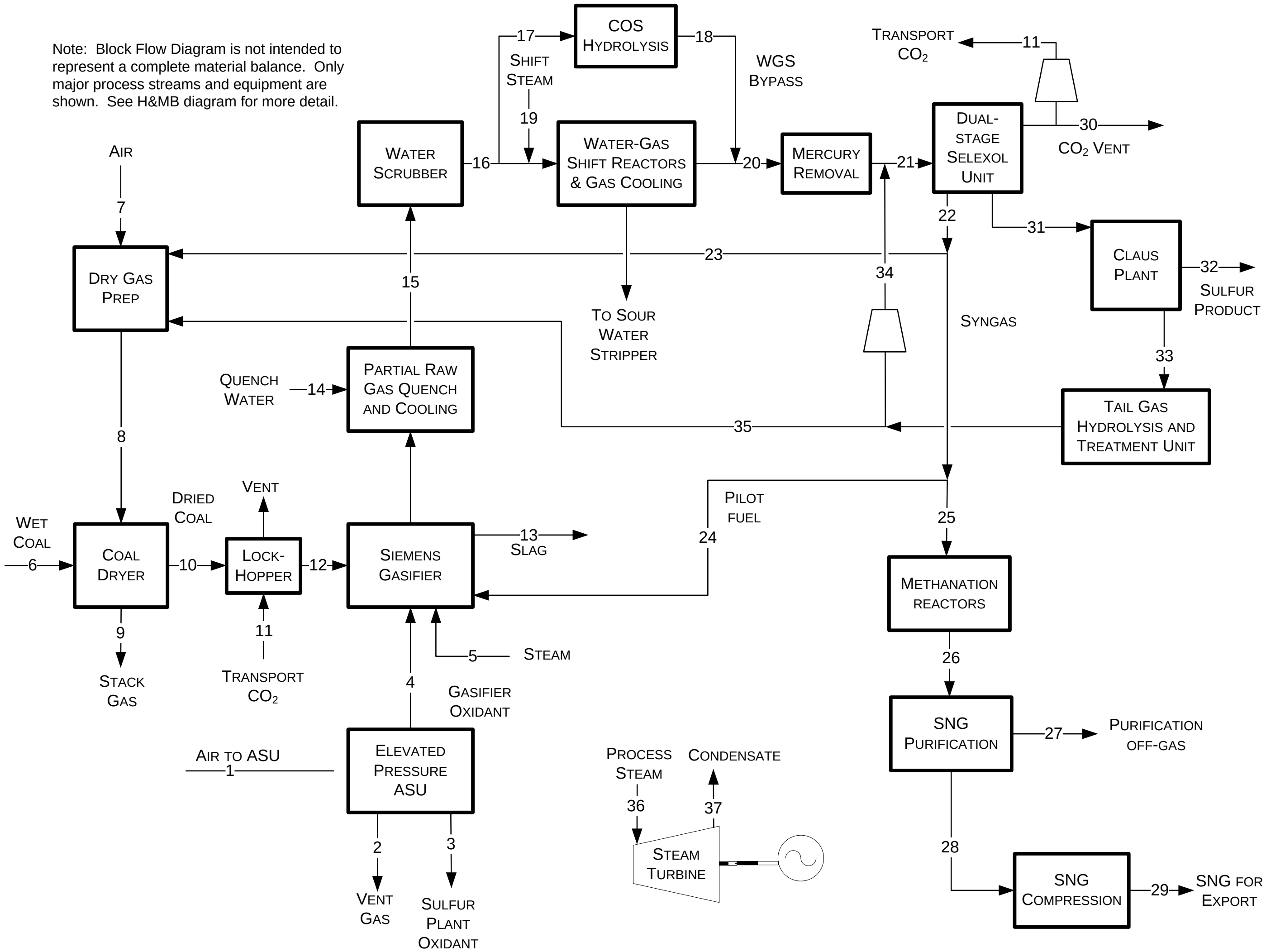
### 4.1.1 Process Description

In this section the overall Siemens gasification process for the production of SNG for Cases 1 and 2 is described. The only major process difference is that Case 2 employs $CO_2$ compression and sequestration. The system description follows the block flow diagrams (BFDs) in Exhibit 4-1 and Exhibit 4-3 and stream numbers reference the same exhibit. The tables in Exhibit 4-2 and Exhibit 4-4 provide process data for the numbered streams in the BFD.

#### Coal Preparation and Feed Systems

Coal receiving and handling is covered in Section 3.1.1. The receiving and handling subsystem ends at the coal silo. The Siemens process uses a dry feed system which is sensitive to the coal moisture content. Coal moisture consists of two parts, surface moisture and inherent moisture. For coal to flow smoothly through the lock hoppers, the surface moisture must be removed. The bituminous coal used in this study contains 11.12 percent moisture on an as-received basis (stream 6). It was assumed that the bituminous coal must be dried to 6 percent moisture to allow for smooth flow through the dry feed system (stream 10).

**Exhibit 4-1 Case 1 Block Flow Diagram, Bituminous Coal to SNG without Carbon Sequestration**



# Exhibit 4-2 Case 1 Stream Table, Bituminous Coal to SNG without Carbon Sequestration

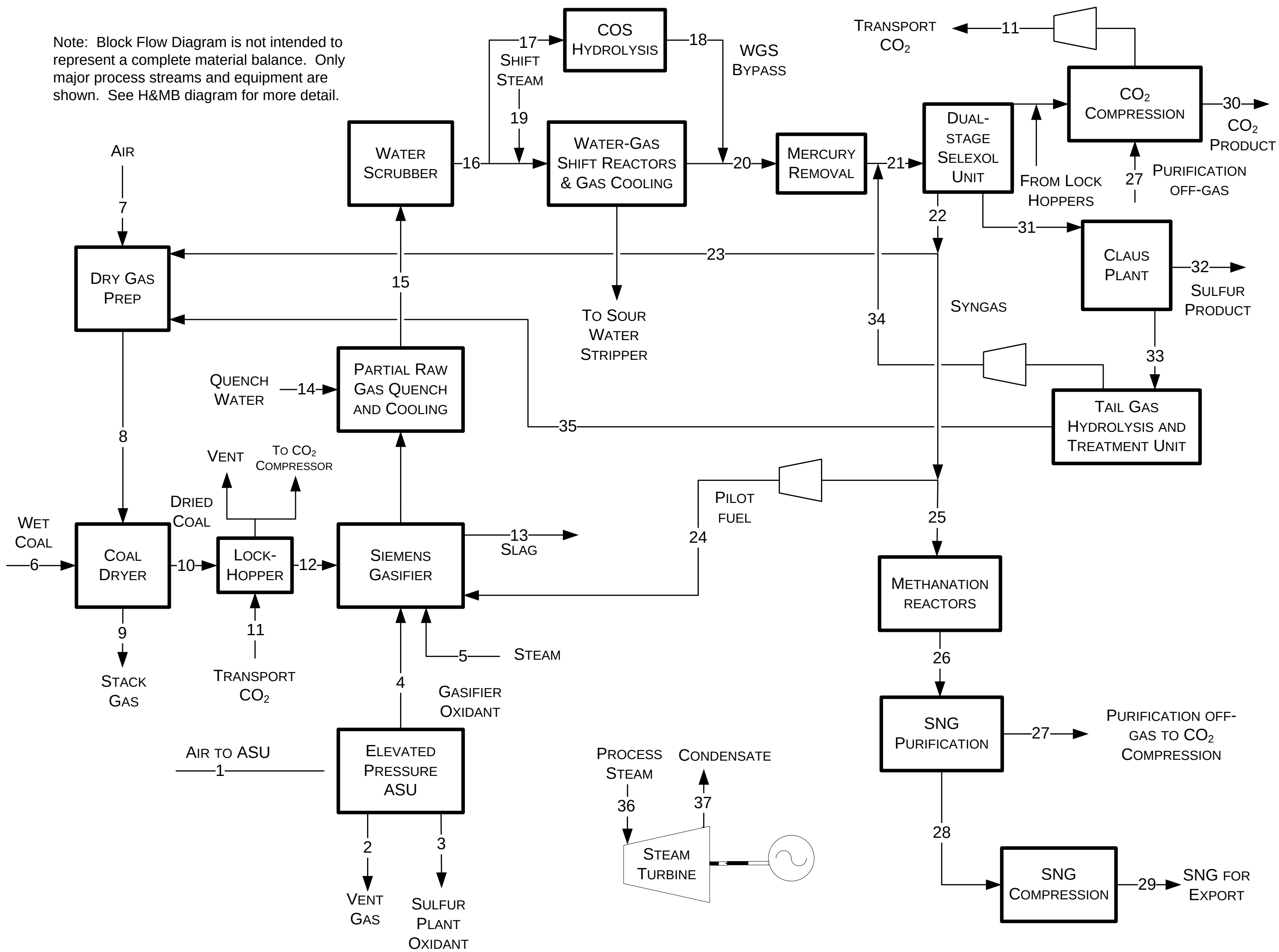
| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | | | |
| Ar | 0.0092 | 0.0091 | 0.0101 | 0.0101 | 0.0000 | 0.0000 | 0.0092 | 0.0084 | 0.0064 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0019 | 0.0020 | 0.0020 | 0.0020 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0119 | 0.0015 | 0.0000 | 0.0000 | 0.4228 | 0.4387 | 0.4387 | 0.4387 | 0.0000 |
| $CO_2$ | 0.0003 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0003 | 0.1134 | 0.0869 | 0.0000 | 0.9736 | 0.1231 | 0.0000 | 0.0000 | 0.0380 | 0.0394 | 0.0394 | 0.0400 | 0.0000 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0005 | 0.0006 | 0.0006 | 0.0000 | 0.0000 |
| $H_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0127 | 0.6936 | 0.0000 | 0.0000 | 0.2001 | 0.2076 | 0.2076 | 0.2076 | 0.0000 |
| $H_2O$ | 0.0099 | 0.0039 | 0.0000 | 0.0000 | 1.0000 | 0.0000 | 0.0099 | 0.1726 | 0.3661 | 0.0000 | 0.0015 | 0.0979 | 0.0000 | 1.0000 | 0.3261 | 0.3009 | 0.3009 | 0.3004 | 1.0000 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0058 | 0.0060 | 0.0060 | 0.0066 | 0.0000 |
| $N_2$ | 0.7732 | 0.9790 | 0.0004 | 0.0004 | 0.0000 | 0.0000 | 0.7732 | 0.6276 | 0.4809 | 0.0000 | 0.0001 | 0.0171 | 0.0000 | 0.0000 | 0.0046 | 0.0047 | 0.0047 | 0.0047 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.2074 | 0.0076 | 0.9895 | 0.9895 | 0.0000 | 0.0000 | 0.2074 | 0.0780 | 0.0597 | 0.0000 | 0.0000 | 0.0668 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 51,091 | 40,346 | 151 | 10,249 | 3,445 | 0 | 3,513 | 4,336 | 5,659 | 0 | 3,569 | 14,109 | 0 | 14,021 | 53,959 | 52,011 | 7,721 | 7,721 | 14,746 |
| V-L Flowrate (kg/hr) | 1,474,339 | 1,134,515 | 4,838 | 328,760 | 62,065 | 0 | 101,361 | 123,623 | 147,458 | 0 | 154,354 | 158,594 | 0 | 252,597 | 1,091,692 | 1,056,587 | 156,847 | 156,847 | 265,661 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 437,604 | 0 | 0 | 0 | 413,769 | 0 | 332,351 | 43,828 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 15 | 24 | 32 | 32 | 343 | 15 | 15 | 1,278 | 143 | 71 | 83 | 18 | 1,500 | 216 | 260 | 215 | 215 | 215 | 288 |
| Pressure (MPa, abs) | 0.10 | 0.11 | 0.86 | 0.86 | 5.10 | 0.10 | 0.10 | 0.10 | 0.10 | 0.10 | 5.62 | 5.62 | 4.24 | 8.274 | 3.89 | 3.76 | 3.76 | 3.69 | 4.14 |
| Enthalpy (kJ/kg) | 30.93 | 30.58 | 26.94 | 26.94 | 3,177.81 | --- | 30.93 | 1,884.17 | 835.89 | --- | 31.45 | 4,685.53 | --- | 1,017.873 | 1,148.27 | 1,011.50 | 1,011.50 | 1,011.03 | 3,070.04 |
| Density ($kg/m^3$) | 1.2 | 1.3 | 11.0 | 11.0 | 20.1 | --- | 1.2 | 0.2 | 0.8 | --- | 98.9 | 28.4 | --- | 782.0 | 17.9 | 19.0 | 19.0 | 18.7 | 18.2 |
| V-L Molecular Weight | 28.857 | 28.120 | 32.078 | 32.078 | 18.015 | --- | 28.857 | 28.514 | 26.059 | --- | 43.248 | 11.241 | --- | 18.015 | 20.232 | 20.314 | 20.314 | 20.314 | 18.015 |
| | | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 112,637 | 88,948 | 332 | 22,595 | 7,595 | 0 | 7,744 | 9,558 | 12,475 | 0 | 7,868 | 31,105 | 0 | 30,912 | 118,960 | 114,666 | 17,022 | 17,022 | 32,510 |
| V-L Flowrate (lb/hr) | 3,250,361 | 2,501,177 | 10,665 | 724,793 | 136,831 | 0 | 223,464 | 272,542 | 325,090 | 0 | 340,292 | 349,640 | 0 | 556,880 | 2,406,768 | 2,329,375 | 345,789 | 345,789 | 585,682 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 964,752 | 0 | 0 | 0 | 912,204 | 0 | 732,709 | 96,625 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 59 | 74 | 90 | 90 | 650 | 59 | 59 | 2,332 | 290 | 160 | 181 | 65 | 2,732 | 420 | 500 | 418 | 418 | 419 | 550 |
| Pressure (psia) | 14.7 | 16.4 | 125.0 | 125.0 | 740.0 | 14.7 | 14.7 | 14.7 | 14.7 | 14.4 | 815.0 | 815.0 | 614.7 | 1,200.0 | 564.7 | 544.7 | 544.7 | 534.7 | 600.0 |
| Enthalpy (Btu/lb) | 13.3 | 13.1 | 11.6 | 11.6 | 1,366.2 | --- | 13.3 | 810.0 | 359.4 | --- | 13.5 | 2,014.4 | --- | 437.6 | 493.7 | 434.9 | 434.9 | 434.7 | 1,319.9 |
| Density ($lb/ft^3$) | 0.076 | 0.081 | 0.685 | 0.685 | 1.257 | --- | 0.076 | 0.014 | 0.048 | --- | 6.174 | 1.774 | --- | 48.817 | 1.120 | 1.189 | 1.189 | 1.166 | 1.135 |

A - Reference conditions are 32.02 F & 0.089 PSIA

# Exhibit 4-2 Case 1 Stream Table, Bituminous Coal to SNG without Carbon Sequestration (continued)

| | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | 34 | 35 | 36 | 37 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | | |
| Ar | 0.0021 | 0.0020 | 0.0029 | 0.0029 | 0.0029 | 0.0029 | 0.0103 | 0.0000 | 0.0118 | 0.0118 | 0.0000 | 0.0005 | 0.0000 | 0.0021 | 0.0000 | 0.0077 | 0.0000 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.8334 | 0.0000 | 0.9495 | 0.9495 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.1648 | 0.1644 | 0.2366 | 0.2366 | 0.2366 | 0.2366 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0028 | 0.0484 | 0.0000 | 0.1158 | 0.0000 | 0.0327 | 0.0000 | 0.0000 |
| $CO_2$ | 0.3242 | 0.3256 | 0.0361 | 0.0361 | 0.0361 | 0.0361 | 0.1217 | 0.9811 | 0.0021 | 0.0021 | 0.9896 | 0.4310 | 0.0000 | 0.3134 | 0.9589 | 0.0713 | 0.0000 | 0.0000 |
| COS | 0.0001 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0002 | 0.0017 | 0.0000 | 0.0005 | 0.0001 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.4956 | 0.4945 | 0.7172 | 0.7172 | 0.7172 | 0.7172 | 0.0076 | 0.0000 | 0.0086 | 0.0086 | 0.0028 | 0.0927 | 0.0000 | 0.0623 | 0.0006 | 0.1325 | 0.0000 | 0.0000 |
| $H_2O$ | 0.0018 | 0.0018 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0023 | 0.0189 | 0.0000 | 0.0000 | 0.0045 | 0.0330 | 0.0000 | 0.5017 | 0.0059 | 0.0258 | 1.0000 | 1.0000 |
| $H_2S$ | 0.0066 | 0.0067 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.3921 | 0.0000 | 0.0021 | 0.0345 | 0.0000 | 0.0000 | 0.0000 |
| $N_2$ | 0.0048 | 0.0048 | 0.0070 | 0.0070 | 0.0070 | 0.0070 | 0.0246 | 0.0000 | 0.0280 | 0.0280 | 0.0000 | 0.0006 | 0.0000 | 0.0008 | 0.0000 | 0.5758 | 0.0000 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1542 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0012 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 50,897 | 51,011 | 34,955 | 638 | 115 | 34,202 | 9,763 | 1,193 | 8,570 | 8,570 | 11,619 | 868 | 0 | 991 | 114 | 4,726 | 41,891 | 64,580 |
| V-L Flowrate (kg/hr) | 1,036,367 | 1,041,318 | 348,816 | 6,363 | 1,152 | 341,301 | 194,158 | 51,931 | 142,227 | 142,227 | 508,119 | 30,029 | 0 | 26,262 | 4,952 | 123,623 | 754,680 | 1,163,420 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10,888 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 35 | 34 | 31 | 31 | 31 | 31 | 35 | 35 | 35 | 35 | 18 | 48 | 178 | 232 | 21 | 23 | 566 | 39 |
| Pressure (MPa, abs) | 3.31 | 3.24 | 3.24 | 3.24 | 3.24 | 3.24 | 2.731 | 0.138 | 2.627 | 6.202 | 0.9 | 0.163 | 0.119 | 0.085 | 3.272 | 0.103 | 12.512 | 0.827 |
| Enthalpy (kJ/kg) | 45.27 | 44.52 | 86.41 | 86.41 | 86.41 | 86.41 | 42.166 | 48.781 | 47.826 | 13.195 | 8.0 | 91.877 | --- | 1,202.561 | -13.721 | 67.724 | 3,513.235 | 164.567 |
| Density ($kg/m^3$) | 26.7 | 26.3 | 12.6 | 12.6 | 12.6 | 12.6 | 22.4 | 2.4 | 17.8 | 44.6 | 17.8 | 2.1 | 5,280.4 | 0.5 | 73.9 | 1.1 | 34.9 | 992.9 |
| V-L Molecular Weight | 20.362 | 20.414 | 9.979 | 9.979 | 9.979 | 9.979 | 19.887 | 43.519 | 16.597 | 16.597 | 44 | 34.607 | --- | 26.492 | 43.489 | 26.158 | 18.015 | 18.015 |
| | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 112,209 | 112,460 | 77,063 | 1,406 | 255 | 75,402 | 21,523 | 2,631 | 18,893 | 18,893 | 25,616 | 1,913 | 0 | 2,185 | 251 | 10,419 | 92,354 | 142,374 |
| V-L Flowrate (lb/hr) | 2,284,798 | 2,295,714 | 769,007 | 14,028 | 2,540 | 752,440 | 428,046 | 114,489 | 313,557 | 313,557 | 1,120,211 | 66,204 | 0 | 57,897 | 10,916 | 272,542 | 1,663,785 | 2,564,903 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 24,003 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 95 | 94 | 87 | 87 | 87 | 87 | 95 | 95 | 95 | 95 | 65 | 119 | 352 | 450 | 70 | 74 | 1,050 | 103 |
| Pressure (psia) | 479.7 | 469.7 | 469.6 | 469.6 | 469.6 | 469.6 | 396.1 | 20.1 | 381.0 | 899.5 | 135.0 | 23.7 | 17.3 | 12.3 | 474.5 | 14.9 | 1,814.7 | 120.0 |
| Enthalpy (Btu/lb) | 19.5 | 19.1 | 37.1 | 37.1 | 37.1 | 37.1 | 18.1 | 21.0 | 20.6 | 5.7 | 3.4 | 39.5 | --- | 517.0 | -5.9 | 29.1 | 1,510.4 | 70.8 |
| Density ($lb/ft^3$) | 1.667 | 1.639 | 0.789 | 0.789 | 0.789 | 0.789 | 1.401 | 0.148 | 1.114 | 2.785 | 1.11 | 0.133 | 329.643 | 0.033 | 4.616 | 0.068 | 2.177 | 61.987 |

**Exhibit 4-3 Case 2 Block Flow Diagram, Bituminous Coal to SNG with Carbon Sequestration**



# Exhibit 4-4 Case 2 Stream Table, Bituminous Coal to SNG with Carbon Sequestration

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | | | |
| Ar | 0.0092 | 0.0091 | 0.0101 | 0.0101 | 0.0000 | 0.0000 | 0.0092 | 0.0084 | 0.0064 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0019 | 0.0020 | 0.0020 | 0.0020 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0046 | 0.0006 | 0.0000 | 0.0000 | 0.4229 | 0.4388 | 0.4388 | 0.4388 | 0.0000 |
| $CO_2$ | 0.0003 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0003 | 0.1133 | 0.0869 | 0.0000 | 0.9902 | 0.1240 | 0.0000 | 0.0000 | 0.0381 | 0.0395 | 0.0395 | 0.0401 | 0.0000 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0005 | 0.0006 | 0.0006 | 0.0000 | 0.0000 |
| $H_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0048 | 0.6935 | 0.0000 | 0.0000 | 0.1999 | 0.2074 | 0.2074 | 0.2074 | 0.0000 |
| $H_2O$ | 0.0099 | 0.0039 | 0.0000 | 0.0000 | 1.0000 | 0.0000 | 0.0099 | 0.1718 | 0.3648 | 0.0000 | 0.0001 | 0.0978 | 0.0000 | 1.0000 | 0.3262 | 0.3010 | 0.3010 | 0.3004 | 1.0000 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0058 | 0.0060 | 0.0060 | 0.0066 | 0.0000 |
| $N_2$ | 0.7732 | 0.9790 | 0.0004 | 0.0004 | 0.0000 | 0.0000 | 0.7732 | 0.6281 | 0.4817 | 0.0000 | 0.0000 | 0.0172 | 0.0000 | 0.0000 | 0.0046 | 0.0047 | 0.0047 | 0.0047 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.2074 | 0.0076 | 0.9895 | 0.9895 | 0.0000 | 0.0000 | 0.2074 | 0.0785 | 0.0602 | 0.0000 | 0.0000 | 0.0669 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 51,089 | 40,345 | 151 | 10,248 | 3,445 | 0 | 3,531 | 4,355 | 5,678 | 0 | 3,530 | 14,089 | 0 | 14,016 | 53,935 | 51,984 | 7,779 | 7,779 | 14,732 |
| V-L Flowrate (kg/hr) | 1,474,284 | 1,134,472 | 4,848 | 328,738 | 62,065 | 0 | 101,888 | 124,203 | 148,038 | 0 | 154,354 | 158,594 | 0 | 252,508 | 1,091,580 | 1,056,432 | 158,081 | 158,081 | 265,396 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 437,604 | 0 | 0 | 0 | 413,769 | 0 | 332,351 | 43,828 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 15 | 24 | 32 | 32 | 343 | 15 | 15 | 1,274 | 143 | 71 | 82 | 18 | 1,500 | 216 | 260 | 215 | 215 | 215 | 288 |
| Pressure (MPa, abs) | 0.10 | 0.11 | 0.86 | 0.86 | 5.10 | 0.10 | 0.10 | 0.10 | 0.10 | 0.10 | 5.62 | 5.62 | 4.24 | 8.274 | 3.89 | 3.76 | 3.76 | 3.69 | 4.14 |
| Enthalpy (kJ/kg) | 30.93 | 30.58 | 26.94 | 26.94 | 3,177.81 | --- | 30.93 | 1,875.99 | 833.13 | --- | 27.52 | 4,692.64 | --- | 1,017.873 | 1,148.09 | 1,011.26 | 1,011.26 | 1,010.79 | 3,070.04 |
| Density ($kg/m^3$) | 1.2 | 1.3 | 11.0 | 11.0 | 20.1 | --- | 1.2 | 0.2 | 0.8 | --- | 101.2 | 28.5 | --- | 782.0 | 17.9 | 19.1 | 19.1 | 18.7 | 18.2 |
| V-L Molecular Weight | 28.857 | 28.120 | 32.078 | 32.078 | 18.015 | **---** | 28.857 | **28.521** | **26.073** | **---** | **43.732** | 11.256 | --- | 18.015 | 20.239 | 20.322 | 20.322 | 20.322 | 18.015 |
| | | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 112,633 | 88,944 | 333 | 22,593 | 7,595 | 0 | 7,784 | 9,601 | 12,518 | 0 | 7,781 | 31,061 | 0 | 30,901 | 118,905 | 114,606 | 17,149 | 17,149 | 32,478 |
| V-L Flowrate (lb/hr) | 3,250,240 | 2,501,083 | 10,688 | 724,742 | 136,831 | 0 | 224,625 | 273,820 | 326,368 | 0 | 340,292 | 349,640 | 0 | 556,684 | 2,406,522 | 2,329,035 | 348,508 | 348,508 | 585,098 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 964,752 | 0 | 0 | 0 | 912,204 | 0 | 732,709 | 96,625 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 59 | 74 | 90 | 90 | 650 | 59 | 59 | 2,325 | 290 | 160 | 179 | 65 | 2,732 | 420 | 500 | 418 | 418 | 419 | 550 |
| Pressure (psia) | 14.7 | 16.4 | 125.0 | 125.0 | 740.0 | 14.7 | 14.7 | 14.7 | 14.7 | 14.4 | 815.0 | 815.0 | 614.7 | 1,200.0 | 564.7 | 544.7 | 544.7 | 534.7 | 600.0 |
| Enthalpy (Btu/lb) | 13.3 | 13.1 | 11.6 | 11.6 | 1,366.2 | --- | 13.3 | 806.5 | 358.2 | --- | 11.8 | 2,017.5 | --- | 437.6 | 493.6 | 434.8 | 434.8 | 434.6 | 1,319.9 |
| Density ($lb/ft^3$) | 0.076 | 0.081 | 0.685 | 0.685 | 1.257 | --- | 0.076 | 0.014 | 0.048 | --- | 6.320 | 1.778 | --- | 48.817 | 1.120 | 1.190 | 1.190 | 1.167 | 1.135 |

A - Reference conditions are 32.02 F & 0.089 PSIA

## Exhibit 4-4 Case 2 Stream Table, Bituminous Coal to SNG with Carbon Sequestration (continued)

| | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | 34 | 35 | 36 | 37 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | | |
| Ar | 0.0020 | 0.0020 | 0.0029 | 0.0029 | 0.0029 | 0.0029 | 0.0103 | 0.0000 | 0.0118 | 0.0118 | 0.0001 | 0.0005 | 0.0000 | 0.0021 | 0.0000 | 0.0036 | 0.0000 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.8318 | 0.0000 | 0.9495 | 0.9495 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.1652 | 0.1649 | 0.2372 | 0.2372 | 0.2372 | 0.2372 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0046 | 0.0485 | 0.0000 | 0.1153 | 0.0000 | 0.0065 | 0.0000 | 0.0000 |
| $CO_2$ | 0.3241 | 0.3255 | 0.0361 | 0.0361 | 0.0361 | 0.0361 | 0.1236 | 0.9814 | 0.0021 | 0.0021 | 0.9902 | 0.4310 | 0.0000 | 0.3149 | 0.9630 | 0.5446 | 0.0000 | 0.0000 |
| COS | 0.0001 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0017 | 0.0000 | 0.0005 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.4953 | 0.4942 | 0.7166 | 0.7166 | 0.7166 | 0.7166 | 0.0075 | 0.0000 | 0.0085 | 0.0085 | 0.0048 | 0.0927 | 0.0000 | 0.0611 | 0.0006 | 0.2924 | 0.0000 | 0.0000 |
| $H_2O$ | 0.0018 | 0.0018 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0023 | 0.0186 | 0.0000 | 0.0000 | 0.0001 | 0.0330 | 0.0000 | 0.5024 | 0.0059 | 0.1514 | 1.0000 | 1.0000 |
| $H_2S$ | 0.0066 | 0.0067 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.3920 | 0.0000 | 0.0021 | 0.0304 | 0.0000 | 0.0000 | 0.0000 |
| $N_2$ | 0.0048 | 0.0048 | 0.0070 | 0.0070 | 0.0070 | 0.0070 | 0.0245 | 0.0000 | 0.0280 | 0.0280 | 0.0000 | 0.0006 | 0.0000 | 0.0008 | 0.0000 | 0.0014 | 0.0000 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0008 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 50,851 | 50,965 | 34,927 | 638 | 115 | 34,174 | 9,775 | 1,212 | 8,562 | 8,562 | 14,470 | 867 | 0 | 992 | 114 | 576 | 42,211 | 65,028 |
| V-L Flowrate (kg/hr) | 1,035,837 | 1,040,786 | 349,074 | 6,379 | 1,152 | 341,543 | 194,891 | 52,765 | 142,126 | 142,126 | 632,824 | 29,996 | 0 | 26,303 | 4,949 | 15,936 | 760,442 | 1,171,495 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10,891 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 35 | 34 | 31 | 31 | 80 | 31 | 35 | 35 | 35 | 35 | 32 | 48 | 178 | 232 | 21 | 51 | 566 | 39 |
| Pressure (MPa, abs) | 3.31 | 3.24 | 3.24 | 3.24 | 5.10 | 3.24 | 2.73 | 0.138 | 2.627 | 6.202 | 15.272 | 0.2 | 0.119 | 0.085 | 3.272 | 0.103 | 12.512 | 0.827 |
| Enthalpy (kJ/kg) | 45.24 | 44.50 | 86.27 | 86.27 | 234.71 | 86.27 | 42.02 | 48.480 | 47.817 | 13.183 | -216.068 | 91.9 | --- | 1,202.323 | -13.694 | 240.482 | 3,513 | 164.549 |
| Density ($kg/m^3$) | 26.7 | 26.3 | 12.7 | 12.7 | 17.0 | 12.7 | 22.5 | 2.4 | 17.9 | 44.6 | 799.5 | 2.1 | 5,279.9 | 0.5 | 74.0 | 1.1 | 34.9 | 992.9 |
| V-L Molecular Weight | 20.370 | 20.422 | 9.994 | 9.994 | 9.994 | 9.994 | 19.938 | 43.526 | 16.599 | 16.599 | 43.732 | 35 | --- | 26.526 | 43.529 | 27.652 | 18.015 | 18.015 |
| | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 112,108 | 112,359 | 77,001 | 1,407 | 254 | 75,340 | 21,549 | 2,673 | 18,877 | 18,877 | 31,902 | 1,911 | 0 | 2,186 | 251 | 1,271 | 93,059 | 143,362 |
| V-L Flowrate (lb/hr) | 2,283,630 | 2,294,540 | 769,577 | 14,063 | 2,540 | 752,974 | 429,662 | 116,328 | 313,334 | 313,334 | 1,395,138 | 66,130 | 0 | 57,988 | 10,910 | 35,132 | 1,676,488 | 2,582,704 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 24,010 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 95 | 94 | 87 | 87 | 177 | 87 | 95 | 95 | 95 | 95 | 90 | 119 | 352 | 450 | 70 | 123 | 1,050 | 103 |
| Pressure (psia) | 479.7 | 469.7 | 469.6 | 469.6 | 740.0 | 469.6 | 396.1 | 20.1 | 381.0 | 899.5 | 2,215.0 | 23.7 | 17.3 | 12.3 | 474.5 | 14.9 | 1,814.7 | 120.0 |
| Enthalpy (Btu/lb) | 19.5 | 19.1 | 37.1 | 37.1 | 100.9 | 37.1 | 18.1 | 20.8 | 20.6 | 5.7 | -92.9 | 39.5 | --- | 516.9 | -5.9 | 103.4 | 1,510 | 70.7 |
| Density ($lb/ft^3$) | 1.667 | 1.640 | 0.790 | 0.790 | 1.061 | 0.790 | 1.405 | 0.148 | 1.114 | 2.786 | 49.913 | 0 | 329.612 | 0.033 | 4.620 | 0.069 | 2.177 | 61.987 |

The coal is simultaneously crushed and dried in the coal mill then delivered to a surge hopper with an approximate 2-hour capacity. The drying medium is provided by combining the off-gas from the Claus plant TGTU and a slipstream of clean syngas (stream 23) and passing them through the incinerator. The incinerator flue gas, with an oxygen content of 8 vol%, is then used to dry the coal in the mill (stream 8).

The coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense phase pneumatic conveyor, which uses $CO_2$ from the Selexol unit (stream 11) to convey the coal to the gasifiers. $CO_2$ is chosen instead of $N_2$ as the transport medium due to the process needs of the methanation section, which requires a low level of inerts for a more high value product and also eliminates the need for further SNG product purification. Approximately 50 percent of the coal transport $CO_2$ is vented from the lock hopper. In the non-sequestration case, the $CO_2$ is simply vented to the atmosphere and in the sequestration case approximately 97 percent of the $CO_2$ is recovered and compressed for sequestration.

**Gasifier**

There are six Siemens dry feed, pressurized, down flow, entrained, slagging gasifiers, operating at 4.24 MPa (615 psia). Coal reacts with oxygen (stream 4) and steam (stream 5) at a temperature of 1,499°C (2,730°F) to produce principally hydrogen and carbon monoxide. The clean syngas recycle stream is used as pilot fuel for the gasifier burners. The Siemens gasifiers use a ‘cooling screen’ (heat exchanger tubes) in the gasification zone of the gasifier. Circulating water passes through these heat exchangers and LP steam is generated in an external steam drum. Molten slag builds up a layer on the cooling screen and protects the cooling screen against the high reaction temperatures. Liquid slag from the reaction phase only comes in contact with the solidified slag layer and hence no corrosion of the reactor wall can take place. The gasifier reaction temperature is maintained by setting the oxygen/feedstock (primary) and steam to feedstock (secondary) ratios within precise limits.

**Raw Gas Cooling/Particulate Removal**

The Siemens system in this study uses a partial water quench to 899°C (1,650°F) with a syngas cooler for steam generation. High-temperature heat recovery in each gasifier train is accomplished in three steps, including the gasifier cooling screen. The product gas from the gasifier is cooled using quench water (stream 14) to lower the temperature below the ash melting point. Syngas then goes through a raw gas cooler, which lowers the gas temperature to 260°C (500°F), and produces HP steam for use in the steam cycle.

The ash that is not carried out with the gas forms slag and runs down the interior walls, exiting the gasifier in liquid form. The slag is solidified in a quench tank for disposal (stream 13). Lockhoppers are used to reduce the pressure of the solids from 4.24 MPa (615 psia) to ambient.

**Syngas Scrubber/Sour Water Stripper**

Downstream of the quench section is a raw gas scrubber system including two venturi scrubbers and a partial condenser. In two successive venturi scrubbers the raw gas is intensely mixed with water to remove the fine ash and soot particles. Cyclone-type separators remove water and particles from the raw gas flow. The scrubbers run on circuit water and additional water (e.g., gas condensate from the downstream gas cooling after CO shift, if applicable). To reduce fine particles and the salty fogs from the raw gas, a partial condensation of the raw gas is done downstream of the scrubber unit. In the partial condenser the raw gas is cooled down to condense a portion of the water content and separate fine particles. This generated gas condensate is separated from the raw gas and collected in a separator drum with a demister.

The sour water stripper removes $NH_3$, $SO_2$, and other impurities from the waste stream of the scrubber. The sour gas stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from synthesis gas coolers (SGCs). Sour water from the drum flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the SRU. Remaining water is sent to wastewater treatment.

**Sour Gas Shift and COS Hydrolysis**

The SGS and COS hydrolysis units are employed in combination to achieve a 3:1 $H_2$ to CO ratio required for methanation and to reduce the amount of COS entering the AGR system. $H_2S$ and COS are at significant concentrations, requiring removal to achieve the low design level necessary for methanation synthesis. $H_2S$ is removed in an AGR process; however, because COS is not readily removed, it is catalytically converted to $H_2S$ in the SGS reactor and COS hydrolysis unit.

Following the water scrubber, the gas is reheated to 214°C (418°F) and fed to the SGS and COS hydrolysis reactors. This raw gas reheat temperature is controlled to achieve a nominal temperature of 232°C (450°F) entering the SGS reactor when combined with steam injection. Approximately 85 percent of the raw gas stream exiting the scrubber is sent to the SGS reactor. Steam (stream 19) is injected into the SGS reactor at 288°C (550°F) and 4.14 MPa (600 psia) to drive the shift reaction and maintain a 0.3 $H_2O$ to dry gas ratio exiting the reactor. The remaining raw gas from the scrubber is sent to the hydrolysis reactor in the bypass (stream 17) and COS is hydrolyzed with steam over a catalyst bed to $H_2S$, which is more easily removed by the AGR solvent.

**Mercury Removal and Acid Gas Removal**

Before the raw fuel gas can be treated in the AGR process, it must be cooled to about 35°C (95°F). During this cooling through a series of heat exchangers, most of the water vapor condenses. This water, which contains some $NH_3$, is sent to the sour water stripper. The cooled syngas (stream 20) then passes through a carbon bed to remove 95 percent of the Hg (Section 3.1.6).

The AGR process is a two stage Selexol process where $H_2S$ is removed in the first stage and $CO_2$ in the second stage of absorption. The process results in three product streams, the clean syngas (stream 22), a $CO_2$-rich stream (streams 11 and 30), and an acid gas feed to the Claus plant (stream 31). In both cases, the acid gas contains about 40 percent $H_2S$ and 45 percent $CO_2$ with the balance primarily $H_2$ and CO. The $CO_2$-rich stream is discussed further in the $CO_2$ compression section.

**$CO_2$ Compression**

$CO_2$ from the AGR process is generated at two pressure levels, a LP stream at 0.12 MPa (17 psia) and an HP stream at 1.03 MPa (150 psia). For the non-sequestration case (Case 1), a portion of the captured $CO_2$ from the HP stream is further compressed to 5.62 MPa (815 psia) to provide the transport medium in the coal feed/lock hopper system and the remaining $CO_2$ is vented to the atmosphere. For the carbon sequestration case (Case 2), the LP stream is compressed to 1.0 MPa (150 psia) and then combined with the HP stream. The combined stream is further compressed to a SC condition at 15.3 MPa (2,215 psia) using a multiple-stage, intercooled compressor. At 5.17 MPa (750 psia), a portion of the $CO_2$ compression stream is withdrawn, further compressed to 5.86 MPa (850 psia) using a boost compressor (stream 11), and sent to the lock hopper system. During compression, the $CO_2$ stream is dehydrated to a dewpoint of -40ºC (-40°F) with triethylene glycol prior to extraction of the transport $CO_2$. The raw $CO_2$ stream from the Selexol process contains over 99 percent $CO_2$. The dehydrated $CO_2$ (stream 30 in Exhibit 4-3) is transported to the plant fence line and is sequestration ready. $CO_2$ TS&M costs were estimated using the methodology described in Section 2.8.

**Methanation System and SNG Compression**

Clean syngas from the Selexol unit (stream 25) is sent to the methanation reactors after syngas is withdrawn to fulfill the requirements of the gasifier (stream 24) and the coal drying system (stream 23). The syngas is reacted over three methanation reactors with interstage cooling that generates superheated HP and medium pressure steam as described in Section 3.1.8. Process condensate is removed from the SNG product (stream 26), which now contains approximately 83 vol% $CH_4$. The SNG product is further purified by a molecular sieve, which removes 98.5 vol% of the $CO_2$ and all of the water. The purification off-gas (stream 27) contains over 98 vol% $CO_2$ and is either vented in the non-sequestration case or sent to $CO_2$ compression for sequestration in the carbon sequestration case. The purified SNG product stream (stream 28) is compressed to 6.21 MPa (900 psia) to meet natural gas pipeline specifications (stream 29).

**Claus Unit**

The SRU is a Claus bypass type SRU utilizing oxygen instead of air. The Claus plant produces molten sulfur (stream 32) by reacting approximately one third of the $H_2S$ in the feed to $SO_2$, then reacting the $H_2S$ and $SO_2$ to sulfur and water. The combination of Claus technology and SCOT tail gas technology results in an overall sulfur recovery exceeding 99 percent.

Utilizing oxygen instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The sulfur plant produces elemental sulfur. Feed for each case consists of acid gas from both the acid gas cleanup unit (stream 31) and a vent stream from the sour water stripper in the gasifier section.

In the furnace waste heat boiler steam is generated. This steam is used to satisfy all Claus process preheating and reheating requirements as well as to provide some steam to the medium-pressure steam header. The sulfur condensers produce 0.45 MPa (65 psia) steam for the LP steam header.

**Power Block**

The methanation system and syngas cooler in the gasifier generates superheated steam that is used to power a commercially available ST using a 12.4MPa/566°C (1800 psig/1050°F) steam cycle (stream 36).

**Air Separation Unit (ASU)**

The ASU is designed to produce 99 mol% $O_2$ for use in the gasifier (stream 4) and SRU (stream 3). The plant is designed with four production trains. The main air compressor is powered by an electric motor. Nitrogen is vented to the atmosphere or used for ammonia synthesis in the co-production cases.

**Balance of Plant**

Balance of plant items were covered in Section 3.1.

### 4.1.2 Key System Assumptions

System assumptions for Cases 1 and 2, Siemens gasifier using Illinois No. 6 coal with and without carbon sequestration, are compiled in Exhibit 4-5.

**Exhibit 4-5 Cases 1 and 2 Plant Study Configuration Matrix**

| | Case 1 | Case 2 |
|---|---|---|
| Gasifier Pressure, MPa (psia) | 4.2 (615) | 4.2 (615) |
| $O_2$:Coal Ratio, kg $O_2$/kg dry coal | 0.784 | 0.784 |
| Carbon Conversion, % | 99.5 | 99.5 |
| Syngas HHV at Gasifier Outlet, kJ/Nm$^3$ (Btu/scf) | 10,283 (276) | 10,283 (276) |
| Steam Cycle, MPa/°C (psig/°F) | 12.4/566/534 (1800/1050/994) | 12.4/566/533 (1800/1050/992) |
| Condenser Pressure, mm Hg (in Hg) | 51 (2.0) | 51 (2.0) |
| Combustion Turbine | N/A | N/A |
| Gasifier Technology | Siemens | Siemens |
| Oxidant | 99 vol% Oxygen | 99 vol% Oxygen |
| Coal | Illinois No. 6 | Illinois No. 6 |
| Coal Feed Moisture Content, % | 11.12 | 11.12 |
| COS Hydrolysis Reactor | Yes | Yes |
| Water Gas Shift | Yes | Yes |
| $H_2S$ Separation | Selexol (1$^{st}$ Stage) | Selexol (1$^{st}$ Stage) |
| Sulfur Removal, % | 99.3 | 99.9 |
| $CO_2$ Separation | Selexol (2$^{nd}$ Stage) | Selexol (2$^{nd}$ Stage) |
| $CO_2$ Sequestered, % | N/A | 62.3 |
| Sulfur Recovery | Claus Plant with Tail Gas Treatment / Elemental Sulfur | Claus Plant with Tail Gas Treatment / Elemental Sulfur |
| Methanation System | Based on Haldor Topsoe TREMP™ Process | Based on Haldor Topsoe TREMP™ Process |
| Particulate Control | Scrubber and AGR Absorber | Scrubber and AGR Absorber |
| Mercury Control | Carbon Bed | Carbon Bed |
| NOx Control | N/A | N/A |

**Balance of Plant – All Cases**

The balance of plant assumptions are common to all cases and are presented in Exhibit 4-6.

**Exhibit 4-6  Balance of Plant Assumptions**

| | |
|---|---|
| **Cooling water system** | Recirculating Wet Cooling Tower |
| **Fuel and Other storage**<br>Coal | 30 days |
| Slag | 30 days |
| Sulfur | 30 days |
| Sorbent | 30 days |
| **Plant Distribution Voltage**<br>Motors below 1hp | 110/220 volt |
| Motors between 1 hp and 250 hp | 480 volt |
| Motors between 250 hp and 5,000 hp | 4,160 volt |
| Motors above 5,000 hp | 13,800 volt |
| Steam Generators | 24,000 volt |
| Grid Interconnection Voltage | 345 kV |
| **Water and Waste Water** | |
| Makeup Water | The water supply is 50 percent from a local Publicly Owned Treatment Works and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup requirements.<br>Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources |
| Process Wastewater | Water associated with gasification activity and storm water that contacts equipment surfaces is collected and treated for discharge through a permitted discharge. |
| Sanitary Waste Disposal | Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant was sized for 5.68 cubic meters per day (1,500 gallons per day) |
| Water Discharge | Most of the process wastewater is recycled to the cooling tower basin. Blowdown is treated for chloride and metals, and discharged. |

### 4.1.3 Sparing Philosophy

The sparing philosophy for Cases 1 and 2 is provided below. Single trains are utilized throughout with exceptions where equipment capacity requires an additional train. There is no redundancy other than normal sparing of rotating equipment.

The plant design consists of the following major subsystems:

- Three ASUs (3 x 33%).
- Six trains of coal drying and dry feed systems (6 x 17%).
- Six trains of gasification, including gasifier and SGC (6 x 17%).
- Three trains of syngas clean-up process (3 x 33%).
- Three trains of two-stage Selexol (3 x 33%).
- Three trains of methanation reactors (3 x 33%).
- Two trains of Claus-based sulfur recovery (2 x 50%).
- One ST (1 x 100%).

### 4.1.4 Cases 1 and 2 Performance Results

The Siemens SNG plant without and with carbon sequestration and using Illinois No. 6 coal produces an average net output of 57 Bscf/year at 90 percent CF. The HHV conversion efficiency of the non-sequestration case is 61.4 percent and 61.3 percent for the carbon sequestration case.

Overall performance for the two plants is summarized in Exhibit 4-7, which includes auxiliary power requirements and production values. The ASU accounts for approximately 68 percent of the total auxiliary load in the non-sequestration case and 54 percent in the carbon sequestration case, distributed between the main air compressor, the oxygen compressor, and ASU auxiliaries. $CO_2$ compression accounts for about 19 percent of the total auxiliary load for the carbon sequestration case. The AGR process accounts for about 13 percent and 11 percent of the auxiliary load for the non-sequestration and sequestration cases, respectively. All other individual auxiliary loads are less than 3 percent of the total.

**Exhibit 4-7 Cases 1 and 2 Plant Performance Summary**

| **POWER SUMMARY (Gross Power at Generator Terminals, kWe)** | **Case 1** | **Case 2** |
|---|---|---|
| Steam Turbine Power | 308,000 | 310,600 |
| **TOTAL POWER, kWe** | **308,000** | **310,600** |
| **AUXILIARY LOAD SUMMARY, kWe** | | |
| Coal Handling | 700 | 700 |
| Coal Milling | 4,500 | 4,500 |
| Slag Handling | 1,140 | 1,140 |
| Air Separation Unit Auxiliaries | 1,000 | 1,000 |
| Air Separation Unit Main Air Compressor | 120,220 | 120,220 |
| Oxygen Compressor | 20,910 | 20,910 |
| $CO_2$ Solid Feed System Compressors | 4,950 | 240 |
| SNG Compressors | 6,290 | 6,290 |
| Methanation Plant Recycle Compressor | 5,840 | 5,840 |
| Gasifier Pilot Fuel Compressor and Incinerator Air Blower | 300 | 300 |
| $CO_2$ Compressor | 0 | 50,400 |
| Boiler Feedwater Pumps | 4,980 | 5,000 |
| Condensate Pump | 420 | 430 |
| Quench Water Pump | 640 | 640 |
| Circulating Water Pump | 6,240 | 6,200 |
| Ground Water Pumps | 650 | 640 |
| Cooling Tower Fans | 3,640 | 3,610 |
| Scrubber Pumps | 1,000 | 1,000 |
| Acid Gas Removal | 27,790 | 27,750 |
| Steam Turbine Auxiliaries | 100 | 100 |
| Claus Plant/TGTU Auxiliaries | 250 | 250 |
| Claus Plant TG Recycle Compressor | 280 | 280 |
| Miscellaneous Balance of Plant[2] | 3,000 | 3,000 |
| Transformer Losses | 1,510 | 1,650 |
| **TOTAL AUXILIARIES, kWe** | **216,350** | **262,090** |
| **NET POWER, kWe** | **91,650** | **48,510** |
| SNG Production Rate, $MNm^3$/hr (Mscf/hr) | 202.9 (7,169) | 202.7 (7,163) |
| Capacity Factor | 90% | 90% |
| Net Exported Power Efficiency (HHV) | 2.8% | 1.5% |
| SNG Conversion Efficiency ($HHV_{product}/HHV_{coal}$), % | 61.4% | 61.3% |
| Coal to SNG Product Yield, $MNm^3_{SNG}/tonne_{coal}$ ($Mscf_{SNG}/ton_{coal}$) | 0.44 (14.9) | 0.44 (14.8) |
| Coal Feed Flow Rate, kg/hr (lb/hr) | 437,604 (964,752) | 437,604 (964,752) |
| Thermal Input,[1] kWth | 3,298,455 | 3,298,455 |
| Condenser Cooling Duty, GJ/hr (MMBtu/hr) | 1,709 (1,620) | 1,730 (1,640) |
| Raw Water Withdrawal, $m^3$/min (gpm) | 27.1 (7,169) | 27.0 (7,123) |
| Raw Water Consumption, $m^3$/min (gpm) | 21.6 (5,708) | 21.5 (5,672) |

1 - HHV of Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)
2 - Includes plant control systems, lighting, heating, ventilating, and air conditioning (HVAC), and miscellaneous low voltage loads

### Environmental Performance

The environmental targets for emissions of Hg, NOx, $SO_2$, and PM were discussed in Section 2.5. A summary of the plant air emissions for Cases 1 and 2 is presented in Exhibit 4-8. The first emissions metric is based on coal heat input to the gasifier.

**Exhibit 4-8 Cases 1 and 2 Air Emissions**

| | kg/GJ ($lb/10^6$ Btu) | | Tonne/year (ton/year) 90% capacity factor | |
|---|---|---|---|---|
| | **Case 1** | **Case 2** | **Case 1** | **Case 2** |
| **$SO_2$** | 0.012 (0.027) | 0.0001 (0.0003) | 1,103 (1,216) | 11 (12) |
| **$NO_X$** | Negligible | Negligible | Negligible | Negligible |
| **Particulates** | 0.003 (0.0071) | 0.003 (0.0071) | 286 (315) | 286 (315) |
| **Hg** | 2.46E-7 (5.71E-7) | 2.46E-7 (5.71E-7) | 0.023 (0.025) | 0.023 (0.025) |
| **$CO_2$** | 55 (128) | 2.1 (4.9) | 5,169,343 (5,698,226) | 195,324 (215,308) |

The low level of $SO_2$ emissions in Case 2 is achieved by capture of the sulfur in the gas by the two-stage Selexol AGR process and co-sequestration with the $CO_2$ product stream. The clean syngas exiting the AGR process has a sulfur concentration of less than 5 ppmv in both cases. The $H_2S$-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is treated using an amine based system to capture most of the remaining sulfur. The clean syngas from the tail gas treatment unit is combined with a slipstream of clean syngas from the Selexol process, passed through an incinerator, and the hot, inert incinerator offgas is used to dry coal prior to being vented to the atmosphere. The higher sulfur emissions in Case 1 are due to the sulfur contained in the $CO_2$-rich stream from the Selexol process, which is vented rather than sequestered.

$NO_X$ emissions are negligible because no CT is used for power generation.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Ninety five percent of mercury is captured from the syngas by an activated carbon bed. $CO_2$ emissions represent the uncontrolled (Case 1) and controlled (Case 2) discharge from the process.

The carbon balance for the two cases is shown in Exhibit 4-9. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not used in the carbon capture equation below, but it is not neglected in the balance since Aspen accounts for air components throughout. Carbon leaves the plant as SNG product, unburned carbon in the slag,

$CO_2$ in the coal dryer incinerator stack gas and ASU vent gas, and the $CO_2$ product (either vented or sequestered). The carbon capture efficiency is presented in two distinct ways. The first way defines capture as the amount of carbon in the $CO_2$ product for sequestration relative to the amount of carbon in the coal less carbon contained in the slag, and is represented by the following fraction:

(Carbon in Product for Sequestration)/[(Carbon in the Coal)-(Carbon in Slag)] or

Non sequestration (Case 1)
381,246/(614,979-3,075)*100 = 62.3% (Case 2)

The second way does not penalize the production facility for the carbon converted to SNG product, as the end use of the SNG is unknown. In a carbon constrained scenario, the SNG may or may not be used in a scenario where carbon capture and sequestration (CCS) is implemented (or to what extent the capture is accomplished). For this method the burden of carbon mitigation falls on the end-user. The following fraction represents this scenario:

[(Carbon in Product for Sequestration) + (Carbon in SNG Product)]/[(Carbon in the Coal)-(Carbon in Slag)] or
(0 + 215,937)/(614,979 – 3,075)*100 = 35.3% (Case 1)
(381,246 + 215,774)/(614,979-3,075)*100 = 97.6% (Case 2)

**Exhibit 4-9 Cases 1 and 2 Carbon Balance**

| Carbon In, kg/hr (lb/hr) | | | Carbon Out, kg/hr (lb/hr) | | |
|---|---|---|---|---|---|
| | **Case 1** | **Case 2** | | **Case 1** | **Case 2** |
| **Coal** | 278,950 (614,979) | 278,950 (614,979) | **Slag** | 1,395 (3,075) | 1,395 (3,075) |
| **Air ($CO_2$)** | 214 (472) | 214 (473) | **SNG Purification Off-Gas** | 14,062 (31,001) | N/A |
| | | | **ASU Vent** | 201 (442) | 200 (442) |
| | | | **SNG** | 97,947 (215,937) | 97,874 (215,774) |
| | | | **$CO_2$ Selexol Vent** | 138,526 (305,398) | N/A |
| | | | **$CO_2$ Lock Hopper Vent** | 21,126 (46,575) | 841 (1,855) |
| | | | **$CO_2$ Product** | N/A | 172,930 (381,246) |
| | | | **Stack** | 5,907 (13,023) | 5,924 (13,061) |
| **Total** | 279,164 (615,451) | 279,164 (615,451) | **Total** | 279,164 (615,451) | 279,164 (615,451) |

Exhibit 4-10 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the stack gas, and sulfur co-sequestered with the $CO_2$ product. Sulfur in the slag is considered negligible. The total sulfur capture is represented by the following fraction:

(Sulfur byproduct + Sulfur in $CO_2$ product)/Sulfur in the coal or
(24,003 + 0)/24,181 = 99.3% (Case 1)
(24,010 + 157)/24,181 = 99.9% (Case 2)

**Exhibit 4-10 Cases 1 and 2 Sulfur Balance**

| Sulfur In, kg/hr (lb/hr) | | | Sulfur Out, kg/hr (lb/hr) | | |
|---|---|---|---|---|---|
| | **Case 1** | **Case 2** | | **Case 1** | **Case 2** |
| **Coal** | 10,968 (24,181) | 10,968 (24,181) | **Elemental Sulfur** | 10,888 (24,003) | 10,891 (24,010) |
| | | | **SNG** | 5 (12) | 5 (12) |
| | | | **$CO_2$ Vent Streams/Stack** | 75 (166) | 1 (3) |
| | | | **$CO_2$ Product** | N/A | 71 (157) |
| **Total** | 10,968 (24,181) | 10,968 (24,181) | **Total** | 10,968 (24,181) | 10,968 (24,181) |

Exhibit 4-11 shows the overall water balance for the plant. Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as syngas condensate, and is re-used as internal recycle. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is defined as the water removed from the ground or diverted from a surface-water source for use in the plant and was assumed to be provided 50 percent by a POTW and 50 percent from groundwater. Raw water withdrawal can be represented by the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, quench system makeup, and slag handling makeup. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source.

**Exhibit 4-11 Cases 1 and 2 Water Balance**

| | Case 1 | Case 2 |
|---|---|---|
| **Water Demand, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.95 (251)** | **0.95 (251)** |
| **Quench/Wash** | **4.2 (1,114)** | **4.2 (1,113)** |
| **Condenser Makeup** | **5.9 (1,548)** | **5.9 (1,547)** |
| Gasifier Steam | 1.0 (274) | 1.0 (274) |
| Shift Steam | 4.4 (1,171) | 4.4 (1,170) |
| BFW Makeup | 0.39 (103) | 0.39 (103) |
| **Cooling Tower** | **24.3 (6,417)** | **24.1 (6,370)** |
| **Total** | **35.3 (9,330)** | **35.1 (9,282)** |
| **Internal Recycle, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.95 (251)** | **0.95 (251)** |
| **Quench/Wash** | **4.2 (1,114)** | **4.2 (1,113)** |
| **Condenser Makeup** | **0.0 (0)** | **0.0 (0)** |
| **Cooling Tower** | **3.0 (796)** | **3.0 (794)** |
| Water from Coal Drying | 0.0 (0) | 0.0 (0) |
| BFW Blowdown | 0.39 (103) | 0.39 (103) |
| SWS Blowdown | 0.71 (187) | 0.71 (187) |
| SWS Excess | 1.9 (505) | 1.9 (504) |
| **Total** | **8.2 (2,161)** | **8.2 (2,159)** |
| **Raw Water Withdrawal, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.0 (0)** | **0.0 (0)** |
| **Quench/Wash** | **0.0 (0)** | **0.0 (0)** |
| **Condenser Makeup** | **5.9 (1,548)** | **5.9 (1,547)** |
| Gasifier Steam | 1.0 (274) | 1.0 (274) |
| Shift Steam | 4.4 (1,171) | 4.4 (1,170) |
| BFW Makeup | 0.39 (103) | 0.39 (103) |
| **Cooling Tower** | **21.3 (5,621)** | **21.1 (5,576)** |
| **Total** | **27.1 (7,169)** | **27.0 (7,123)** |
| **Process Water Discharge, $m^3$/min (gpm)** | | |
| **SWS Blowdown** | **0.07 (19)** | **0.07 (19)** |
| **Cooling Tower Blowdown** | **5.5 (1,443)** | **5.4 (1,433)** |
| **Total** | **5.5 (1,462)** | **5.5 (1,451)** |
| **Raw Water Consumption, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.0 (0)** | **0.0 (0)** |
| **Quench/Wash** | **0.0 (0)** | **0.0 (0)** |
| **SWS Blowdown** | **-0.07 (-19)** | **-0.07 (-19)** |
| **Condenser Makeup** | **5.9 (1,548)** | **5.9 (1,547)** |
| **Cooling Tower** | **15.8 (4,178)** | **15.7 (4,144)** |
| **Total** | **21.6 (5,708)** | **21.5 (5,672)** |

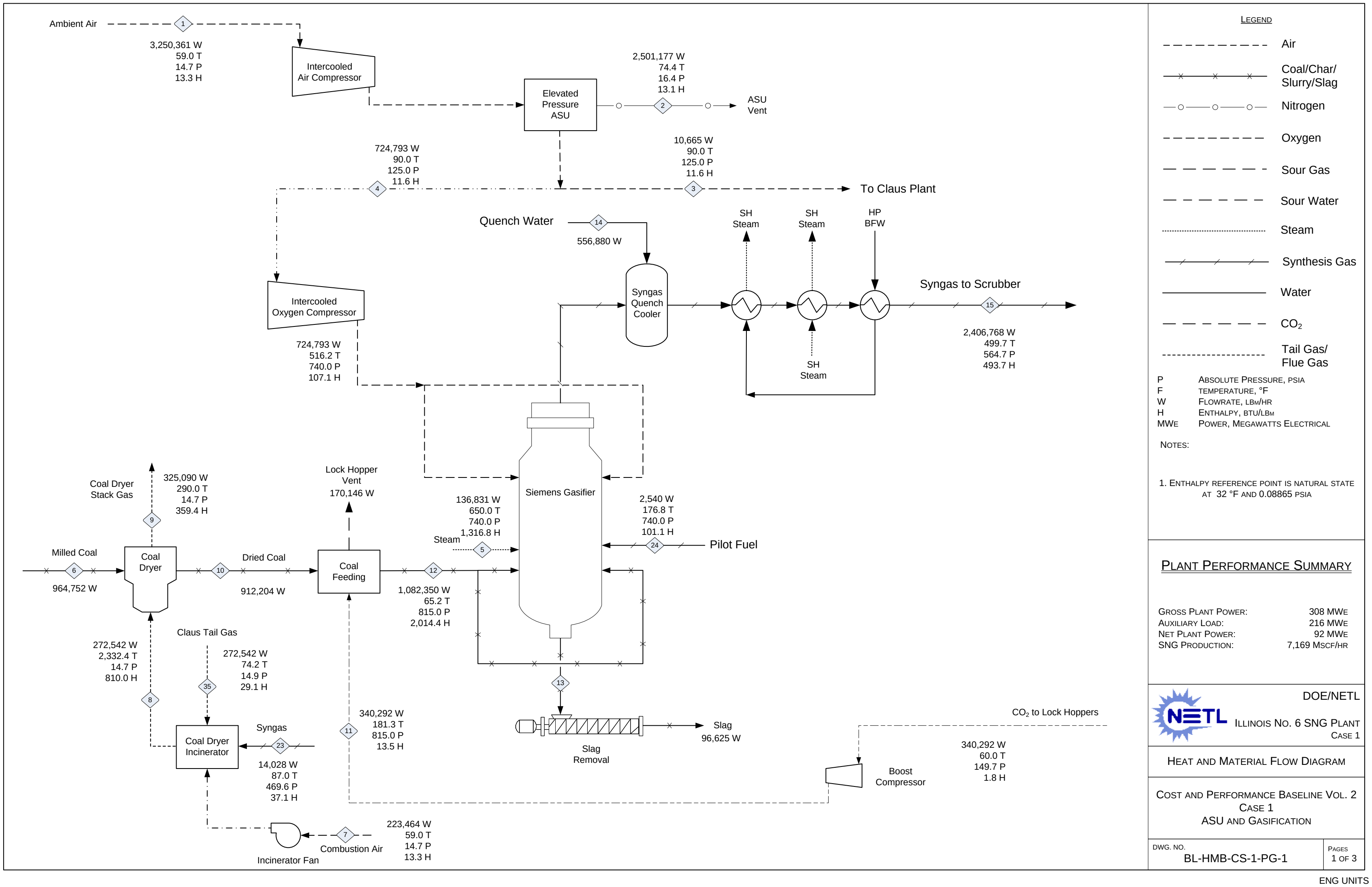
### Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 4-12 through Exhibit 4-17:

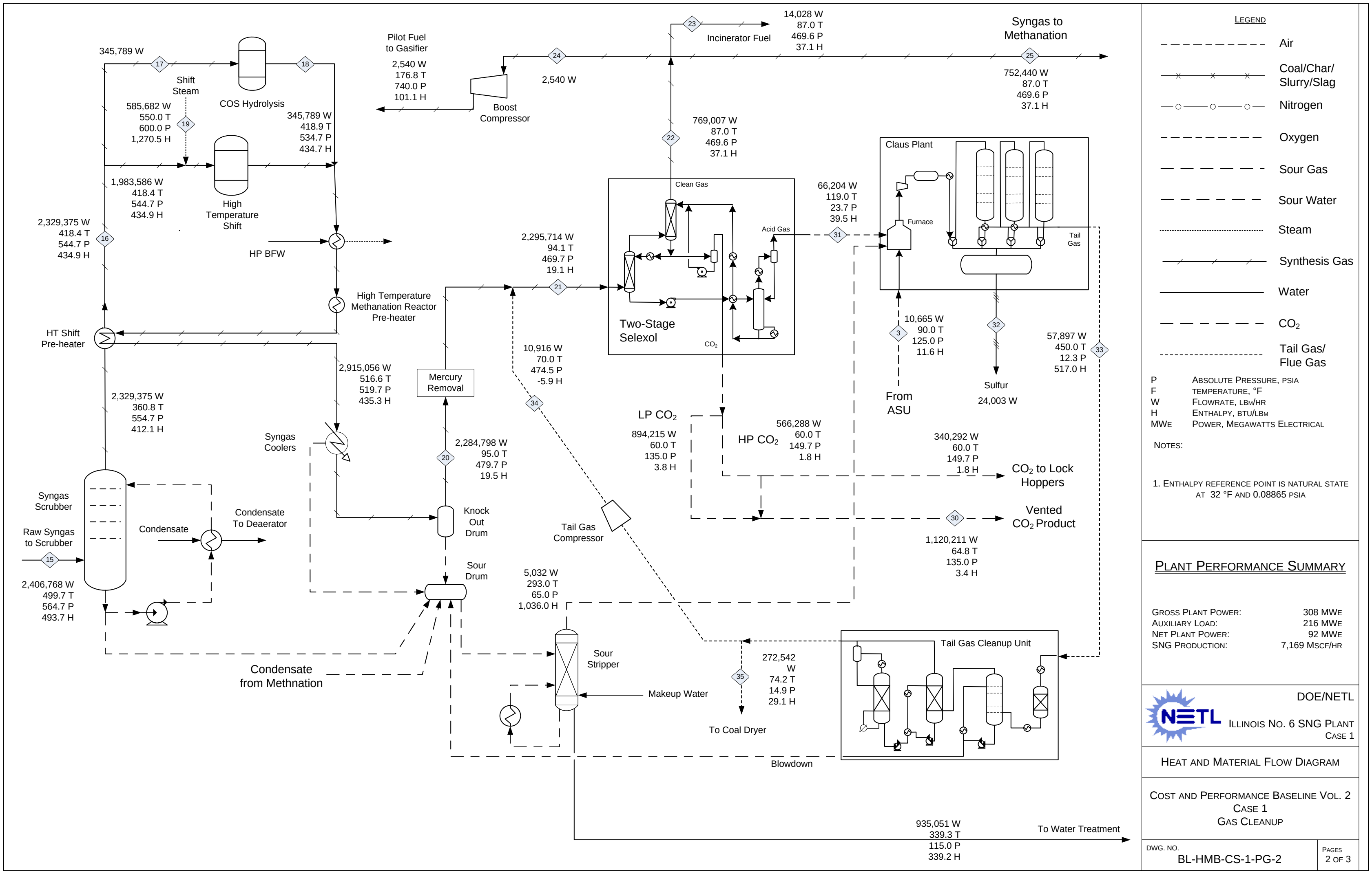
- Coal gasification and ASU
- Syngas cleanup (including sulfur recovery and tail gas recycle)
- Methanation, SNG purification, compression, and power cycle

An overall plant energy balance is provided in tabular form in Exhibit 4-18 for the two cases.

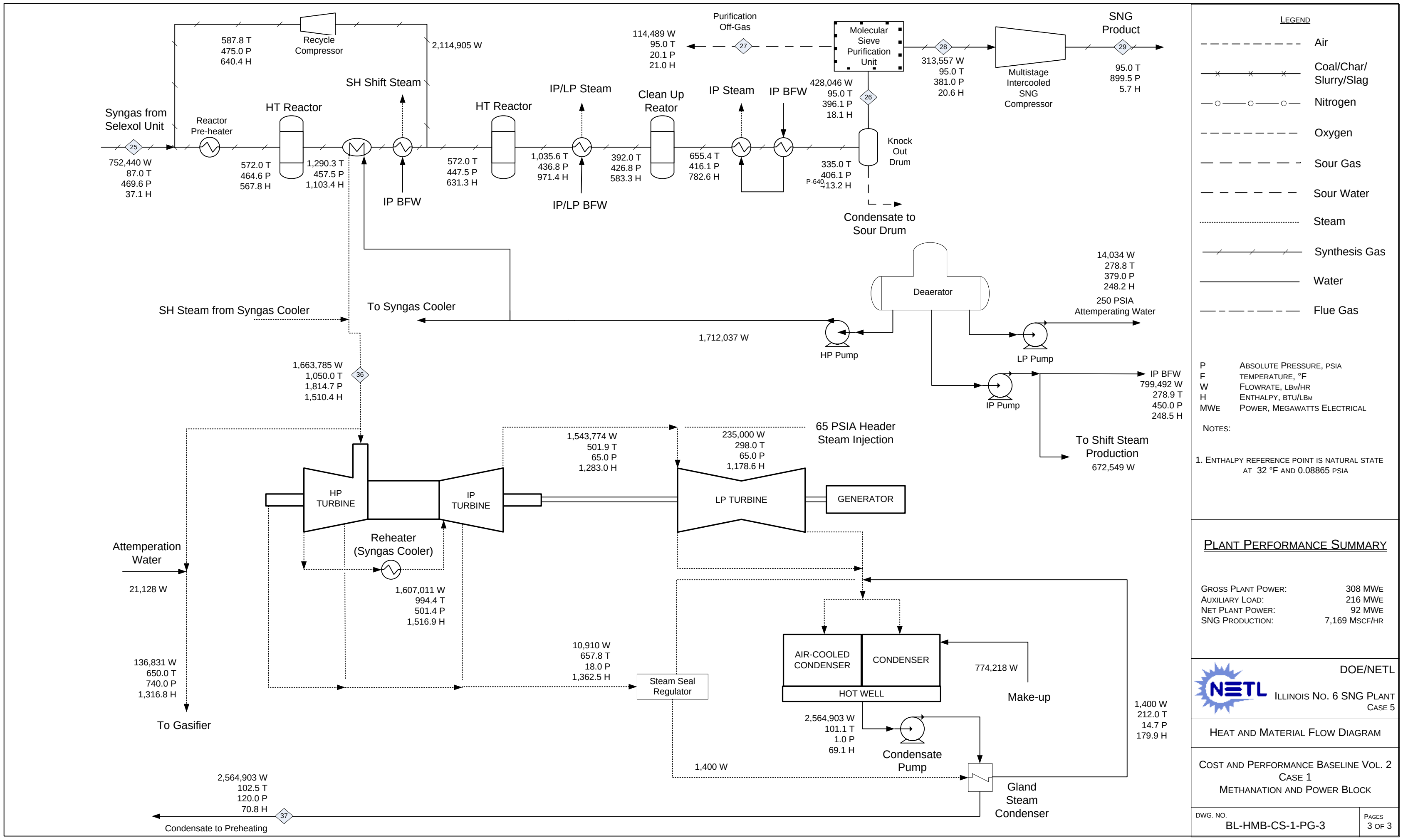
## Exhibit 4-12  Case 1 Gasification and ASU Heat and Mass Balance



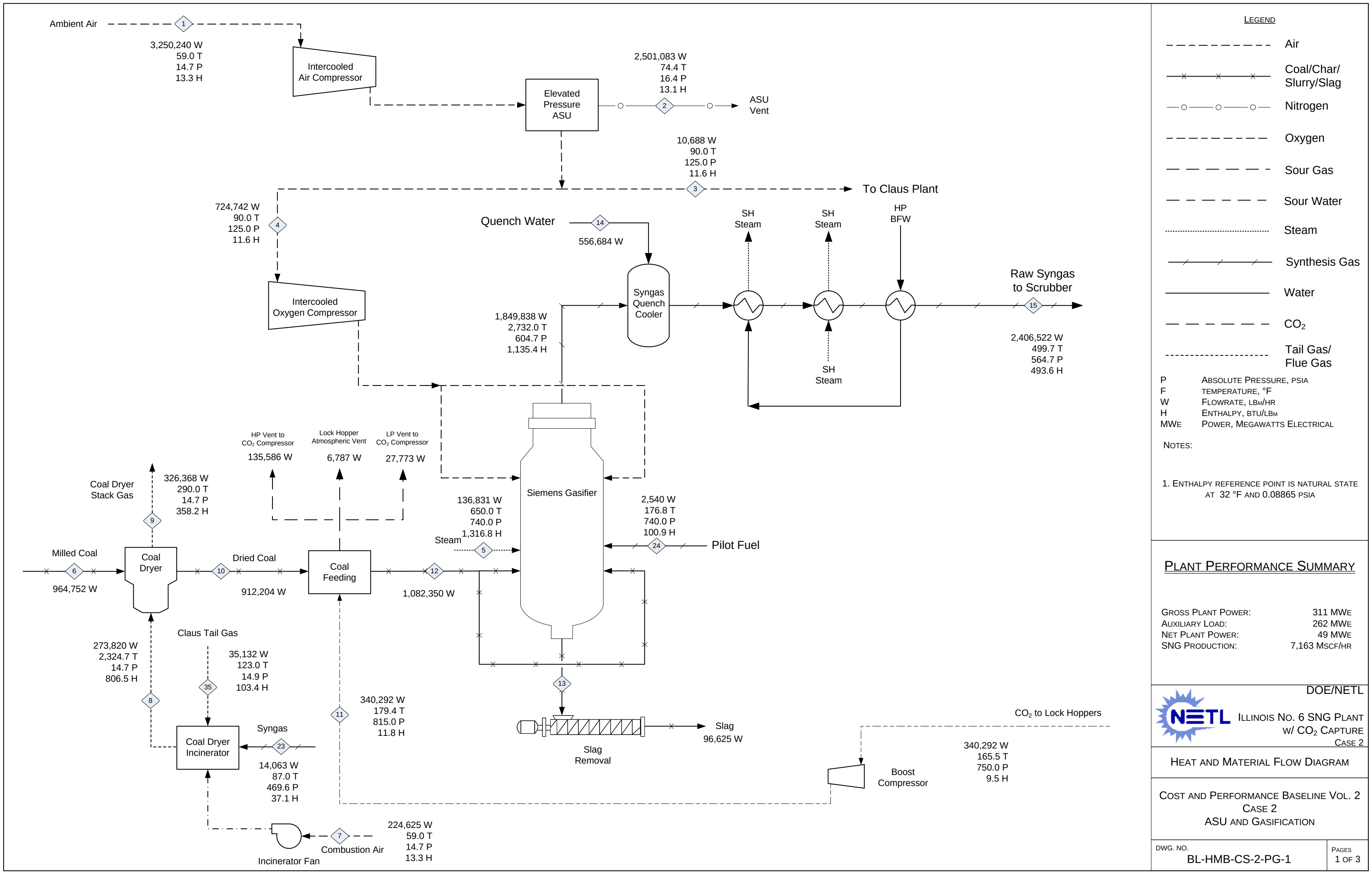
## Exhibit 4-13 Case 1 Gas Cleanup Heat and Mass Balance



## Exhibit 4-14 Case 1 Methanation and Power Block Heat and Mass Balance



## Exhibit 4-15 Case 2 Gasifier and ASU Heat and Mass Balance

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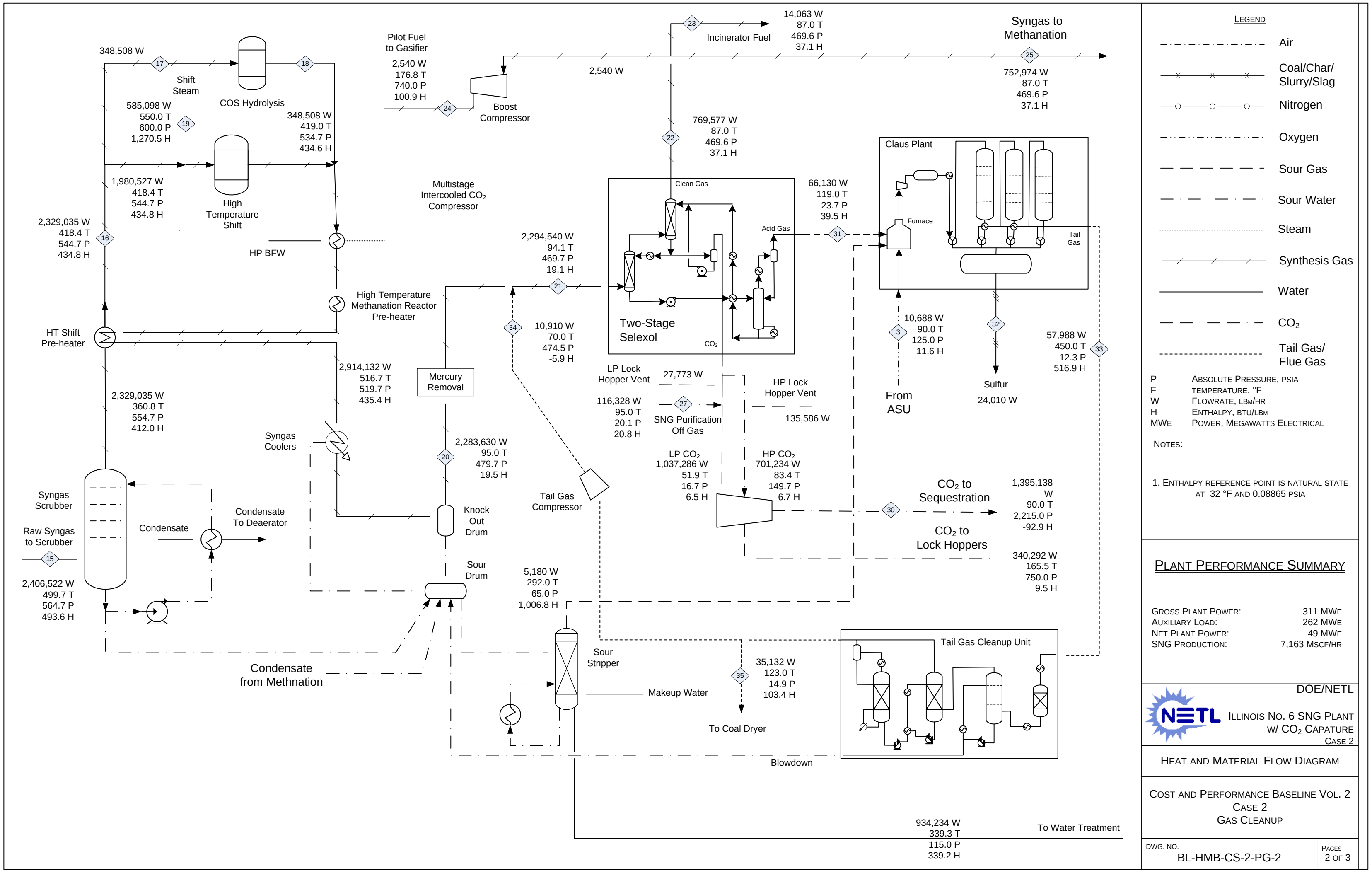
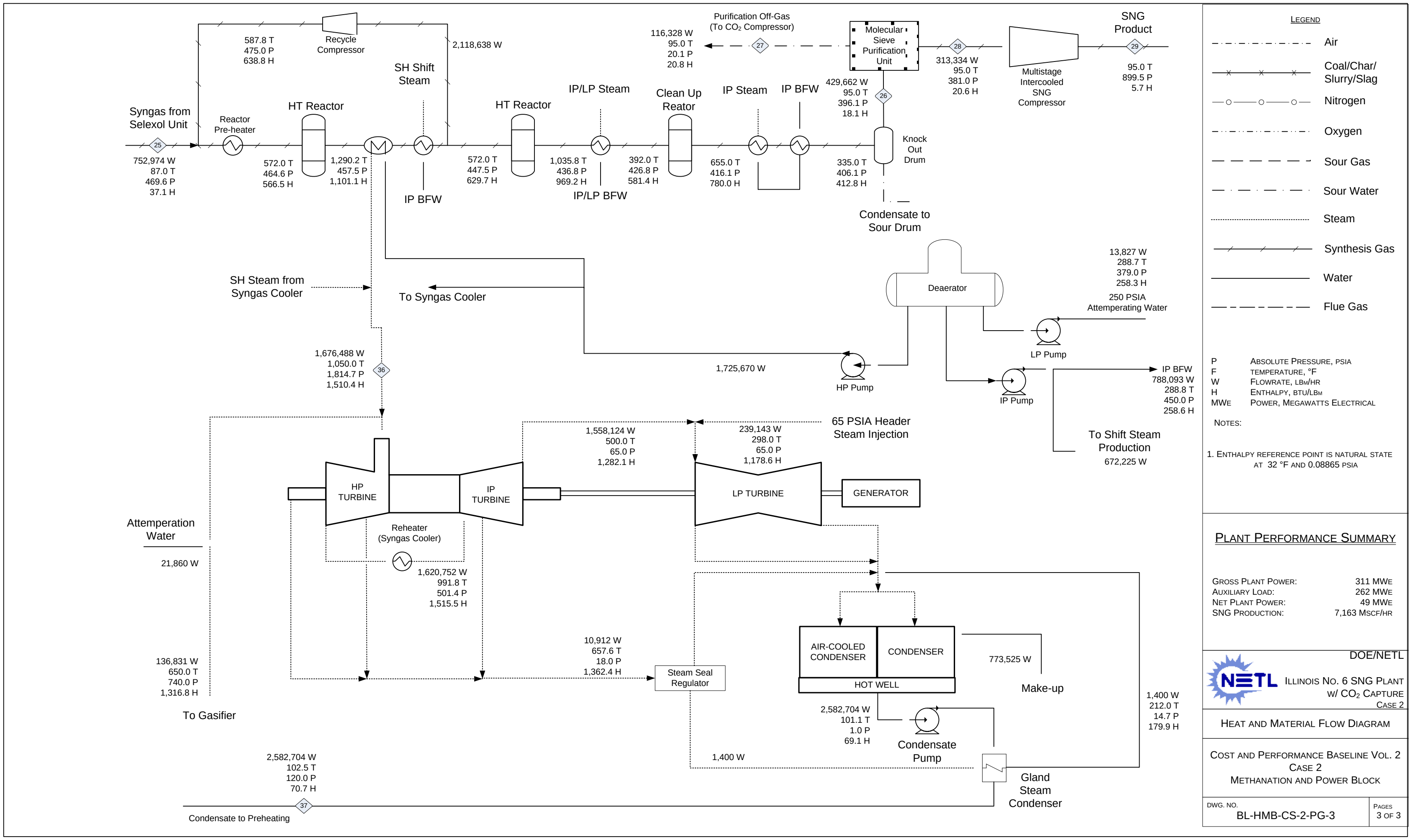
# Exhibit 4-16 Case 2 Gas Cleanup Heat and Mass Balance



# Exhibit 4-17 Case 2 Methanation and Power Block Heat and Mass Balance




To Gasifier

**Exhibit 4-18 Cases 1 and 2 Energy Balance**

| | HHV | | Sensible + Latent | | Power | | Total | |
|---|---|---|---|---|---|---|---|---|
| | Case 1 | Case 2 | Case 1 | Case 2 | Case 1 | Case 2 | Case 1 | Case 2 |
| **Heat In, GJ/hr (MMBtu/hr)** | | | | | | | | |
| Coal | 11,874 (11,255) | 11,874 (11,255) | 12.1 (11.5) | 12.1 (11.5) | 0 (0) | 0 (0) | 11,887 (11,266) | 11,887 (11,266) |
| ASU Air | 0 (0) | 0 (0) | 48.7 (46.2) | 49.6 (47.0) | 0 (0) | 0 (0) | 49 (46) | 50 (47) |
| Raw Water Makeup | 0 (0) | 0 (0) | 102.0 (96.7) | 101.4 (96.1) | 0 (0) | 0 (0) | 102 (97) | 101 (96) |
| Auxiliary Power | 0 (0) | 0 (0) | 0.0 (0.0) | 0.0 (0.0) | 779 (738) | 944 (894) | 779 (738) | 944 (894) |
| **Totals** | **11,874 (11,255)** | **11,874 (11,255)** | **162.9 (154.4)** | **163.1 (154.6)** | **779 (738)** | **944 (894)** | **12,816 (12,147)** | **12,981 (12,304)** |
| **Heat Out, GJ/hr (MMBtu/hr)** | | | | | | | | |
| ASU Intercoolers | 0 (0) | 0 (0) | 325 (308) | 325 (308) | 0 (0) | 0 (0) | 325 (308) | 325 (308) |
| ASU Vent | 0 (0) | 0 (0) | 34.7 (32.9) | 34.7 (32.9) | 0 (0) | 0 (0) | 35 (33) | 35 (33) |
| Slag | 46 (43) | 46 (43) | 79.1 (75.0) | 79.1 (75.0) | 0 (0) | 0 (0) | 125 (118) | 125 (118) |
| Coal Drying Stack Gas | 0 (0) | 0 (0) | 123.3 (116.8) | 123.3 (116.9) | 0 (0) | 0 (0) | 123 (117) | 123 (117) |
| Sulfur | 101 (96) | 101 (96) | 1.2 (1.2) | 1.2 (1.2) | 0 (0) | 0 (0) | 102 (97) | 102 (97) |
| Auxiliary Cooling Loads | 0 (0) | 0 (0) | 882 (836) | 842 (798) | 0 (0) | 0 (0) | 882 (836) | 842 (798) |
| $CO_2$ | 0 (0) | 0 (0) | 6.5 (6.1) | -136.7 (-129.6) | 0 (0) | 0 (0) | 6 (6) | -137 (-130) |
| Cooling Tower Blowdown | 0 (0) | 0 (0) | 40.6 (38.5) | 40.3 (38.2) | 0 (0) | 0 (0) | 41 (38) | 40 (38) |
| SNG | 7,272 (6,892) | 7,266 (6,887) | 1.9 (1.8) | 1.9 (1.8) | 0 (0) | 0 (0) | 7,274 (6,894) | 7,268 (6,888) |
| SNG Purification Off-Gas | 0 (0) | 0 (0) | 3 (2) | 3 (2) | 0 (0) | 0 (0) | 3 (2) | 3 (2) |
| Condenser | 0 (0) | 0 (0) | 1,710 (1,621) | 1,727 (1,637) | 0 (0) | 0 (0) | 1,710 (1,621) | 1,727 (1,637) |
| Process Losses* | 0 (0) | 0 (0) | 1,082 (1,026) | 1,410 (1,336) | 0 (0) | 0 (0) | 1,082 (1,026) | 1,410 (1,336) |
| Power | 0 (0) | 0 (0) | 0.0 (0.0) | 0.0 (0.0) | 1,109 (1,051) | 1,118 (1,060) | 1,109 (1,051) | 1,118 (1,060) |
| **Totals** | **7,418 (7,031)** | **7,412 (7,026)** | **4,289 (4,065)** | **4,450 (4,218)** | **1,109 (1,051)** | **1,118 (1,060)** | **12,816 (12,147)** | **12,981 (12,304)** |

* Includes other energy losses not explicitly accounted for in the model.

### 4.1.5 Cases 1 and 2 Equipment Lists

Major equipment items for the Siemens gasifier SNG plant without and with carbon sequestration using Illinois No. 6 coal are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 4.1.6. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

**ACCOUNT 1 COAL HANDLING**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Bottom Trestle Dumper and Receiving Hoppers | N/A | 2 | 0 | 181 tonne (200 ton) | 181 tonne (200 ton) |
| 2 | Feeder | Belt | 2 | 0 | 572 tonne/hr (630 tph) | 572 tonne/hr (630 tph) |
| 3 | Conveyor No. 1 | Belt | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 4 | Transfer Tower No. 1 | Enclosed | 1 | 0 | N/A | N/A |
| 5 | Conveyor No. 2 | Belt | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 6 | As-Received Coal Sampling System | Two-stage | 1 | 0 | N/A | N/A |
| 7 | Stacker/Reclaimer | Traveling, linear | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 8 | Reclaim Hopper | N/A | 2 | 1 | 91 tonne (100 ton) | 91 tonne (100 ton) |
| 9 | Feeder | Vibratory | 2 | 1 | 363 tonne/hr (400 tph) | 363 tonne/hr (400 tph) |
| 10 | Conveyor No. 3 | Belt w/ tripper | 1 | 0 | 726 tonne/hr (800 tph) | 726 tonne/hr (800 tph) |
| 11 | Crusher Tower | N/A | 1 | 0 | N/A | N/A |
| 12 | Coal Surge Bin w/ Vent Filter | Dual outlet | 2 | 0 | 363 tonne (400 ton) | 363 tonne (400 ton) |
| 13 | Crusher | Impactor reduction | 2 | 0 | 8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0) | 8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0) |
| 14 | As-Fired Coal Sampling System | Swing hammer | 1 | 1 | N/A | N/A |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 15 | Conveyor No. 4 | Belt w/tripper | 1 | 0 | 726 tonne/hr (800 tph) | 726 tonne/hr (800 tph) |
| 16 | Transfer Tower No. 2 | Enclosed | 1 | 0 | N/A | N/A |
| 17 | Conveyor No. 5 | Belt w/ tripper | 1 | 0 | 726 tonne/hr (800 tph) | 726 tonne/hr (800 tph) |
| 18 | Coal Silo w/ Vent Filter and Slide Gates | Field erected | 3 | 0 | 1,633 tonne (1,800 ton) | 1,633 tonne (1,800 ton) |

**ACCOUNT 2 COAL PREPARATION AND FEED**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Feeder | Vibratory | 6 | 0 | 82 tonne/hr (90 tph) | 82 tonne/hr (90 tph) |
| 2 | Conveyor No. 6 | Belt w/tripper | 3 | 0 | 163 tonne/hr (180 tph) | 163 tonne/hr (180 tph) |
| 3 | Roller Mill Feed Hopper | Dual Outlet | 3 | 0 | 318 tonne (350 ton) | 318 tonne (350 ton) |
| 4 | Weigh Feeder | Belt | 6 | 0 | 82 tonne/hr (90 tph) | 82 tonne/hr (90 tph) |
| 5 | Pulverizer | Rotary | 6 | 0 | 82 tonne/hr (90 tph) | 82 tonne/hr (90 tph) |

**ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Demineralized Water Storage Tank | Vertical, cylindrical, outdoor | 3 | 0 | 635,949 liters (168,000 gal) | 635,949 liters (168,000 gal) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 2 | Condensate Pumps | Vertical canned | 2 | 1 | 10,751 lpm @ 91 m $H_2O$ (2,840 gpm @ 300 ft $H_2O$) | 10,826 lpm @ 91 m $H_2O$ (2,860 gpm @ 300 ft $H_2O$) |
| 3 | Deaerator | Horizontal spray type | 2 | 0 | 806,034 kg/hr (1,777,000 lb/hr) | 806,487 kg/hr (1,778,000 lb/hr) |
| 4 | Intermediate Pressure Feedwater Pump | Horizontal centrifugal, single stage | 2 | 1 | 6,359 lpm @ 341 m $H_2O$ (1,680 gpm @ 1120 ft $H_2O$) | 6,322 lpm @ 335 m $H_2O$ (1,670 gpm @ 1100 ft $H_2O$) |
| 5 | High Pressure Feedwater Pump | Barrel type, multi-stage, centrifugal | 2 | 1 | HP water: 7,419 lpm @ 1,676 m $H_2O$ (1,960 gpm @ 5,500 ft $H_2O$) | HP water: 7,457 lpm @ 1,676 m $H_2O$ (1,970 gpm @ 5,500 ft $H_2O$) |
| 6 | Auxiliary Boiler | Shop fabricated, water tube | 1 | 0 | 18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F) | 18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F) |
| 7 | Service Air Compressors | Flooded Screw | 3 | 1 | 28 $m^3$/min @ 0.7 MPa (1,000 scfm @ 100 psig) | 28 $m^3$/min @ 0.7 MPa (1,000 scfm @ 100 psig) |
| 8 | Instrument Air Dryers | Duplex, regenerative | 3 | 1 | 28 $m^3$/min (1,000 scfm) | 28 $m^3$/min (1,000 scfm) |
| 9 | Closed Cycle Cooling Heat Exchangers | Plate and frame | 2 | 0 | 664 GJ/hr (629 MMBtu/hr) | 642 GJ/hr (608 MMBtu/hr) each |
| 10 | Closed Cycle Cooling Water Pumps | Horizontal centrifugal | 2 | 1 | 238,102 lpm @ 21 m $H_2O$ (62,900 gpm @ 70 ft $H_2O$) | 230,153 lpm @ 21 m $H_2O$ (60,800 gpm @ 70 ft $H_2O$) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 11 | Engine-Driven Fire Pump | Vertical turbine, diesel engine | 1 | 1 | 3,785 lpm @ 107 m $H_2O$ (1,000 gpm @ 350 ft $H_2O$) | 3,785 lpm @ 107 m $H_2O$ (1,000 gpm @ 350 ft $H_2O$) |
| 12 | Fire Service Booster Pump | Two-stage horizontal centrifugal | 1 | 1 | 2,650 lpm @ 76 m $H_2O$ (700 gpm @ 250 ft $H_2O$) | 2,650 lpm @ 76 m $H_2O$ (700 gpm @ 250 ft $H_2O$) |
| 13 | Raw Water Pumps | Stainless steel, single suction | 2 | 1 | 7,533 lpm @ 18 m $H_2O$ (1,990 gpm @ 60 ft $H_2O$) | 7,457 lpm @ 18 m $H_2O$ (1,970 gpm @ 60 ft $H_2O$) |
| 14 | Ground Water Pumps | Stainless steel, single suction | 4 | 1 | 3,748 lpm @ 268 m $H_2O$ (990 gpm @ 880 ft $H_2O$) | 3,748 lpm @ 268 m $H_2O$ (990 gpm @ 880 ft $H_2O$) |
| 15 | Filtered Water Pumps | Stainless steel, single suction | 2 | 1 | 6,095 lpm @ 49 m $H_2O$ (1,610 gpm @ 160 ft $H_2O$) | 6,095 lpm @ 49 m $H_2O$ (1,610 gpm @ 160 ft $H_2O$) |
| 16 | Filtered Water Tank | Vertical, cylindrical | 2 | 0 | 2,933,694 liter (775,000 gal) | 2,929,909 liter (774,000 gal) |
| 17 | Makeup Water Demineralizer | Anion, cation, and mixed bed | 2 | 0 | 2,688 lpm (710 gpm) | 2,688 lpm (710 gpm) |
| 18 | Liquid Waste Treatment System | | 1 | 0 | 10 years, 24-hour storm | 10 years, 24-hour storm |

**ACCOUNT 4 GASIFIER, ASU, AND ACCESSORIES INCLUDING LOW TEMPERATURE HEAT RECOVERY**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Gasifier | Pressurized dry-feed, entrained bed | 6 | 0 | 1,905 tonne/day, 4.2 MPa (2,100 tpd, 615 psia) | 1,905 tonne/day, 4.2 MPa (2,100 tpd, 615 psia) |
| 2 | Synthesis Gas Cooler | Convective spiral-wound tube boiler | 6 | 0 | 200,034 kg/hr (441,000 lb/hr) | 200,034 kg/hr (441,000 lb/hr) |
| 3 | Syngas Scrubber Including Sour Water Stripper | Vertical upflow | 6 | 0 | 200,034 kg/hr (441,000 lb/hr) | 200,034 kg/hr (441,000 lb/hr) |
| 4 | Raw Gas Coolers | Shell and tube with condensate drain | 12 | 0 | 415,491 kg/hr (916,000 lb/hr) | 415,491 kg/hr (916,000 lb/hr) |
| 5 | Raw Gas Knockout Drum | Vertical with mist eliminator | 3 | 0 | 380,110 kg/hr, 35°C, 3.3 MPa (838,000 lb/hr, 95°F, 485 psia) | 379,657 kg/hr, 35°C, 3.3 MPa (837,000 lb/hr, 95°F, 485 psia) |
| 6 | Flare Stack | Self-supporting, carbon steel, stainless steel top, pilot ignition | 3 | 0 | 400,068 kg/hr (882,000 lb/hr) syngas | 400,068 kg/hr (882,000 lb/hr) syngas |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 7 | ASU Main Air Compressor | Centrifugal, multi-stage | 3 | 0 | 7,391 $m^3$/min @ 1.3 MPa (261,000 scfm @ 190 psia) | 7,391 $m^3$/min @ 1.3 MPa (261,000 scfm @ 190 psia) |
| 8 | Cold Box | Vendor design | 3 | 0 | 2,903 tonne/day (3,200 tpd) of 99% purity oxygen | 2,903 tonne/day (3,200 tpd) of 99% purity oxygen |
| 9 | Oxygen Compressor | Centrifugal, multi-stage | 3 | 0 | 1,472 $m^3$/min (52,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia) | 1,472 $m^3$/min (52,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia) |
| 10 | Coal Feed System $CO_2$ Compressor | Centrifugal, multi-stage | 3 | 0 | 510 $m^3$/min (18,000 scfm) Suction - 1.0 MPa (150 psia) Discharge - 5.7 MPa (820 psia) | 510 $m^3$/min (18,000 scfm) Suction - 5.2 MPa (750 psia) Discharge - 5.7 MPa (820 psia) |

**ACCOUNT 5A SYNGAS CLEANUP**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Mercury Adsorber | Sulfated carbon bed | 3 | 0 | 380,110 kg/hr (838,000 lb/hr) 35°C (95°F) 3.3 MPa (480 psia) | 379,657 kg/hr (837,000 lb/hr) 35°C (95°F) 3.3 MPa (480 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 2 | Zinc Oxide Guard Bed | Fixed bed | 1 | 0 | 450 $Nm^3$/min (15,891 acfm) | 450 $Nm^3$/min (15,877 acfm) |
| 3 | Sulfur Plant | Claus type | 2 | 0 | 144 tonne/day (158 tpd) | 144 tonne/day (158 tpd) |
| 4 | COS Hydrolysis Reactor | Fixed bed, catalytic | 3 | 0 | 57,606 kg/hr (127,000 lb/hr) 216°C (420°F) 3.7 MPa (540 psia) | 58,060 kg/hr (128,000 lb/hr) 216°C (420°F) 3.7 MPa (540 psia) |
| 5 | Water Gas Shift Reactor | Fixed bed, catalytic | 3 | 0 | 427,284 kg/hr (942,000 lb/hr) 232°C (450°F) 3.7 MPa (540 psia) | 426,830 kg/hr (941,000 lb/hr) 232°C (450°F) 3.7 MPa (540 psia) |
| 6 | Shift Reactor Heat Recovery Exchanger | Shell and Tube | 3 | 0 | Exchanger 1: 45 GJ/hr (43 MMBtu/hr) Exchanger 2: 81 GJ/hr (77 MMBtu/hr) Exchanger 3: 18 GJ/hr (17 MMBtu/hr) | Exchanger 1: 45 GJ/hr (43 MMBtu/hr) Exchanger 2: 81 GJ/hr (77 MMBtu/hr) Exchanger 3: 18 GJ/hr (17 MMBtu/hr) |
| 7 | Acid Gas Removal Plant | Two-stage Selexol | 3 | 0 | 381,925 kg/hr (842,000 lb/hr) 34°C (94°F) 3.2 MPa (470 psia) | 381,471 kg/hr (841,000 lb/hr) 34°C (94°F) 3.2 MPa (470 psia) |
| 8 | Tail Gas Treatment Unit | Proprietary amine, absorber/stripper | 2 | 0 | 11,552 kg/hr (25,467 lb/hr) 49°C (120°F) 0.1 MPa (10.6 psia) | 11,570 kg/hr (25,508 lb/hr) 49°C (120°F) 0.1 MPa (10.6 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 9 | Tail Gas Treatment Incinerator | N/A | 2 | 0 | 59,021 kg/hr (130,120 lb/hr) 23°C (74°F) 0.1 MPa (14.9 psia) | 59,021 kg/hr (130,120 lb/hr) 23°C (74°F) 0.1 MPa (14.9 psia) |
| 10 | Tail Gas Recycle Compressor | Centrifugal | 2 | 1 | 2,807 kg/hr @ 03.3 MPa (6,189 lb/hr @ 475 psia) | 2,806 kg/hr @ 03.3 MPa (6,186 lb/hr @ 475 psia) |

**ACCOUNT 5A SNG PRODUCTION/METHANATION SYSTEM**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Methanation Reactor Preheater | Shell and Tube | 3 | 0 | 95 GJ/hr (90 MMBtu/hr) | 95 GJ/hr (90 MMBtu/hr) |
| 2 | Methanation Reactor 1 | Fixed Bed, catalytic | 3 | 0 | 476,726 kg/hr (1,051,000 lb/hr) 699°C (1,290°F) 3.2 MPa (460 psia) | 477,633 kg/hr (1,053,000 lb/hr) 699°C (1,290°F) 3.2 MPa (460 psia) |
| 3 | Methanation Reactor Intercooler 1 | Shell and Tube | 3 | 0 | 413 GJ/hr (391 MMBtu/hr) | 413 GJ/hr (391 MMBtu/hr) |
| 4 | Methanation Reactor Intercooler 2 | Shell and Tube | 3 | 0 | 111 GJ/hr (105 MMBtu/hr) | 111 GJ/hr (105 MMBtu/hr) |
| 5 | Methanation Reactor 2 | Fixed Bed, catalytic | 3 | 0 | 125,191 kg/hr (276,000 lb/hr) 560°C (1040°F) 3.1 MPa (450 psia) | 125,191 kg/hr (276,000 lb/hr) 560°C (1040°F) 3.1 MPa (450 psia) |
| 6 | Methanation Reactor Intercooler 3 | Shell and Tube | 3 | 0 | 29 GJ/hr (28 MMBtu/hr) | 29 GJ/hr (28 MMBtu/hr) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 7 | Methanation Reactor 3 | Fixed Bed, catalytic | 3 | 0 | 125,191 kg/hr (276,000 lb/hr) 349°C (660°F) 3.0 MPa (430 psia) | 125,191 kg/hr (276,000 lb/hr) 343°C (650°F) 3.0 MPa (430 psia) |
| 8 | Methanation Reactor Intercooler 4 | Shell and Tube | 3 | 0 | 41 GJ/hr (39 MMBtu/hr) | 41 GJ/hr (39 MMBtu/hr) |
| 9 | Methanation Reactor Intercooler 5 | Shell and Tube | 3 | 0 | 66 GJ/hr (63 MMBtu/hr) | 66 GJ/hr (62 MMBtu/hr) |
| 10 | SNG Purification Condenser 1 | Shell and Tube | 3 | 0 | 55 GJ/hr (52 MMBtu/hr) | 55 GJ/hr (52 MMBtu/hr) |
| 11 | SNG Purification Condenser 2 | Shell and Tube | 3 | 0 | 33 GJ/hr (31 MMBtu/hr) | 33 GJ/hr (32 MMBtu/hr) |
| 12 | Methanation Recycle Compressor | Centrifugal | 3 | 1 | 8,668 $m^3$/min (306,100 scfm) Suction - 3.1 MPa (448 psia) Discharge - 3.3 MPa (475 psia) | 8,668 $m^3$/min (306,100 scfm) Suction - 3.1 MPa (448 psia) Discharge - 3.3 MPa (475 psia) |
| 13 | Molecular Sieve Purification Reactor | Fixed bed | 6 | 0 | 39,463 kg/hr (87,000 lb/hr) 129°C (265°F) 2.8 MPa (401 psia) | 39,463 kg/hr (87,000 lb/hr) 129°C (265°F) 2.8 MPa (401 psia) |
| 14 | SNG Product Compressor | Centrifugal, Multi-staged | 3 | 1 | 1,240 $m^3$/min (43,800 scfm) 2.6 MPa (381 psia) 6.2 MPa (899 psia) | 1,240 $m^3$/min (43,800 scfm) 2.6 MPa (381 psia) 6.2 MPa (899 psia) |

**ACCOUNT 5B $CO_2$ COMPRESSION**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | $CO_2$ Compressor | Integrally geared, multi-stage centrifugal | 4 | 0 | N/A | 1,572 $m^3$/min @ 15.3 MPa (55,500 scfm @ 2,215 psia) |

**ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Combustion Turbine | N/A | N/A | N/A | N/A | N/A |
| 2 | Combustion Turbine Generator | N/A | N/A | N/A | N/A | N/A |

**ACCOUNT 7 DUCTING & STACK**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Stack | CS plate, type 409SS liner | 1 | 0 | 76 m (250 ft) high x 3.0 m (10 ft) diameter | 76 m (250 ft) high x 2.8 m (9 ft) diameter |

**ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Steam Turbine | Commercially available | 1 | 0 | 324 MW 12.4 MPa/566°C (1800 psig/ 1050°F) | 327 MW 12.4 MPa/566°C (1800 psig/ 1050°F) |
| 2 | Steam Turbine Generator | Hydrogen cooled, static excitation | 1 | 0 | 360 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase | 360 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase |
| 3 | Surface Condenser | Single pass, divided waterbox including vacuum pumps | 1 | 0 | 1,878 GJ/hr (1,780 MMBtu/hr), Condensing temperature 16°C (60°F), Ambient temperature 11°C (20°F) | 1,899 GJ/hr (1,800 MMBtu/hr), Condensing temperature 16°C (60°F), Ambient temperature 11°C (20°F) |
| 4 | Air-cooled Condenser | --- | N/A | N/A | N/A | N/A |

**ACCOUNT 9 COOLING WATER SYSTEM**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Circulating Water Pumps | Vertical, wet pit | 2 | 1 | 624,593 lpm @ 30 m (165,000 gpm @ 100 ft) | 620,808 lpm @ 30 m (164,000 gpm @ 100 ft) |
| 2 | Cooling Tower | Evaporative, mechanical draft, multi-cell | 1 | 0 | 11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 3,492 GJ/hr (3,310 MMBtu/hr) heat duty | 11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 3,461 GJ/hr (3,280 MMBtu/hr) heat duty |

## ACCOUNT 10 SLAG RECOVERY AND HANDLING

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Slag Quench Tank | Water bath | 6 | 0 | 151,416 liters (40,000 gal) | 151,416 liters (40,000 gal) |
| 2 | Slag Crusher | Roll | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 3 | Slag Depressurizer | Proprietary | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 4 | Slag Receiving Tank | Horizontal, weir | 6 | 0 | 90,850 liters (24,000 gal) | 90,850 liters (24,000 gal) |
| 5 | Black Water Overflow Tank | Shop fabricated | 6 | | 41,640 liters (11,000 gal) | 41,640 liters (11,000 gal) |
| 6 | Slag Conveyor | Drag chain | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 7 | Slag Separation Screen | Vibrating | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 8 | Coarse Slag Conveyor | Belt/bucket | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 9 | Fine Ash Settling Tank | Vertical, gravity | 6 | 0 | 132,489 liters (35,000 gal) | 132,489 liters (35,000 gal) |
| 10 | Fine Ash Recycle Pumps | Horizontal centrifugal | 6 | 3 | 38 lpm @ 14 m $H_2O$ (10 gpm @ 46 ft $H_2O$) | 38 lpm @ 14 m $H_2O$ (10 gpm @ 46 ft $H_2O$) |
| 11 | Grey Water Storage Tank | Field erected | 6 | 0 | 41,640 liters (11,000 gal) | 41,640 liters (11,000 gal) |
| 12 | Grey Water Pumps | Centrifugal | 6 | 3 | 151 lpm @ 433 m $H_2O$ (40 gpm @ 1,420 ft $H_2O$) | 151 lpm @ 433 m $H_2O$ (40 gpm @ 1,420 ft $H_2O$) |
| 13 | Slag Storage Bin | Vertical, field erected | 6 | 0 | 544 tonne (600 tons) | 544 tonne (600 tons) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 14 | Unloading Equipment | Telescoping chute | 2 | 0 | 100 tonne/hr (110 tph) | 100 tonne/hr (110 tph) |

**ACCOUNT 11 ACCESSORY ELECTRIC PLANT**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | STG Step-up Transformer | Oil-filled | 1 | 0 | 24 kV/345 kV, 360 MVA, 3-ph, 60 Hz | 24 kV/345 kV, 360 MVA, 3-ph, 60 Hz |
| 2 | Auxiliary Transformer | Oil-filled | 1 | 1 | 24 kV/4.16 kV, 239 MVA, 3-ph, 60 Hz | 24 kV/4.16 kV, 289 MVA, 3-ph, 60 Hz |
| 3 | Low Voltage Transformer | Dry ventilated | 1 | 1 | 4.16 kV/480 V, 36 MVA, 3-ph, 60 Hz | 4.16 kV/480 V, 43 MVA, 3-ph, 60 Hz |
| 4 | STG Isolated Phase Bus Duct and Tap Bus | Aluminum, self-cooled | 1 | 1 | 24 kV, 3-ph, 60 Hz | 24 kV, 3-ph, 60 Hz |
| 5 | Medium Voltage Switchgear | Metal clad | 1 | 1 | 4.16 kV, 3-ph, 60 Hz | 4.16 kV, 3-ph, 60 Hz |
| 6 | Low Voltage Switchgear | Metal enclosed | 1 | 0 | 480 V, 3-ph, 60 Hz | 480 V, 3-ph, 60 Hz |
| 7 | Emergency Diesel Generator | Sized for emergency shutdown | 1 | 0 | 750 kW, 480 V, 3-ph, 60 Hz | 750 kW, 480 V, 3-ph, 60 Hz |

## ACCOUNT 12 INSTRUMENTATION AND CONTROLS

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 1 Design Condition | Case 2 Design Condition |
|---|---|---|---|---|---|---|
| 1 | DCS - Main Control | Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W) | 1 | 0 | Operator stations/printers and engineering stations/printers | Operator stations/printers and engineering stations/printers |
| 2 | DCS - Processor | Microprocessor with redundant input/output | 1 | 0 | N/A | N/A |
| 3 | DCS - Data Highway | Fiber optic | 1 | 0 | Fully redundant, 25% spare | Fully redundant, 25% spare |

### 4.1.6 Cases 1 and 2 Cost Estimating

The cost estimating methodology was described previously in Section 2.8.

The TPC organized by cost account; owner's costs; TOC; and initial and annual O&M costs for the SNG plant without sequestration using Illinois No. 6 coal (Case 1) are shown in Exhibit 4-19 and Exhibit 4-20. The same data for the SNG plant with carbon sequestration using Illinois No. 6 coal (Case 2) are shown in Exhibit 4-21, and Exhibit 4-22.

The estimated TOC of the SNG plant without carbon sequestration is 3.24 billion dollars and with carbon sequestration is 3.31 billion dollars. Project and process contingencies represent 11.9 and 3.9 percent for the non-sequestration case and 11.9 and 3.8 percent for the sequestration case. The FYCOP is $19.27/MMBtu for the non-sequestration case and $20.95/MMBtu for the carbon sequestration case as shown in Exhibit ES-6.

# Exhibit 4-19 Case 1 Total Plant Cost Details

| | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | Client: | USDOE/NETL | | | | | | | | Report Date: | 2010-Feb-18 |
| | Project: | Cost & Performance Baseline for Fossil Energy Plants Vol.2 | | | | | | | | | |
| | | TOTAL PLANT COST SUMMARY | | | | | | | | | |
| | Case: | Case 1 - Siemens Quench SNG Production w/o CO2 | | | | | | | | | |
| | Plant Size: | 91.65 | MW,net | Estimate Type: | | Conceptual | | | Cost Base (Jun) | 2007 | ($x1000) |
| Acct | | Equipment | Material | Labor | | Sales | Bare Erected | Eng'g CM | Contingencies | | Total Plant Cost |
| No. | Item/Description | Cost | Cost | Direct | Indirect | Tax | Cost $ | H.O.& Fee | Process | Project | $ |
| 1 | COAL & SORBENT HANDLING | | | | | | | | | | |
| 1.1 | Coal Receive & Unload | $5,729 | $0 | $2,800 | $0 | $0 | $8,529 | $764 | $0 | $1,859 | $11,152 |
| 1.2 | Coal Stackout & Reclaim | $7,404 | $0 | $1,795 | $0 | $0 | $9,198 | $806 | $0 | $2,001 | $12,005 |
| 1.3 | Coal Conveyors & Yd Crush | $6,883 | $0 | $1,776 | $0 | $0 | $8,659 | $760 | $0 | $1,884 | $11,303 |
| 1.4 | Other Coal Handling | $1,801 | $0 | $411 | $0 | $0 | $2,212 | $194 | $0 | $481 | $2,886 |
| 1.5 | Sorbent Receive & Unload | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.6 | Sorbent Stackout & Reclaim | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.7 | Sorbent Conveyors | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.8 | Other Sorbent Handling | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.9 | Coal & Sorbent Hnd.Foundations | $0 | $4,054 | $10,138 | $0 | $0 | $14,192 | $1,360 | $0 | $3,110 | $18,662 |
| | **SUBTOTAL 1.** | **$21,817** | **$4,054** | **$16,919** | **$0** | **$0** | **$42,790** | **$3,884** | **$0** | **$9,335** | **$56,009** |
| 2 | COAL & SORBENT PREP & FEED | | | | | | | | | | |
| 2.1 | Coal Crushing & Drying | $92,942 | $0 | $13,543 | $0 | $0 | $106,485 | $9,671 | $0 | $23,231 | $139,387 |
| 2.2 | Prepared Coal Storage & Feed | $4,402 | $1,053 | $690 | $0 | $0 | $6,146 | $526 | $0 | $1,334 | $8,006 |
| 2.3 | Dry Coal Injection System | $144,878 | $1,681 | $13,455 | $0 | $0 | $160,014 | $13,782 | $0 | $34,759 | $208,555 |
| 2.4 | Misc.Coal Prep & Feed | $2,421 | $1,762 | $5,282 | $0 | $0 | $9,464 | $870 | $0 | $2,067 | $12,401 |
| 2.5 | Sorbent Prep Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.6 | Sorbent Storage & Feed | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.7 | Sorbent Injection System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.8 | Booster Air Supply System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.9 | Coal & Sorbent Feed Foundation | $0 | $9,409 | $7,725 | $0 | $0 | $17,135 | $1,587 | $0 | $3,744 | $22,466 |
| | **SUBTOTAL 2.** | **$244,643** | **$13,906** | **$40,695** | **$0** | **$0** | **$299,244** | **$26,435** | **$0** | **$65,136** | **$390,814** |
| 3 | FEEDWATER & MISC. BOP SYSTEMS | | | | | | | | | | |
| 3.1 | Feedwater System | $3,562 | $1,655 | $2,008 | $0 | $0 | $7,225 | $653 | $0 | $1,576 | $9,454 |
| 3.2 | Water Makeup & Pretreating | $832 | $87 | $465 | $0 | $0 | $1,385 | $132 | $0 | $455 | $1,971 |
| 3.3 | Other Feedwater Subsystems | $3,867 | $1,169 | $1,913 | $0 | $0 | $6,949 | $631 | $0 | $1,516 | $9,096 |
| 3.4 | Service Water Systems | $476 | $981 | $3,404 | $0 | $0 | $4,861 | $474 | $0 | $1,601 | $6,936 |
| 3.5 | Other Boiler Plant Systems | $2,556 | $991 | $2,455 | $0 | $0 | $6,002 | $569 | $0 | $1,314 | $7,885 |
| 3.6 | FO Supply Sys & Nat Gas | $102 | $210 | $188 | $0 | $0 | $500 | $48 | $0 | $110 | $658 |
| 3.7 | Waste Treatment Equipment | $1,164 | $0 | $710 | $0 | $0 | $1,873 | $182 | $0 | $617 | $2,673 |
| 3.8 | Misc. Power Plant Equipment | $1,721 | $230 | $884 | $0 | $0 | $2,836 | $274 | $0 | $933 | $4,042 |
| | **SUBTOTAL 3.** | **$14,281** | **$5,323** | **$12,026** | **$0** | **$0** | **$31,631** | **$2,964** | **$0** | **$8,121** | **$42,715** |
| 4 | GASIFIER & ACCESSORIES | | | | | | | | | | |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries | $325,844 | $0 | $150,858 | $0 | $0 | $476,702 | $42,340 | $71,505 | $88,582 | $679,129 |
| 4.2 | Syngas Cooling | w/4.1 | $0 | w/ 4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.3 | ASU/Oxidant Compression | $238,715 | $0 | w/equip. | $0 | $0 | $238,715 | $23,139 | $0 | $26,185 | $288,039 |
| 4.4 | LT Heat Recovery & FG Saturation | $49,590 | $0 | $18,374 | $0 | $0 | $67,963 | $6,510 | $0 | $14,895 | $89,369 |
| 4.5 | Misc. Gasification Equipment | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.6 | Flare Stack System | $0 | $2,333 | $950 | $0 | $0 | $3,283 | $315 | $0 | $719 | $4,317 |
| 4.7 | CO2 Solid Feed System Compressors | $10,520 | $2,525 | $3,787 | $0 | $0 | $16,832 | $1,683 | $0 | $3,703 | $22,218 |
| 4.8 | Major Component Rigging | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.9 | Gasification Foundations | $0 | $21,143 | $12,064 | $0 | $0 | $33,207 | $3,040 | $0 | $9,062 | $45,309 |
| | **SUBTOTAL 4.** | **$624,669** | **$26,001** | **$186,032** | **$0** | **$0** | **$836,702** | **$77,027** | **$71,505** | **$143,146** | **$1,128,380** |

## Exhibit 4-19 Case 1 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 5A | GAS CLEANUP & PIPING | | | | | | | | | | |
| 5A.1 | Double Stage Selexol | $125,892 | $0 | $106,823 | $0 | $0 | $232,715 | $22,506 | $46,543 | $60,353 | $362,116 |
| 5A.2 | Elemental Sulfur Plant | $16,274 | $3,243 | $20,996 | $0 | $0 | $40,513 | $3,935 | $0 | $8,890 | $53,338 |
| 5A.3 | Mercury Removal | $2,454 | $0 | $1,868 | $0 | $0 | $4,322 | $417 | $216 | $991 | $5,947 |
| 5A.4a | Shift Reactors | $10,461 | $0 | $4,211 | $0 | $0 | $14,672 | $1,407 | $0 | $3,216 | $19,294 |
| 5A.4b | COS Hydrolysis | $1,988 | $0 | $2,596 | $0 | $0 | $4,583 | $446 | $0 | $1,006 | $6,035 |
| 5A.5 | Methanation | $28,050 | $11,220 | $16,830 | $0 | $0 | $56,100 | $5,610 | $5,610 | $13,464 | $80,784 |
| 5A.6 | SNG Purification & Compression | $19,850 | $7,940 | $11,910 | $0 | $0 | $39,700 | $3,970 | $0 | $8,734 | $52,404 |
| 5A.7 | Fuel Gas Piping | $0 | $1,753 | $1,228 | $0 | $0 | $2,981 | $276 | $0 | $652 | $3,909 |
| 5A.9 | Process Interconnects | $0 | $12,000 | $18,000 | $0 | $0 | $30,000 | $3,000 | $0 | $6,600 | $39,600 |
| 5A.10 | HGCU Foundations | $0 | $1,776 | $1,145 | $0 | $0 | $2,920 | $268 | $0 | $957 | $4,145 |
| 5A.11 | Zinc Oxide Guard Bed | $793 | $0 | $146 | $0 | $0 | $939 | $94 | $0 | $207 | $1,239 |
| | **SUBTOTAL 5A.** | **$205,762** | **$37,932** | **$185,752** | **$0** | **$0** | **$429,446** | **$41,930** | **$52,369** | **$105,068** | **$628,812** |
| 5B | CO2 COMPRESSION | | | | | | | | | | |
| 5B.1 | CO2 Removal System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 5B.2 | CO2 Compression & Drying | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 5B.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 6 | COMBUSTION TURBINE/ACCESSORIES | | | | | | | | | | |
| 6.1 | Combustion Turbine Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.2 | Syngas Expander | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.3 | Compressed Air Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.9 | Combustion Turbine Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 6.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 7 | HRSG, DUCTING & STACK | | | | | | | | | | |
| 7.1 | Heat Recovery Steam Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.2 | HRSG Accessories | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $480 | $343 | $0 | $0 | $823 | $72 | $0 | $179 | $1,074 |
| 7.4 | Stack | $1,814 | $0 | $682 | $0 | $0 | $2,496 | $239 | $0 | $274 | $3,009 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $364 | $349 | $0 | $0 | $713 | $66 | $0 | $234 | $1,013 |
| | **SUBTOTAL 7.** | **$1,814** | **$844** | **$1,373** | **$0** | **$0** | **$4,032** | **$378** | **$0** | **$686** | **$5,096** |
| 8 | STEAM TURBINE GENERATOR | | | | | | | | | | |
| 8.1 | Steam TG & Accessories | $31,384 | $0 | $5,672 | $0 | $0 | $37,056 | $3,557 | $0 | $4,061 | $44,674 |
| 8.2 | Turbine Plant Auxiliaries | $178 | $0 | $520 | $0 | $0 | $698 | $68 | $0 | $77 | $843 |
| 8.3 | Condenser & Auxiliaries | $5,445 | $0 | $1,666 | $0 | $0 | $7,111 | $680 | $0 | $779 | $8,570 |
| 8.4 | Steam Piping | $5,365 | $0 | $3,774 | $0 | $0 | $9,139 | $785 | $0 | $2,481 | $12,405 |
| 8.9 | TG Foundations | $0 | $1,125 | $1,901 | $0 | $0 | $3,026 | $287 | $0 | $994 | $4,307 |
| | **SUBTOTAL 8.** | **$42,372** | **$1,125** | **$13,534** | **$0** | **$0** | **$57,031** | **$5,377** | **$0** | **$8,392** | **$70,800** |

## Exhibit 4-19 Case 1 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 9 | COOLING WATER SYSTEM | | | | | | | | | | |
| 9.1 | Cooling Towers | $8,833 | $0 | $1,607 | $0 | $0 | $10,440 | $994 | $0 | $1,715 | $13,149 |
| 9.2 | Circulating Water Pumps | $2,343 | $0 | $91 | $0 | $0 | $2,434 | $204 | $0 | $396 | $3,034 |
| 9.3 | Circ.Water System Auxiliaries | $186 | $0 | $27 | $0 | $0 | $213 | $20 | $0 | $35 | $268 |
| 9.4 | Circ.Water Piping | $0 | $7,768 | $2,014 | $0 | $0 | $9,782 | $884 | $0 | $2,133 | $12,800 |
| 9.5 | Make-up Water System | $440 | $0 | $629 | $0 | $0 | $1,069 | $103 | $0 | $234 | $1,406 |
| 9.6 | Component Cooling Water Sys | $917 | $1,097 | $780 | $0 | $0 | $2,794 | $262 | $0 | $611 | $3,667 |
| 9.9 | Circ.Water System Foundations | $0 | $2,850 | $4,844 | $0 | $0 | $7,694 | $729 | $0 | $2,527 | $10,951 |
| | **SUBTOTAL 9.** | **$12,719** | **$11,715** | **$9,992** | **$0** | **$0** | **$34,426** | **$3,197** | **$0** | **$7,652** | **$45,274** |
| 10 | ASH/SPENT SORBENT HANDLING SYS | | | | | | | | | | |
| 10.1 | Slag Dewatering & Cooling | $36,268 | $0 | $35,772 | $0 | $0 | $72,040 | $6,980 | $0 | $7,902 | $86,922 |
| 10.2 | Gasifier Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.3 | Cleanup Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.4 | High Temperature Ash Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.5 | Other Ash Recovery Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.6 | Ash Storage Silos | $818 | $0 | $890 | $0 | $0 | $1,708 | $166 | $0 | $281 | $2,154 |
| 10.7 | Ash Transport & Feed Equipment | $1,097 | $0 | $265 | $0 | $0 | $1,362 | $127 | $0 | $223 | $1,712 |
| 10.8 | Misc. Ash Handling Equipment | $1,694 | $2,076 | $620 | $0 | $0 | $4,390 | $418 | $0 | $721 | $5,529 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $72 | $91 | $0 | $0 | $163 | $15 | $0 | $54 | $232 |
| | **SUBTOTAL 10.** | **$39,877** | **$2,148** | **$37,637** | **$0** | **$0** | **$79,662** | **$7,706** | **$0** | **$9,181** | **$96,549** |
| 11 | ACCESSORY ELECTRIC PLANT | | | | | | | | | | |
| 11.1 | Generator Equipment | $446 | $0 | $441 | $0 | $0 | $887 | $85 | $0 | $97 | $1,069 |
| 11.2 | Station Service Equipment | $5,293 | $0 | $477 | $0 | $0 | $5,770 | $532 | $0 | $630 | $6,933 |
| 11.3 | Switchgear & Motor Control | $9,786 | $0 | $1,780 | $0 | $0 | $11,566 | $1,073 | $0 | $1,896 | $14,535 |
| 11.4 | Conduit & Cable Tray | $0 | $4,546 | $14,997 | $0 | $0 | $19,543 | $1,890 | $0 | $5,358 | $26,792 |
| 11.5 | Wire & Cable | $0 | $8,686 | $5,707 | $0 | $0 | $14,393 | $1,046 | $0 | $3,860 | $19,298 |
| 11.6 | Protective Equipment | $0 | $855 | $3,112 | $0 | $0 | $3,967 | $387 | $0 | $653 | $5,008 |
| 11.7 | Standby Equipment | $330 | $0 | $322 | $0 | $0 | $652 | $62 | $0 | $107 | $821 |
| 11.8 | Main Power Transformers | $8,795 | $0 | $182 | $0 | $0 | $8,977 | $682 | $0 | $1,449 | $11,108 |
| 11.9 | Electrical Foundations | $0 | $198 | $518 | $0 | $0 | $716 | $69 | $0 | $235 | $1,020 |
| | **SUBTOTAL 11.** | **$24,651** | **$14,285** | **$27,537** | **$0** | **$0** | **$66,472** | **$5,825** | **$0** | **$14,286** | **$86,583** |
| 12 | INSTRUMENTATION & CONTROL | | | | | | | | | | |
| 12.1 | IGCC Control Equipment | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.2 | Combustion Turbine Control | w/6.1 | $0 | w/6.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.3 | Steam Turbine Control | w/8.1 | $0 | w/8.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.4 | Other Major Component Control | $539 | $0 | $360 | $0 | $0 | $899 | $85 | $45 | $154 | $1,183 |
| 12.5 | Signal Processing Equipment | w/12.7 | $0 | w/12.7 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.6 | Control Boards,Panels & Racks | $358 | $0 | $229 | $0 | $0 | $587 | $56 | $29 | $134 | $806 |
| 12.7 | Computer & Accessories | $5,727 | $0 | $183 | $0 | $0 | $5,910 | $542 | $296 | $675 | $7,423 |
| 12.8 | Instrument Wiring & Tubing | $0 | $2,901 | $5,930 | $0 | $0 | $8,831 | $749 | $442 | $2,505 | $12,527 |
| 12.9 | Other I & C Equipment | $4,270 | $0 | $2,073 | $0 | $0 | $6,343 | $597 | $317 | $1,089 | $8,346 |
| | **SUBTOTAL 12.** | **$10,894** | **$2,901** | **$8,776** | **$0** | **$0** | **$22,571** | **$2,029** | **$1,129** | **$4,558** | **$30,286** |

# Exhibit 4-19 Case 1 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 13 | IMPROVEMENTS TO SITE | | | | | | | | | | |
| 13.1 | Site Preparation | $0 | $147 | $3,143 | $0 | $0 | $3,290 | $327 | $0 | $1,085 | $4,702 |
| 13.2 | Site Improvements | $0 | $2,615 | $3,476 | $0 | $0 | $6,091 | $601 | $0 | $2,008 | $8,700 |
| 13.3 | Site Facilities | $4,687 | $0 | $4,946 | $0 | $0 | $9,632 | $950 | $0 | $3,175 | $13,756 |
| | **SUBTOTAL 13.** | **$4,687** | **$2,762** | **$11,564** | **$0** | **$0** | **$19,013** | **$1,877** | **$0** | **$6,267** | **$27,158** |
| 14 | BUILDINGS & STRUCTURES | | | | | | | | | | |
| 14.1 | Combustion Turbine Area | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 14.2 | Steam Turbine Building | $0 | $2,691 | $3,834 | $0 | $0 | $6,525 | $600 | $0 | $1,069 | $8,194 |
| 14.3 | Administration Building | $0 | $870 | $630 | $0 | $0 | $1,500 | $134 | $0 | $245 | $1,879 |
| 14.4 | Circulation Water Pumphouse | $0 | $160 | $90 | $0 | $0 | $250 | $22 | $0 | $41 | $312 |
| 14.5 | Water Treatment Buildings | $0 | $696 | $679 | $0 | $0 | $1,375 | $124 | $0 | $225 | $1,724 |
| 14.6 | Machine Shop | $0 | $445 | $305 | $0 | $0 | $750 | $67 | $0 | $123 | $939 |
| 14.7 | Warehouse | $0 | $726 | $464 | $0 | $0 | $1,190 | $105 | $0 | $194 | $1,489 |
| 14.8 | Other Buildings & Structures | $0 | $528 | $411 | $0 | $0 | $940 | $84 | $0 | $205 | $1,229 |
| 14.9 | Waste Treating Building & Str. | $0 | $1,181 | $2,258 | $0 | $0 | $3,439 | $321 | $0 | $752 | $4,511 |
| | **SUBTOTAL 14.** | **$0** | **$7,297** | **$8,671** | **$0** | **$0** | **$15,968** | **$1,456** | **$0** | **$2,853** | **$20,277** |
| | | | | | | | | | | | |
| | **TOTAL COST** | **$1,248,184** | **$130,294** | **$560,509** | **$0** | **$0** | **$1,938,987** | **$180,084** | **$125,003** | **$384,680** | **$2,628,754** |
| | | | | | | | | | | | |
| | **Owner's Costs** | | | | | | | | | | |
| | **Preproduction Costs** | | | | | | | | | | |
| | 6 Months All Labor | | | | | | | | | | $21,523 |
| | 1 Month Maintenance Materials | | | | | | | | | | $3,796 |
| | 1 Month Non-fuel Consumables | | | | | | | | | | $816 |
| | 1 Month Waste Disposal | | | | | | | | | | $577 |
| | 25% of 1 Months Fuel Cost at 100% CF | | | | | | | | | | $3,362 |
| | 2% of TPC | | | | | | | | | | $52,575 |
| | **Total** | | | | | | | | | | **$82,649** |
| | **Inventory Capital** | | | | | | | | | | |
| | 60 day supply of fuel and consumables at 100% CF | | | | | | | | | | $28,184 |
| | 0.5% of TPC (spare parts) | | | | | | | | | | $13,144 |
| | **Total** | | | | | | | | | | **$41,328** |
| | | | | | | | | | | | |
| | **Initial Cost for Catalyst and Chemicals** | | | | | | | | | | $16,342 |
| | **Land** | | | | | | | | | | $900 |
| | **Other Owner's Costs** | | | | | | | | | | $394,313 |
| | **Financing Costs** | | | | | | | | | | $70,976 |
| | **Total Overnight Costs (TOC)** | | | | | | | | | | **$3,235,262** |
| | TASC Multiplier | | | | | | | | | | 1.201 |
| | **Total As-Spent Cost (TASC)** | | | | | | | | | | **$3,885,549** |

## Exhibit 4-20 Case 1 Initial and Annual O&M Costs

| | | | | | | |
|---|---|---|---|---|---|---|
| INITIAL & ANNUAL O&M EXPENSES | | | | | Cost Base (Jun): | 2007 |
| Case 1 - Siemens Quench SNG Production w/o CO2 | | | | | MWe-net: | 92 |
| SNG (MMbtu/hr): | 6892 | | | | Capacity Factor (%): | 90 |
| OPERATING & MAINTENANCE LABOR | | | | | | |
| Operating Labor | | | | | | |
| Operating Labor Rate(base): | 34.65 | $/hour | | | | |
| Operating Labor Burden: | 30.00 | % of base | | | | |
| Labor O-H Charge Rate: | 25.00 | % of labor | | | | |
| | | | Total | | | |
| Operating Labor Requirements(O.J.)per Shift: | 1 unit/mod. | | Plant | | | |
| Skilled Operator | 2.0 | | 2.0 | | | |
| Operator | 12.0 | | 12.0 | | | |
| Foreman | 1.0 | | 1.0 | | | |
| Lab Tech's, etc. | 3.0 | | 3.0 | | | |
| TOTAL-O.J.'s | 18.0 | | 18.0 | | | |
| | | | | | Annual Cost | Annual Unit Cost |
| | | | | | $ | $/MMBtu |
| Annual Operating Labor Cost | | | | | $7,102,696 | $0.131 |
| Maintenance Labor Cost | | | | | $27,333,527 | $0.503 |
| Administrative & Support Labor | | | | | $8,609,056 | $0.158 |
| Property Taxes and Insurance | | | | | $52,575,070 | $0.968 |
| **TOTAL FIXED OPERATING COSTS** | | | | | **$95,620,348** | **$1.760** |
| VARIABLE OPERATING COSTS | | | | | | |
| | | | | | | $/MMBtu |
| **Maintenance Material Cost** | | | | | **$41,000,290** | **$0.75453** |
| Consumables | Consumption | | Unit | Initial Fill | | |
| | Initial Fill | /Day | Cost | Cost | | |
| **Water(/1000 gallons)** | 0 | 5,162 | 1.08 | **$0** | **$1,834,102** | **$0.03375** |
| **Chemicals** | | | | | | |
| MU & WT Chem. (lb) | 0 | 30,752 | 0.17 | $0 | $1,748,328 | $0.03217 |
| Carbon (Mercury Removal) (lb) | 228,466 | 391 | 1.05 | $239,928 | $134,960 | $0.00248 |
| COS Catalyst (m3) | 116 | 0 | 2,397.36 | $277,521 | $78,053 | $0.00144 |
| Water Gas Shift Catalyst (ft3) | 5,515 | 4.72 | 498.83 | $2,750,848 | $773,676 | $0.01424 |
| ZnO Sorbent (ton) | 150 | 0.51 | 12,574.00 | $1,887,818 | $2,123,795 | $0.03908 |
| Methanation Catalyst (ft3) | 12,295 | 9.36 | 440.00 | $5,410,004 | $1,352,501 | $0.02489 |
| Selexol Solution (gal) | 431,115 | 137 | 13.40 | $5,776,182 | $602,843 | $0.01109 |
| SCR Catalyst (m3) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Aqueous Ammonia (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Claus Catalyst (ft3) | w/equip | 3.75 | 131.27 | $0 | $161,773 | $0.00298 |
| **Subtotal Chemicals** | | | | **$16,342,301** | **$6,975,929** | **$0.12838** |
| **Other** | | | | | | |
| Supplemental Fuel (MBtu) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for consumption) (MW | 0 | 0 | 61.60 | $0 | $0 | $0.00000 |
| Gases,N2 etc. (/100scf) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| L.P. Steam (/1000 pounds) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| **Subtotal Other** | | | | **$0** | **$0** | **$0.00000** |
| **Waste Disposal** | | | | | | |
| Spent Mercury Catalyst (lb.) | 0 | 391 | 0.42 | $0 | $53,599 | $0.00099 |
| Spent ZnO Sorbent (ton) | 0 | 0.51 | 16.23 | $0 | $2,740 | $0.00005 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Slag (ton) | 0 | 1,160 | 16.23 | $0 | $6,180,087 | $0.11373 |
| **Subtotal-Waste Disposal** | | | | **$0** | **$6,236,427** | **$0.11477** |
| **By-products & Emissions** | | | | | | |
| Sulfur (tons) | 0 | 288 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for sale) (MWh) | 0 | 2,200 | 58.00 | $0 | -$41,908,979 | -$0.77125 |
| **Subtotal By-Products** | | | | **$0** | **-$41,908,979** | **-$0.77125** |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$16,342,301** | **$14,137,770** | **$0.26018** |
| **Fuel (ton)** | 0 | 11,577 | 38.18 | **$0** | **$145,217,654** | **$2.67245** |

# Exhibit 4-21 Case 2 Total Plant Cost Details

| | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | **Client:** | USDOE/NETL | | | | | | | | **Report Date:** | 2010-Feb-18 |
| | **Project:** | Cost & Performance Baseline for Fossil Energy Plants Vol.2 | | | | | | | | | |
| | | | | | **TOTAL PLANT COST SUMMARY** | | | | | | |
| | **Case:** | Case 2 - Siemens Quench SNG Production w/ CO2 | | | | | | | | | |
| | **Plant Size:** | 48.51 | MW,net | **Estimate Type:** | | Conceptual | | | **Cost Base (Jun)** | 2007 | ($x1000) |

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 1 | COAL & SORBENT HANDLING | | | | | | | | | | |
| 1.1 | Coal Receive & Unload | $5,729 | $0 | $2,800 | $0 | $0 | $8,529 | $764 | $0 | $1,859 | $11,152 |
| 1.2 | Coal Stackout & Reclaim | $7,404 | $0 | $1,795 | $0 | $0 | $9,198 | $806 | $0 | $2,001 | $12,005 |
| 1.3 | Coal Conveyors & Yd Crush | $6,883 | $0 | $1,776 | $0 | $0 | $8,659 | $760 | $0 | $1,884 | $11,303 |
| 1.4 | Other Coal Handling | $1,801 | $0 | $411 | $0 | $0 | $2,212 | $194 | $0 | $481 | $2,886 |
| 1.5 | Sorbent Receive & Unload | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.6 | Sorbent Stackout & Reclaim | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.7 | Sorbent Conveyors | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.8 | Other Sorbent Handling | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.9 | Coal & Sorbent Hnd.Foundations | $0 | $4,054 | $10,138 | $0 | $0 | $14,192 | $1,360 | $0 | $3,110 | $18,662 |
| | **SUBTOTAL 1.** | **$21,817** | **$4,054** | **$16,919** | **$0** | **$0** | **$42,790** | **$3,884** | **$0** | **$9,335** | **$56,009** |
| 2 | COAL & SORBENT PREP & FEED | | | | | | | | | | |
| 2.1 | Coal Crushing & Drying | $92,942 | $0 | $13,543 | $0 | $0 | $106,485 | $9,671 | $0 | $23,231 | $139,387 |
| 2.2 | Prepared Coal Storage & Feed | $4,402 | $1,053 | $690 | $0 | $0 | $6,146 | $526 | $0 | $1,334 | $8,006 |
| 2.3 | Dry Coal Injection System | $144,878 | $1,681 | $13,455 | $0 | $0 | $160,014 | $13,782 | $0 | $34,759 | $208,555 |
| 2.4 | Misc.Coal Prep & Feed | $2,421 | $1,762 | $5,282 | $0 | $0 | $9,464 | $870 | $0 | $2,067 | $12,401 |
| 2.5 | Sorbent Prep Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.6 | Sorbent Storage & Feed | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.7 | Sorbent Injection System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.8 | Booster Air Supply System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.9 | Coal & Sorbent Feed Foundation | $0 | $9,409 | $7,725 | $0 | $0 | $17,135 | $1,587 | $0 | $3,744 | $22,466 |
| | **SUBTOTAL 2.** | **$244,643** | **$13,906** | **$40,695** | **$0** | **$0** | **$299,244** | **$26,435** | **$0** | **$65,136** | **$390,814** |
| 3 | FEEDWATER & MISC. BOP SYSTEMS | | | | | | | | | | |
| 3.1 | FeedwaterSystem | $3,582 | $1,664 | $2,019 | $0 | $0 | $7,265 | $657 | $0 | $1,584 | $9,506 |
| 3.2 | Water Makeup & Pretreating | $829 | $87 | $463 | $0 | $0 | $1,378 | $131 | $0 | $453 | $1,962 |
| 3.3 | Other Feedwater Subsystems | $3,889 | $1,176 | $1,923 | $0 | $0 | $6,988 | $634 | $0 | $1,524 | $9,146 |
| 3.4 | Service Water Systems | $474 | $976 | $3,389 | $0 | $0 | $4,839 | $472 | $0 | $1,593 | $6,905 |
| 3.5 | Other Boiler Plant Systems | $2,545 | $986 | $2,444 | $0 | $0 | $5,974 | $567 | $0 | $1,308 | $7,849 |
| 3.6 | FO Supply Sys & Nat Gas | $102 | $210 | $188 | $0 | $0 | $500 | $48 | $0 | $110 | $658 |
| 3.7 | Waste Treatment Equipment | $1,158 | $0 | $707 | $0 | $0 | $1,865 | $182 | $0 | $614 | $2,660 |
| 3.8 | Misc. Power Plant Equipment | $1,721 | $230 | $884 | $0 | $0 | $2,836 | $274 | $0 | $933 | $4,042 |
| | **SUBTOTAL 3.** | **$14,299** | **$5,330** | **$12,016** | **$0** | **$0** | **$31,645** | **$2,965** | **$0** | **$8,120** | **$42,729** |
| 4 | GASIFIER & ACCESSORIES | | | | | | | | | | |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries | $325,819 | $0 | $150,846 | $0 | $0 | $476,665 | $42,337 | $71,500 | $88,575 | $679,077 |
| 4.2 | Syngas Cooling | w/4.1 | $0 | w/ 4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.3 | ASU/Oxidant Compression | $238,715 | $0 | w/equip. | $0 | $0 | $238,715 | $23,139 | $0 | $26,185 | $288,039 |
| 4.4 | LT Heat Recovery & FG Saturation | $49,540 | $0 | $18,356 | $0 | $0 | $67,896 | $6,504 | $0 | $14,880 | $89,280 |
| 4.5 | Misc. Gasification Equipment | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.6 | Flare Stack System | $0 | $2,332 | $949 | $0 | $0 | $3,281 | $315 | $0 | $719 | $4,315 |
| 4.7 | CO2 Solid Feed System Compressors | $6,696 | $1,607 | $2,411 | $0 | $0 | $10,713 | $1,071 | $0 | $2,357 | $14,142 |
| 4.8 | Major Component Rigging | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.9 | Gasification Foundations | $0 | $21,143 | $12,064 | $0 | $0 | $33,207 | $3,040 | $0 | $9,062 | $45,309 |
| | **SUBTOTAL 4.** | **$620,771** | **$25,082** | **$184,626** | **$0** | **$0** | **$830,479** | **$76,405** | **$71,500** | **$141,779** | **$1,120,162** |

## Exhibit 4-21 Case 2 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 5A | GAS CLEANUP & PIPING | | | | | | | | | | |
| 5A.1 | Double Stage Selexol | $125,787 | $0 | $106,734 | $0 | $0 | $232,521 | $22,487 | $46,504 | $60,302 | $361,815 |
| 5A.2 | Elemental Sulfur Plant | $16,277 | $3,244 | $21,000 | $0 | $0 | $40,521 | $3,936 | $0 | $8,891 | $53,349 |
| 5A.3 | Mercury Removal | $2,453 | $0 | $1,867 | $0 | $0 | $4,320 | $417 | $216 | $991 | $5,943 |
| 5A.4a | Shift Reactors | $10,448 | $0 | $4,205 | $0 | $0 | $14,653 | $1,405 | $0 | $3,212 | $19,269 |
| 5A.4b | COS Hydrolysis | $1,999 | $0 | $2,611 | $0 | $0 | $4,610 | $448 | $0 | $1,012 | $6,070 |
| 5A.5 | Methanation | $28,050 | $11,220 | $16,830 | $0 | $0 | $56,100 | $5,610 | $5,610 | $13,464 | $80,784 |
| 5A.6 | SNG Purification & Compression | $19,850 | $7,940 | $11,910 | $0 | $0 | $39,700 | $3,970 | $0 | $8,734 | $52,404 |
| 5A.7 | Fuel Gas Piping | $0 | $1,752 | $1,227 | $0 | $0 | $2,979 | $276 | $0 | $651 | $3,907 |
| 5A.9 | Process Interconnects | $0 | $12,000 | $18,000 | $0 | $0 | $30,000 | $3,000 | $0 | $6,600 | $39,600 |
| 5A.10 | HGCU Foundations | $0 | $1,774 | $1,144 | $0 | $0 | $2,918 | $268 | $0 | $956 | $4,143 |
| 5A.11 | Zinc Oxide Guard Bed | $792 | $0 | $146 | $0 | $0 | $938 | $94 | $0 | $206 | $1,239 |
| | **SUBTOTAL 5A.** | **$205,656** | **$37,931** | **$185,674** | **$0** | **$0** | **$429,261** | **$41,912** | **$52,330** | **$105,019** | **$628,522** |
| 5B | CO2 COMPRESSION | | | | | | | | | | |
| 5B.1 | CO2 Removal System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 5B.2 | CO2 Compression & Drying | $34,256 | $0 | $20,173 | $0 | $0 | $54,428 | $5,240 | $0 | $11,934 | $71,602 |
| | **SUBTOTAL 5B.** | **$34,256** | **$0** | **$20,173** | **$0** | **$0** | **$54,428** | **$5,240** | **$0** | **$11,934** | **$71,602** |
| 6 | COMBUSTION TURBINE/ACCESSORIES | | | | | | | | | | |
| 6.1 | Combustion Turbine Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.2 | Syngas Expander | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.3 | Compressed Air Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.9 | Combustion Turbine Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 6** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 7 | HRSG, DUCTING & STACK | | | | | | | | | | |
| 7.1 | Heat Recovery Steam Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.2 | HRSG Accessories | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.4 | Stack | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 7.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 8 | STEAM TURBINE GENERATOR | | | | | | | | | | |
| 8.1 | Steam TG & Accessories | $31,587 | $0 | $5,484 | $0 | $0 | $37,071 | $3,557 | $0 | $4,063 | $44,691 |
| 8.2 | Turbine Plant Auxiliaries | $173 | $0 | $504 | $0 | $0 | $677 | $66 | $0 | $74 | $817 |
| 8.3 | Condenser & Auxiliaries | $5,488 | $0 | $1,616 | $0 | $0 | $7,104 | $679 | $0 | $778 | $8,561 |
| 8.4 | Steam Piping | $5,591 | $0 | $3,933 | $0 | $0 | $9,525 | $818 | $0 | $2,586 | $12,929 |
| 8.9 | TG Foundations | $0 | $1,091 | $1,844 | $0 | $0 | $2,934 | $278 | $0 | $964 | $4,176 |
| | **SUBTOTAL 8.** | **$42,839** | **$1,091** | **$13,381** | **$0** | **$0** | **$57,310** | **$5,399** | **$0** | **$8,465** | **$71,174** |

## Exhibit 4-21 Case 2 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 9 | COOLING WATER SYSTEM | | | | | | | | | | |
| 9.1 | Cooling Towers | $8,777 | $0 | $1,597 | $0 | $0 | $10,373 | $988 | $0 | $1,704 | $13,065 |
| 9.2 | Circulating Water Pumps | $2,333 | $0 | $88 | $0 | $0 | $2,421 | $203 | $0 | $394 | $3,018 |
| 9.3 | Circ.Water System Auxiliaries | $185 | $0 | $26 | $0 | $0 | $212 | $20 | $0 | $35 | $267 |
| 9.4 | Circ.Water Piping | $0 | $7,740 | $2,007 | $0 | $0 | $9,747 | $881 | $0 | $2,125 | $12,753 |
| 9.5 | Make-up Water System | $438 | $0 | $627 | $0 | $0 | $1,065 | $102 | $0 | $233 | $1,401 |
| 9.6 | Component Cooling Water Sys | $914 | $1,093 | $778 | $0 | $0 | $2,784 | $261 | $0 | $609 | $3,654 |
| 9.9 | Circ.Water System Foundations | $0 | $2,834 | $4,818 | $0 | $0 | $7,652 | $725 | $0 | $2,513 | $10,891 |
| | **SUBTOTAL 9.** | **$12,647** | **$11,667** | **$9,940** | **$0** | **$0** | **$34,254** | **$3,181** | **$0** | **$7,614** | **$45,048** |
| 10 | ASH/SPENT SORBENT HANDLING SYS | | | | | | | | | | |
| 10.1 | Slag Dewatering & Cooling | $36,268 | $0 | $35,772 | $0 | $0 | $72,040 | $6,980 | $0 | $7,902 | $86,922 |
| 10.2 | Gasifier Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.3 | Cleanup Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.4 | High Temperature Ash Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.5 | Other Ash Recovery Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.6 | Ash Storage Silos | $818 | $0 | $890 | $0 | $0 | $1,708 | $166 | $0 | $281 | $2,154 |
| 10.7 | Ash Transport & Feed Equipment | $1,097 | $0 | $265 | $0 | $0 | $1,362 | $127 | $0 | $223 | $1,712 |
| 10.8 | Misc. Ash Handling Equipment | $1,694 | $2,076 | $620 | $0 | $0 | $4,390 | $418 | $0 | $721 | $5,529 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $72 | $91 | $0 | $0 | $163 | $15 | $0 | $54 | $232 |
| | **SUBTOTAL 10.** | **$39,877** | **$2,148** | **$37,637** | **$0** | **$0** | **$79,662** | **$7,706** | **$0** | **$9,181** | **$96,549** |
| 11 | ACCESSORY ELECTRIC PLANT | | | | | | | | | | |
| 11.1 | Generator Equipment | $435 | $0 | $431 | $0 | $0 | $866 | $83 | $0 | $95 | $1,043 |
| 11.2 | Station Service Equipment | $5,740 | $0 | $517 | $0 | $0 | $6,258 | $577 | $0 | $683 | $7,518 |
| 11.3 | Switchgear & Motor Control | $10,612 | $0 | $1,930 | $0 | $0 | $12,542 | $1,163 | $0 | $2,056 | $15,761 |
| 11.4 | Conduit & Cable Tray | $0 | $4,930 | $16,263 | $0 | $0 | $21,193 | $2,050 | $0 | $5,811 | $29,053 |
| 11.5 | Wire & Cable | $0 | $9,419 | $6,189 | $0 | $0 | $15,608 | $1,134 | $0 | $4,185 | $20,927 |
| 11.6 | Protective Equipment | $0 | $855 | $3,112 | $0 | $0 | $3,967 | $387 | $0 | $653 | $5,008 |
| 11.7 | Standby Equipment | $330 | $0 | $322 | $0 | $0 | $652 | $62 | $0 | $107 | $821 |
| 11.8 | Main Power Transformers | $9,427 | $0 | $182 | $0 | $0 | $9,609 | $729 | $0 | $1,551 | $11,889 |
| 11.9 | Electrical Foundations | $0 | $198 | $518 | $0 | $0 | $716 | $69 | $0 | $235 | $1,020 |
| | **SUBTOTAL 11.** | **$26,544** | **$15,402** | **$29,464** | **$0** | **$0** | **$71,410** | **$6,254** | **$0** | **$15,377** | **$93,041** |
| 12 | INSTRUMENTATION & CONTROL | | | | | | | | | | |
| 12.1 | IGCC Control Equipment | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.2 | Combustion Turbine Control | w/6.1 | $0 | w/6.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.3 | Steam Turbine Control | w/8.1 | $0 | w/8.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.4 | Other Major Component Control | $536 | $0 | $358 | $0 | $0 | $894 | $85 | $45 | $153 | $1,177 |
| 12.5 | Signal Processing Equipment | w/12.7 | $0 | w/12.7 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.6 | Control Boards,Panels & Racks | $356 | $0 | $228 | $0 | $0 | $584 | $55 | $29 | $134 | $802 |
| 12.7 | Computer & Accessories | $5,696 | $0 | $182 | $0 | $0 | $5,879 | $540 | $294 | $671 | $7,383 |
| 12.8 | Instrument Wiring & Tubing | $0 | $2,885 | $5,899 | $0 | $0 | $8,784 | $745 | $439 | $2,492 | $12,460 |
| 12.9 | Other I & C Equipment | $4,247 | $0 | $2,062 | $0 | $0 | $6,309 | $594 | $315 | $1,083 | $8,301 |
| | **SUBTOTAL 12.** | **$10,835** | **$2,885** | **$8,729** | **$0** | **$0** | **$22,450** | **$2,018** | **$1,123** | **$4,533** | **$30,124** |

## Exhibit 4-21 Case 2 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 13 | IMPROVEMENTS TO SITE | | | | | | | | | | |
| 13.1 | Site Preparation | $0 | $147 | $3,143 | $0 | $0 | $3,290 | $327 | $0 | $1,085 | $4,702 |
| 13.2 | Site Improvements | $0 | $2,615 | $3,476 | $0 | $0 | $6,091 | $601 | $0 | $2,008 | $8,700 |
| 13.3 | Site Facilities | $4,687 | $0 | $4,946 | $0 | $0 | $9,632 | $950 | $0 | $3,175 | $13,756 |
| | **SUBTOTAL 13.** | **$4,687** | **$2,762** | **$11,564** | **$0** | **$0** | **$19,013** | **$1,877** | **$0** | **$6,267** | **$27,158** |
| 14 | BUILDINGS & STRUCTURES | | | | | | | | | | |
| 14.1 | Combustion Turbine Area | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 14.2 | Steam Turbine Building | $0 | $2,624 | $3,738 | $0 | $0 | $6,361 | $585 | $0 | $1,042 | $7,989 |
| 14.3 | Administration Building | $0 | $870 | $630 | $0 | $0 | $1,500 | $134 | $0 | $245 | $1,879 |
| 14.4 | Circulation Water Pumphouse | $0 | $160 | $90 | $0 | $0 | $250 | $22 | $0 | $41 | $312 |
| 14.5 | Water Treatment Buildings | $0 | $693 | $676 | $0 | $0 | $1,368 | $124 | $0 | $224 | $1,716 |
| 14.6 | Machine Shop | $0 | $445 | $305 | $0 | $0 | $750 | $67 | $0 | $123 | $939 |
| 14.7 | Warehouse | $0 | $726 | $464 | $0 | $0 | $1,190 | $105 | $0 | $194 | $1,489 |
| 14.8 | Other Buildings & Structures | $0 | $528 | $411 | $0 | $0 | $940 | $84 | $0 | $205 | $1,229 |
| 14.9 | Waste Treating Building & Str. | $0 | $1,181 | $2,258 | $0 | $0 | $3,439 | $321 | $0 | $752 | $4,511 |
| | **SUBTOTAL 14.** | **$0** | **$7,227** | **$8,572** | **$0** | **$0** | **$15,798** | **$1,441** | **$0** | **$2,825** | **$20,064** |
| | | | | | | | | | | | |
| | **TOTAL COST** | **$1,278,871** | **$129,485** | **$579,389** | **$0** | **$0** | **$1,987,745** | **$184,716** | **$124,952** | **$395,583** | **$2,692,997** |
| | | | | | | | | | | | |
| | **Owner's Costs** | | | | | | | | | | |
| | **Preproduction Costs** | | | | | | | | | | |
| | 6 Months All Labor | | | | | | | | | | $21,750 |
| | 1 Month Maintenance Materials | | | | | | | | | | $3,847 |
| | 1 Month Non-fuel Consumables | | | | | | | | | | $813 |
| | 1 Month Waste Disposal | | | | | | | | | | $577 |
| | 25% of 1 Months Fuel Cost at 100% CF | | | | | | | | | | $3,362 |
| | 2% of TPC | | | | | | | | | | $53,860 |
| | **Total** | | | | | | | | | | **$84,209** |
| | **Inventory Capital** | | | | | | | | | | |
| | 60 day supply of fuel and consumables at 100% CF | | | | | | | | | | $28,181 |
| | 0.5% of TPC (spare parts) | | | | | | | | | | $13,465 |
| | **Total** | | | | | | | | | | **$41,646** |
| | | | | | | | | | | | |
| | **Initial Cost for Catalyst and Chemicals** | | | | | | | | | | $16,327 |
| | **Land** | | | | | | | | | | $900 |
| | **Other Owner's Costs** | | | | | | | | | | $403,950 |
| | **Financing Costs** | | | | | | | | | | $72,711 |
| | **Total Overnight Costs (TOC)** | | | | | | | | | | **$3,312,740** |
| | TASC Multiplier | | | | | | | | | | 1.201 |
| | **Total As-Spent Cost (TASC)** | | | | | | | | | | **$3,978,601** |

# Exhibit 4-22 Case 2 Initial and Annual O&M Costs

| INITIAL & ANNUAL O&M EXPENSES | | | Cost Base (Jun): | 2007 |
|---|---|---|---|---|
| Case 2 - Siemens Quench SNG Production w/ CO2 | | | MWe-net: | 49 |
| SNG (MMbtu/hr): | 6886 | | Capacity Factor (%): | 90 |

OPERATING & MAINTENANCE LABOR

Operating Labor

| | | |
|---|---|---|
| Operating Labor Rate(base): | 34.65 | $/hour |
| Operating Labor Burden: | 30.00 | % of base |
| Labor O-H Charge Rate: | 25.00 | % of labor |

| Operating Labor Requirements(O.J.)per Shift: | 1 unit/mod. | Total Plant |
|---|---|---|
| Skilled Operator | 2.0 | 2.0 |
| Operator | 12.0 | 12.0 |
| Foreman | 1.0 | 1.0 |
| Lab Tech's, etc. | 3.0 | 3.0 |
| TOTAL-O.J.'s | 18.0 | 18.0 |

| | Annual Cost $ | Annual Unit Cost $/MMBtu |
|---|---|---|
| Annual Operating Labor Cost | $7,102,696 | $0.131 |
| Maintenance Labor Cost | $27,697,722 | $0.510 |
| Administrative & Support Labor | $8,700,105 | $0.160 |
| Property Taxes and Insurance | $53,859,942 | $0.992 |
| **TOTAL FIXED OPERATING COSTS** | **$97,360,465** | **$1.793** |

VARIABLE OPERATING COSTS

| | | $/MMBtu |
|---|---|---|
| **Maintenance Material Cost** | **$41,546,584** | **$0.76523** |

| Consumables | Consumption Initial Fill | Consumption /Day | Unit Cost | Initial Fill Cost | | |
|---|---|---|---|---|---|---|
| **Water(/1000 gallons)** | 0 | 5,129 | 1.08 | **$0** | **$1,822,333** | **$0.03356** |
| **Chemicals** | | | | | | |
| MU & WT Chem.(lb) | 0 | 30,554 | 0.17 | $0 | $1,737,110 | $0.03200 |
| Carbon (Mercury Removal) (lb) | 228,257 | 391 | 1.05 | $239,709 | $134,836 | $0.00248 |
| COS Catalyst (m3) | 117 | 0 | 2,397.36 | $279,597 | $78,637 | $0.00145 |
| Water Gas Shift Catalyst (ft3) | 5,505 | 4.71 | 498.83 | $2,746,203 | $772,369 | $0.01423 |
| ZnO Sorbent (ton) | 150 | 0.51 | 12,574.00 | $1,885,197 | $2,120,847 | $0.03906 |
| Methanation Catalyst (ft3) | 12,289 | 9.35 | 440.00 | $5,407,162 | $1,351,791 | $0.02490 |
| Selexol Solution (gal) | 430,618 | 137 | 13.40 | $5,769,523 | $602,148 | $0.01109 |
| SCR Catalyst (m3) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Aqueous Ammonia (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Claus Catalyst (ft3) | w/equip | 3.75 | 131.27 | $0 | $161,817 | $0.00298 |
| **Subtotal Chemicals** | | | | **$16,327,392** | **$6,959,555** | **$0.12819** |
| **Other** | | | | | | |
| Supplemental Fuel (MBtu) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for consumption) (MW | 0 | 0 | 61.60 | $0 | $0 | $0.00000 |
| Gases,N2 etc.(/100scf) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| L.P. Steam(/1000 pounds) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| **Subtotal Other** | | | | **$0** | **$0** | **$0.00000** |
| **Waste Disposal** | | | | | | |
| Spent Mercury Catalyst (lb.) | 0 | 391 | 0.42 | $0 | $53,551 | $0.00099 |
| Spent ZnO Sorbent (ton) | 0 | 0.51 | 16.23 | $0 | $2,737 | $0.00005 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Slag (ton) | 0 | 1,160 | 16.23 | $0 | $6,180,087 | $0.11383 |
| **Subtotal-Waste Disposal** | | | | **$0** | **$6,236,374** | **$0.11487** |
| **By-products & Emissions** | | | | | | |
| Sulfur (tons) | 0 | 288 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for sale) (MWh) | 0 | 1,164 | 58.00 | $0 | -$22,182,265 | -$0.40857 |
| **Subtotal By-Products** | | | | **$0** | **-$22,182,265** | **-$0.40857** |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$16,327,392** | **$34,382,582** | **$0.63328** |
| **Fuel (ton)** | 0 | 11,577 | 38.18 | **$0** | **$145,217,654** | **$2.67471** |

## 4.2 SYNTHETIC NATURAL GAS AND AMMONIA CO-PRODUCTION USING ILLINOIS NO. 6 COAL

### 4.2.1 Process Description

In this section the overall Siemens gasification process for producing SNG and ammonia production is described. All major process systems for coal handling and drying, gasifier, raw gas cooling/particulate removal, syngas scrubbing, and SNG production are the same as previously described for Cases 1 and 2. Only the major process areas for ammonia production and their subsequent integration with the SNG portion of the plant are described in this section. The process BFDs for the Illinois No. 6 coal non-sequestration and sequestration co-production cases are shown in Exhibit 4-23 and Exhibit 4-25. The associated stream tables are shown in Exhibit 4-24 for the non-sequestration case and Exhibit 4-26 for the carbon sequestration case.

**Sour Gas Shift for Ammonia Production**

Two SGS units are employed to maximize the hydrogen production necessary for ammonia synthesis and to reduce the amount of COS entering the AGR system. A $H_2O$ to CO ratio of 3.0 is maintained to drive the shift reaction to an approximate fractional conversion of 99 percent for CO.

Following the water scrubber, approximately 24 percent of the total raw gas stream is utilized for ammonia production. This raw gas stream (stream 30) is combined with shift steam (stream 31) in the first reactor. The stream exiting the first reactor is cooled to 232°C (450°F) to help further drive the shift reaction in the second reactor. The shifted syngas exits the second reactor with nearly all the available carbon monoxide shifted to hydrogen.

**Mercury Removal and Acid Gas Removal**

Before the raw gas can be treated in the AGR process, it must be cooled to about 35°C (95°F). During this cooling sequence through a series of heat exchangers, most of the water vapor condenses. This water, which contains some $NH_3$, is sent to the sour water stripper. The cooled syngas (stream 32) then passes through a carbon bed to remove 95 percent of the Hg (Section 3.1.6).

The AGR process is a two stage Selexol process where $H_2S$ is removed in the first stage and $CO_2$ is removed in the second stage of absorption. The process results in three product streams. A high hydrogen syngas (stream 34) is combined with the purge gas from the ammonia synthesis loop (stream 40) and sent to the PSA process. A $CO_2$-rich stream is combined with the $CO_2$-rich stream from the SNG production trains and either vented to the atmosphere in the non-sequestration case or compressed for sequestration in the carbon sequestration case. A portion of the $CO_2$ product stream is withdrawn and further compressed to 5.62 MPa (815 psia) to provide the fluid for coal transport (stream 12). The acid gas stream (stream 43) is combined with its counterpart in the SNG production train (stream 42) and fed to the Claus plant (stream 44). In both cases, the acid gas contains nearly 40 percent $H_2S$ and nearly 50 percent $CO_2$, with the balance primarily $H_2$ and CO.

**Pressure Swing Absorption**

The PSA unit is used to purify the syngas stream entering the ammonia synthesis loop to acceptable levels of contaminants for the ammonia catalyst as described in Section 3.1.10.

The high hydrogen syngas from the Selexol unit (stream 34) is combined with the ammonia purge gas (stream 40) and feed to the PSA unit at 3.24 MPa (470 psia) and 21°C (69°F). The PSA unit produces a 99 vol% $H_2$ stream for ammonia synthesis (stream 37). Purge nitrogen (stream 35) enters the PSA beds from the ASU to increase the hydrogen recovery to 94 vol%. The tail gas from the PSA unit (stream 36) is sent to a waste heat recovery boiler.

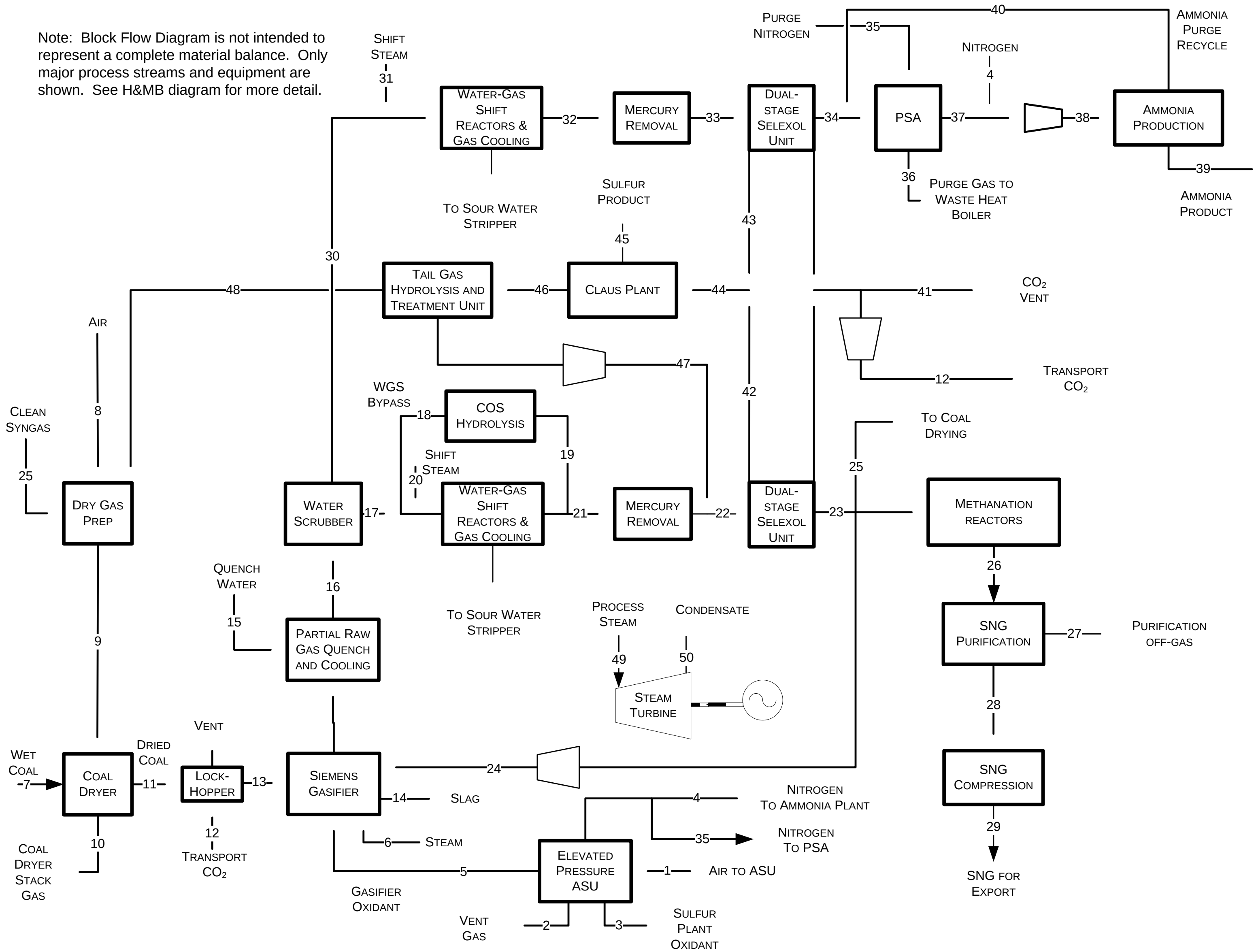
**Ammonia Synthesis Loop**

The purified $H_2$ stream from the PSA unit is combined with additional nitrogen from the ASU (stream 4) to maintain a 3:1 molar ratio of $H_2$ to $N_2$ entering the ammonia reactor. The combined stream is compressed to 13.62 MPa (1,975 psia) before being fed into the synthesis loop (stream 38). The ammonia reactor converts $H_2$ and $N_2$ into $NH_3$ at a pressure of 14.0 MPa (2,030 psia) due to further compression within the synthesis loop. The ammonia product is refrigerated to −25°C (−13°F) and sent to storage vessels (stream 39). Approximately five percent of the maximum flow in the ammonia synthesis loop is treated in the PSA unit (stream 40) to keep concentrations of inerts such as $CH_4$, Ar, and $O_2$ from building up in the synthesis loop. The ammonia synthesis loop produces anhydrous ammonia with a purity of 99.5 wt%.

**$CO_2$ Compression, Claus Unit, Power Block**

The process parameters are the same as previously discussed in Section 4.1.1.

# Exhibit 4-23 Case 3 Block Flow Diagram, Bituminous Coal to SNG and Ammonia without Carbon Sequestration



# Exhibit 4-24 Case 3 Stream Table, Bituminous Coal to SNG and Ammonia without Carbon Sequestration

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | | |
| Ar | 0.0092 | 0.0101 | 0.0101 | 0.0000 | 0.0101 | 0.0000 | 0.0000 | 0.0092 | 0.0084 | 0.0064 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0020 | 0.0020 | 0.0020 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0081 | 0.0010 | 0.0000 | 0.0000 | 0.4205 | 0.4364 | 0.4364 |
| $CO_2$ | 0.0003 | 0.0005 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0003 | 0.1192 | 0.0916 | 0.0000 | 0.9781 | 0.1235 | 0.0000 | 0.0000 | 0.0399 | 0.0415 | 0.0415 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0005 | 0.0006 | 0.0006 |
| $H_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0121 | 0.6937 | 0.0000 | 0.0000 | 0.1971 | 0.2046 | 0.2046 |
| $H_2O$ | 0.0099 | 0.0044 | 0.0000 | 0.0000 | 0.0000 | 1.0000 | 0.0000 | 0.0099 | 0.1710 | 0.3627 | 0.0000 | 0.0014 | 0.0979 | 0.0000 | 1.0000 | 0.3295 | 0.3042 | 0.3042 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0058 | 0.0060 | 0.0060 |
| $N_2$ | 0.7732 | 0.9765 | 0.0004 | 1.0000 | 0.0004 | 0.0000 | 0.0000 | 0.7732 | 0.6234 | 0.4793 | 0.0000 | 0.0001 | 0.0171 | 0.0000 | 0.0000 | 0.0046 | 0.0047 | 0.0047 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.2074 | 0.0085 | 0.9895 | 0.0000 | 0.9895 | 0.0000 | 0.0000 | 0.2074 | 0.0780 | 0.0600 | 0.0000 | 0.0000 | 0.0668 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 51,682 | 36,441 | 151 | 3,127 | 10,369 | 3,442 | 0 | 3,538 | 4,396 | 5,718 | 0 | 3,559 | 14,094 | 0 | 14,055 | 53,960 | 39,309 | 5,856 |
| V-L Flowrate (kg/hr) | 1,491,391 | 1,025,161 | 4,845 | 87,594 | 332,612 | 62,017 | 0 | 102,084 | 125,819 | 149,636 | 0 | 154,233 | 158,471 | 0 | 253,210 | 1,095,758 | 801,642 | 119,422 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 437,263 | 0 | 0 | 0 | 413,446 | 0 | 332,092 | 43,794 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 15 | 25 | 32 | 72 | 32 | 343 | 15 | 15 | 1,260 | 143 | 71 | 83 | 18 | 1,500 | 216 | 260 | 214 | 214 |
| Pressure (MPa, abs) | 0.10 | 0.11 | 0.86 | 3.24 | 0.86 | 5.10 | 0.10 | 0.10 | 0.10 | 0.10 | 0.10 | 5.62 | 5.62 | 4.24 | 8.274 | 3.89 | 3.76 | 3.76 |
| Enthalpy (kJ/kg) | 30.93 | 32.59 | 26.94 | 69.15 | 26.94 | 3,177.81 | --- | 30.93 | 1,853.88 | 826.48 | --- | 31.02 | 4,689.64 | --- | 1,017.873 | 1,152.07 | 1,014.68 | 1,014.68 |
| Density ($kg/m^3$) | 1.2 | 1.3 | 11.0 | 31.6 | 11.0 | 20.1 | --- | 1.2 | 0.2 | 0.8 | --- | 99.3 | 28.4 | --- | 782.0 | 18.0 | 19.1 | 19.1 |
| V-L Molecular Weight | 28.857 | 28.132 | 32.078 | 28.013 | 32.078 | 18.015 | --- | 28.857 | 28.621 | 26.169 | --- | 43.334 | 11.244 | --- | 18.015 | 20.307 | 20.393 | 20.393 |
| | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 113,940 | 80,339 | 333 | 6,894 | 22,859 | 7,589 | 0 | 7,799 | 9,692 | 12,606 | 0 | 7,847 | 31,073 | 0 | 30,987 | 118,962 | 86,662 | 12,910 |
| V-L Flowrate (lb/hr) | 3,287,954 | 2,260,094 | 10,681 | 193,111 | 733,283 | 136,724 | 0 | 225,056 | 277,383 | 329,890 | 0 | 340,027 | 349,368 | 0 | 558,232 | 2,415,734 | 1,767,318 | 263,281 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 964,000 | 0 | 0 | 0 | 911,493 | 0 | 732,139 | 96,550 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 59 | 76 | 90 | 162 | 90 | 650 | 59 | 59 | 2,300 | 290 | 160 | 181 | 65 | 2,732 | 420 | 500 | 418 | 418 |
| Pressure (psia) | 14.7 | 16.4 | 125.0 | 470.0 | 125.0 | 740.0 | 14.7 | 14.7 | 14.7 | 14.7 | 14.4 | 815.0 | 815.0 | 614.7 | 1,200.0 | 564.7 | 544.7 | 544.7 |
| Enthalpy (Btu/lb) | 13.3 | 14.0 | 11.6 | 29.7 | 11.6 | 1,366.2 | --- | 13.3 | 797.0 | 355.3 | --- | 13.3 | 2,016.2 | --- | 437.6 | 495.3 | 436.2 | 436.2 |
| Density ($lb/ft^3$) | 0.076 | 0.080 | 0.685 | 1.973 | 0.685 | 1.257 | --- | 0.076 | 0.014 | 0.048 | --- | 6.197 | 1.775 | --- | 48.817 | 1.125 | 1.195 | 1.195 |

A - Reference conditions are 32.02 F & 0.089 PSIA

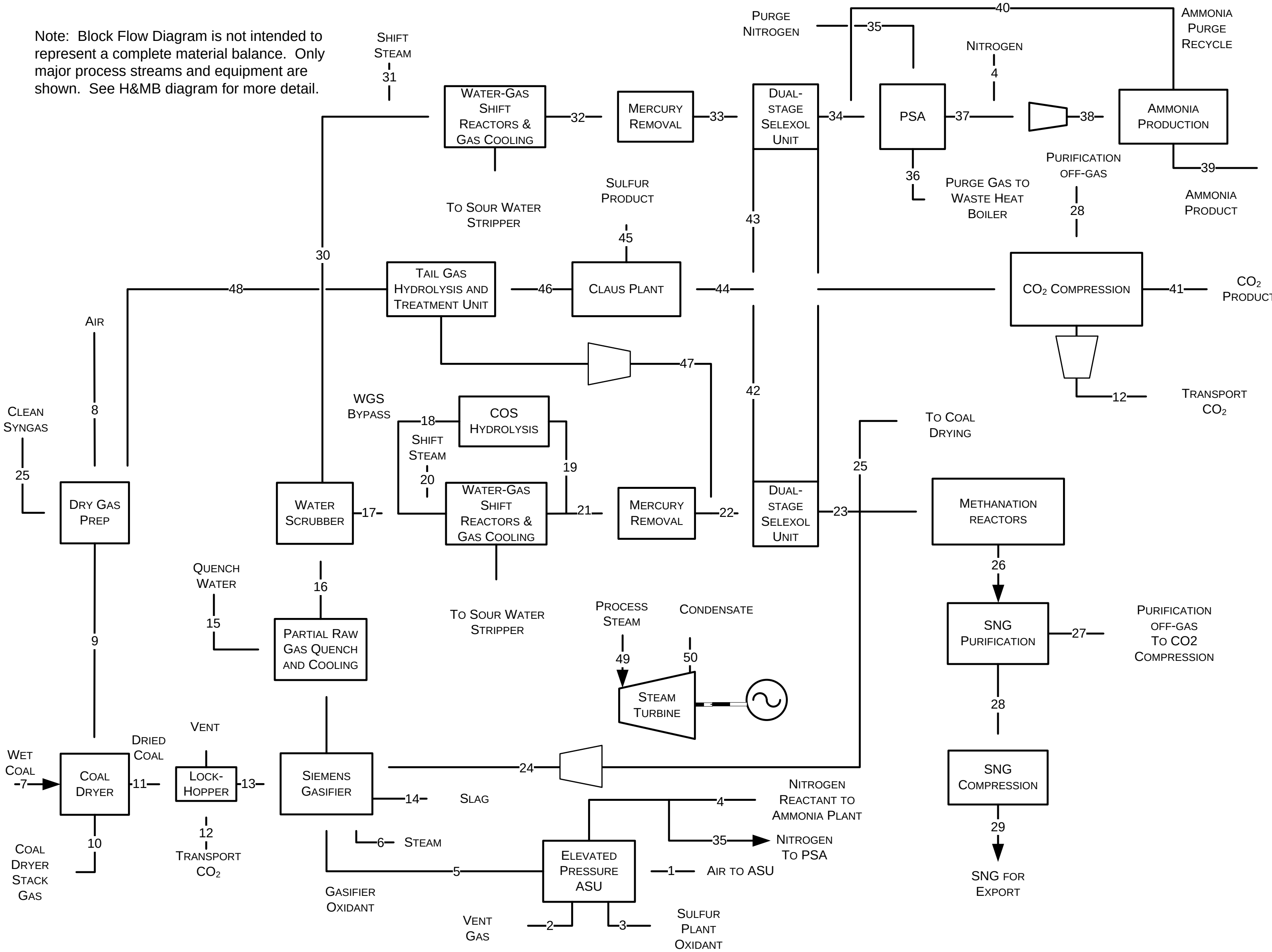
## Exhibit 4-24 Case 3 Stream Table, Bituminous Coal to SNG and Ammonia without Carbon Sequestration (continued)

| | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | 34 | 35 | 36 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | | |
| Ar | 0.0020 | 0.0000 | 0.0021 | 0.0021 | 0.0030 | 0.0030 | 0.0030 | 0.0105 | 0.0000 | 0.0120 | 0.0120 | 0.0020 | 0.0000 | 0.0018 | 0.0018 | 0.0029 | 0.0000 | 0.0072 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0001 | 0.0001 | 0.8302 | 0.0000 | 0.9496 | 0.9496 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0002 |
| CO | 0.4364 | 0.0000 | 0.1647 | 0.1642 | 0.2372 | 0.2372 | 0.2372 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.4364 | 0.0000 | 0.0037 | 0.0037 | 0.0061 | 0.0000 | 0.0166 |
| $CO_2$ | 0.0420 | 0.0000 | 0.3264 | 0.3284 | 0.0366 | 0.0366 | 0.0366 | 0.1253 | 0.9816 | 0.0021 | 0.0021 | 0.0415 | 0.0000 | 0.4192 | 0.4192 | 0.0530 | 0.0000 | 0.1443 |
| COS | 0.0000 | 0.0000 | 0.0001 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0006 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.2046 | 0.0000 | 0.4935 | 0.4919 | 0.7160 | 0.7160 | 0.7160 | 0.0070 | 0.0000 | 0.0080 | 0.0080 | 0.2046 | 0.0000 | 0.5636 | 0.5636 | 0.9309 | 0.0000 | 0.1939 |
| $H_2O$ | 0.3036 | 1.0000 | 0.0018 | 0.0018 | 0.0001 | 0.0001 | 0.0001 | 0.0023 | 0.0184 | 0.0000 | 0.0000 | 0.3042 | 1.0000 | 0.0017 | 0.0017 | 0.0001 | 0.0000 | 0.0004 |
| $H_2S$ | 0.0066 | 0.0000 | 0.0066 | 0.0067 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0060 | 0.0000 | 0.0057 | 0.0057 | 0.0000 | 0.0000 | 0.0000 |
| $N_2$ | 0.0047 | 0.0000 | 0.0049 | 0.0048 | 0.0071 | 0.0071 | 0.0071 | 0.0247 | 0.0000 | 0.0282 | 0.0282 | 0.0047 | 0.0000 | 0.0042 | 0.0042 | 0.0069 | 1.0000 | 0.6195 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0180 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 5,856 | 10,935 | 38,282 | 38,406 | 26,217 | 115 | 645 | 7,292 | 917 | 6,375 | 6,375 | 12,685 | 12,748 | 14,332 | 14,332 | 8,622 | 1,245 | 3,164 |
| V-L Flowrate (kg/hr) | 119,422 | 196,998 | 783,015 | 788,417 | 262,604 | 1,151 | 6,461 | 145,836 | 39,914 | 105,923 | 105,923 | 258,689 | 229,661 | 288,176 | 288,176 | 40,467 | 34,881 | 79,621 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 215 | 288 | 35 | 35 | 31 | 80 | 31 | 35 | 35 | 35 | 35 | 214 | 288 | 35 | 35 | 31 | 72 | 21 |
| Pressure (MPa, abs) | 3.69 | 4.14 | 3.31 | 3.24 | 3.24 | 5.10 | 3.24 | 2.73 | 0.14 | 2.627 | 6.202 | 3.756 | 4.137 | 3.37 | 3.3 | 3.238 | 3.241 | 0.138 |
| Enthalpy (kJ/kg) | 1,014.21 | 3,070.04 | 44.97 | 44.66 | 86.06 | 234.17 | 86.06 | 41.84 | 48.20 | 47.750 | 13.125 | 1,014.683 | 3,070.036 | 44.67 | 44.2 | 190.221 | 69.152 | 24.935 |
| Density ($kg/m^3$) | 18.8 | 18.2 | 26.8 | 26.4 | 12.7 | 17.0 | 12.7 | 22.6 | 2.4 | 17.9 | 44.7 | 19.1 | 18.2 | 27.1 | 26.5 | 5.9 | 31.6 | 1.4 |
| V-L Molecular Weight | 20.393 | 18.015 | 20.454 | 20.529 | 10.017 | 10.017 | 10.017 | 20.000 | 43.532 | 16.615 | 16.615 | 20.393 | 18.015 | 20.107 | 20.107 | 4.693 | 28.013 | 25.164 |
| | | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 12,910 | 24,108 | 84,396 | 84,670 | 57,797 | 253 | 1,422 | 16,076 | 2,021 | 14,055 | 14,055 | 27,966 | 28,105 | 31,597 | 31,597 | 19,009 | 2,745 | 6,976 |
| V-L Flowrate (lb/hr) | 263,281 | 434,306 | 1,726,252 | 1,738,163 | 578,942 | 2,538 | 14,243 | 321,514 | 87,994 | 233,520 | 233,520 | 570,311 | 506,315 | 635,320 | 635,320 | 89,214 | 76,900 | 175,534 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 418 | 550 | 95 | 95 | 87 | 177 | 87 | 95 | 95 | 95 | 95 | 418 | 550 | 95 | 94 | 87 | 162 | 69 |
| Pressure (psia) | 534.7 | 600.0 | 479.7 | 469.7 | 469.6 | 740.0 | 469.6 | 396.1 | 20.1 | 381.0 | 899.5 | 544.7 | 600.0 | 489.2 | 479.2 | 469.6 | 470.0 | 20.0 |
| Enthalpy (Btu/lb) | 436.0 | 1,319.9 | 19.3 | 19.2 | 37.0 | 100.7 | 37.0 | 18.0 | 20.7 | 20.5 | 5.6 | 436.2 | 1,319.9 | 19.2 | 19.0 | 81.8 | 29.7 | 10.7 |
| Density ($lb/ft^3$) | 1.172 | 1.135 | 1.675 | 1.647 | 0.792 | 1.063 | 0.792 | 1.409 | 0.148 | 1.116 | 2.789 | 1.195 | 1.135 | 1.690 | 1.657 | 0.371 | 1.973 | 0.089 |

## Exhibit 4-24 Case 3 Stream Table, Bituminous Coal to SNG and Ammonia without Carbon Sequestration (continued)

| | 37 | 38 | 39 | 40 | 41 | 42 | 43 | 44 | 45 | 46 | 47 | 48 | 49 | 50 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | |
| Ar | 0.0013 | 0.0010 | 0.0005 | 0.0035 | 0.0000 | 0.0005 | 0.0005 | 0.0005 | 0.0000 | 0.0019 | 0.0000 | 0.0034 | 0.0000 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0021 | 0.0480 | 0.0011 | 0.0350 | 0.0000 | 0.1014 | 0.0000 | 0.0060 | 0.0000 | 0.0000 |
| $CO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.9906 | 0.4323 | 0.5377 | 0.4614 | 0.0000 | 0.3385 | 0.9629 | 0.5672 | 0.0000 | 0.0000 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0017 | 0.0001 | 0.0012 | 0.0000 | 0.0005 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.9905 | 0.7491 | 0.0030 | 0.7328 | 0.0030 | 0.0917 | 0.1024 | 0.0946 | 0.0000 | 0.0589 | 0.0005 | 0.2694 | 0.0000 | 0.0000 |
| $H_2O$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0041 | 0.0329 | 0.0314 | 0.0325 | 0.0000 | 0.4950 | 0.0059 | 0.1525 | 1.0000 | 1.0000 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.3923 | 0.3264 | 0.3741 | 0.0000 | 0.0021 | 0.0306 | 0.0000 | 0.0000 | 0.0000 |
| $N_2$ | 0.0082 | 0.2499 | 0.0011 | 0.2448 | 0.0000 | 0.0006 | 0.0005 | 0.0006 | 0.0000 | 0.0008 | 0.0000 | 0.0014 | 0.0000 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.9954 | 0.0190 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0009 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 9,704 | 12,831 | 4,898 | 3,001 | 13,431 | 657 | 252 | 909 | 0 | 1,058 | 124 | 604 | 40,998 | 76,214 |
| V-L Flowrate (kg/hr) | 22,125 | 109,719 | 83,324 | 26,398 | 587,552 | 22,772 | 8,966 | 31,738 | 0 | 28,599 | 5,403 | 17,274 | 738,585 | 1,373,008 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10,893 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | |
| Temperature (°C) | 21 | 21 | -25 | -28 | 5 | 48 | 48 | 48 | 177 | 232 | 35 | 51 | 566 | 39 |
| Pressure (MPa, abs) | 3.238 | 13.614 | 13.048 | 3.238 | 0.115 | 0.163 | 0.163 | 0.163 | 0.119 | 0.085 | 3.272 | 0.103 | 12.512 | 0.827 |
| Enthalpy (kJ/kg) | 267.348 | 68.637 | -1,408.059 | -99.171 | 7.544 | 91.640 | 87.876 | 90.577 | --- | 1,168.962 | -1.583 | 232.751 | 3,513.235 | 164.154 |
| Density ($kg/m^3$) | 3.0 | 44.6 | 668.8 | 13.9 | 2.2 | 2.1 | 2.2 | 2.1 | 5,281.2 | 0.5 | 67.9 | 1.2 | 34.9 | 993.0 |
| V-L Molecular Weight | 2.280 | 8.551 | 17.010 | 8.796 | 43.745 | 34.656 | 35.628 | 34.925 | --- | 27.035 | 43.530 | 28.598 | 18.015 | 18.015 |
| | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 21,394 | 28,288 | 10,799 | 6,616 | 29,611 | 1,449 | 555 | 2,003 | 0 | 2,332 | 274 | 1,332 | 90,384 | 168,022 |
| V-L Flowrate (lb/hr) | 48,777 | 241,888 | 183,698 | 58,197 | 1,295,330 | 50,203 | 19,766 | 69,969 | 0 | 63,051 | 11,911 | 38,083 | 1,628,301 | 3,026,965 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 24,016 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | |
| Temperature (°F) | 69 | 70 | -13 | -19 | 42 | 119 | 119 | 119 | 351 | 450 | 95 | 123 | 1,050 | 102 |
| Pressure (psia) | 469.6 | 1,974.5 | 1,892.5 | 469.6 | 16.7 | 23.7 | 23.7 | 23.7 | 17.3 | 12.3 | 474.5 | 14.9 | 1,814.7 | 120.0 |
| Enthalpy (Btu/lb) | 114.9 | 29.5 | -605.4 | -42.6 | 3.2 | 39.4 | 37.8 | 38.9 | --- | 502.6 | -0.7 | 100.1 | 1,510.4 | 70.6 |
| Density ($lb/ft^3$) | 0.186 | 2.787 | 41.749 | 0.865 | 0.137 | 0.133 | 0.137 | 0.134 | 329.697 | 0.034 | 4.237 | 0.072 | 2.177 | 61.989 |

## Exhibit 4-25  Case 4 Block Flow Diagram, Bituminous Coal to SNG and Ammonia with Carbon Sequestration



# Exhibit 4-26 Case 4 Stream Table, Bituminous Coal to SNG and Ammonia with Carbon Sequestration

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | |
| Ar | 0.0092 | 0.0101 | 0.0101 | 0.0000 | 0.0101 | 0.0000 | 0.0000 | 0.0092 | 0.0084 | 0.0064 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0020 | 0.0020 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0032 | 0.0004 | 0.0000 | 0.0000 | 0.4206 | 0.4365 |
| $CO_2$ | 0.0003 | 0.0005 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0003 | 0.1204 | 0.0924 | 0.0000 | 0.9918 | 0.1242 | 0.0000 | 0.0000 | 0.0400 | 0.0415 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0005 | 0.0006 |
| $H_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0047 | 0.6936 | 0.0000 | 0.0000 | 0.1970 | 0.2044 |
| $H_2O$ | 0.0099 | 0.0044 | 0.0000 | 0.0000 | 0.0000 | 1.0000 | 0.0000 | 0.0099 | 0.1711 | 0.3635 | 0.0000 | 0.0001 | 0.0978 | 0.0000 | 1.0000 | 0.3295 | 0.3042 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0058 | 0.0060 |
| $N_2$ | 0.7732 | 0.9765 | 0.0004 | 1.0000 | 0.0004 | 0.0000 | 0.0000 | 0.7732 | 0.6227 | 0.4782 | 0.0000 | 0.0000 | 0.0172 | 0.0000 | 0.0000 | 0.0046 | 0.0047 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.2074 | 0.0085 | 0.9895 | 0.0000 | 0.9895 | 0.0000 | 0.0000 | 0.2074 | 0.0774 | 0.0595 | 0.0000 | 0.0000 | 0.0669 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 51,694 | 36,451 | 156 | 3,127 | 10,366 | 3,442 | 0 | 3,516 | 4,374 | 5,696 | 0 | 3,525 | 14,077 | 0 | 14,050 | 53,938 | 39,291 |
| V-L Flowrate (kg/hr) | 1,491,725 | 1,025,437 | 4,997 | 87,591 | 332,535 | 62,017 | 0 | 101,453 | 125,255 | 149,072 | 0 | 154,233 | 158,471 | 0 | 253,117 | 1,095,588 | 801,497 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 437,263 | 0 | 0 | 0 | 413,446 | 0 | 332,092 | 43,794 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 15 | 25 | 32 | 117 | 32 | 343 | 15 | 15 | 1,265 | 143 | 71 | 82 | 18 | 1,500 | 216 | 260 | 214 |
| Pressure (MPa, abs) | 0.10 | 0.11 | 0.86 | 3.24 | 0.86 | 5.10 | 0.10 | 0.10 | 0.10 | 0.10 | 0.10 | 5.62 | 5.62 | 4.24 | 8.274 | 3.89 | 3.76 |
| Enthalpy (kJ/kg) | 30.93 | 32.59 | 26.94 | 117.12 | 26.94 | 3,177.81 | --- | 30.93 | 1,860.19 | 827.90 | --- | 27.40 | 4,694.27 | --- | 1,017.873 | 1,151.80 | 1,014.38 |
| Density ($kg/m^3$) | 1.2 | 1.3 | 11.0 | 27.9 | 11.0 | 20.1 | --- | 1.2 | 0.2 | 0.8 | --- | 101.4 | 28.5 | --- | 782.0 | 18.0 | 19.2 |
| V-L Molecular Weight | 28.857 | 28.132 | 32.078 | 28.013 | 32.078 | 18.015 | --- | 28.857 | 28.637 | 26.172 | --- | 43.759 | 11.257 | --- | 18.015 | 20.312 | 20.399 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 113,965 | 80,361 | 343 | 6,893 | 22,854 | 7,589 | 0 | 7,751 | 9,643 | 12,557 | 0 | 7,770 | 31,035 | 0 | 30,975 | 118,912 | 86,623 |
| V-L Flowrate (lb/hr) | 3,288,691 | 2,260,701 | 11,017 | 193,105 | 733,114 | 136,724 | 0 | 223,666 | 276,140 | 328,647 | 0 | 340,026 | 349,368 | 0 | 558,027 | 2,415,359 | 1,766,999 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 964,000 | 0 | 0 | 0 | 911,493 | 0 | 732,138 | 96,550 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 59 | 76 | 90 | 242 | 90 | 650 | 59 | 59 | 2,308 | 290 | 160 | 179 | 65 | 2,732 | 420 | 500 | 418 |
| Pressure (psia) | 14.7 | 16.4 | 125.0 | 470.0 | 125.0 | 740.0 | 14.7 | 14.7 | 14.7 | 14.7 | 14.4 | 815.0 | 815.0 | 614.7 | 1,200.0 | 564.7 | 544.7 |
| Enthalpy (Btu/lb) | 13.3 | 14.0 | 11.6 | 50.4 | 11.6 | 1,366.2 | --- | 13.3 | 799.7 | 355.9 | --- | 11.8 | 2,018.2 | --- | 437.6 | 495.2 | 436.1 |
| Density ($lb/ft^3$) | 0.076 | 0.080 | 0.685 | 1.739 | 0.685 | 1.257 | --- | 0.076 | 0.014 | 0.048 | --- | 6.328 | 1.779 | --- | 48.817 | 1.125 | 1.196 |

A - Reference conditions are 32.02 F & 0.089 PSIA

## Exhibit 4-26 Case 4 Stream Table, Bituminous Coal to SNG and Ammonia with Carbon Sequestration (continued)

| | 18 | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | 34 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | |
| Ar | 0.0020 | 0.0020 | 0.0000 | 0.0021 | 0.0021 | 0.0030 | 0.0030 | 0.0030 | 0.0105 | 0.0000 | 0.0120 | 0.0120 | 0.0020 | 0.0000 | 0.0018 | 0.0018 | 0.0029 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0001 | 0.0001 | 0.8300 | 0.0000 | 0.9496 | 0.9496 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 |
| CO | 0.4365 | 0.4365 | 0.0000 | 0.1647 | 0.1642 | 0.2372 | 0.2372 | 0.2372 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.4365 | 0.0000 | 0.0037 | 0.0037 | 0.0061 |
| $CO_2$ | 0.0415 | 0.0421 | 0.0000 | 0.3265 | 0.3285 | 0.0366 | 0.0366 | 0.0366 | 0.1255 | 0.9817 | 0.0022 | 0.0022 | 0.0415 | 0.0000 | 0.4193 | 0.4193 | 0.0530 |
| COS | 0.0006 | 0.0000 | 0.0000 | 0.0001 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0006 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.2044 | 0.2044 | 0.0000 | 0.4933 | 0.4917 | 0.7160 | 0.7160 | 0.7160 | 0.0070 | 0.0000 | 0.0080 | 0.0080 | 0.2044 | 0.0000 | 0.5634 | 0.5634 | 0.9309 |
| $H_2O$ | 0.3042 | 0.3036 | 1.0000 | 0.0018 | 0.0018 | 0.0001 | 0.0001 | 0.0001 | 0.0023 | 0.0183 | 0.0000 | 0.0000 | 0.3042 | 1.0000 | 0.0017 | 0.0017 | 0.0001 |
| $H_2S$ | 0.0060 | 0.0066 | 0.0000 | 0.0066 | 0.0068 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0060 | 0.0000 | 0.0057 | 0.0057 | 0.0000 |
| $N_2$ | 0.0047 | 0.0047 | 0.0000 | 0.0049 | 0.0048 | 0.0071 | 0.0071 | 0.0071 | 0.0247 | 0.0000 | 0.0282 | 0.0282 | 0.0047 | 0.0000 | 0.0042 | 0.0042 | 0.0069 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 5,852 | 5,852 | 10,939 | 38,269 | 38,397 | 26,203 | 115 | 654 | 7,287 | 918 | 6,369 | 6,369 | 12,679 | 12,748 | 14,327 | 14,327 | 8,618 |
| V-L Flowrate (kg/hr) | 119,381 | 119,381 | 197,069 | 782,967 | 788,515 | 262,524 | 1,151 | 6,555 | 145,785 | 39,955 | 105,827 | 105,827 | 258,642 | 229,652 | 288,150 | 288,150 | 40,455 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 214 | 215 | 288 | 35 | 35 | 31 | 80 | 31 | 35 | 35 | 35 | 35 | 214 | 288 | 35 | 35 | 31 |
| Pressure (MPa, abs) | 3.76 | 3.69 | 4.14 | 3.31 | 3.24 | 3.24 | 5.10 | 3.24 | 2.73 | 0.14 | 2.627 | 6.202 | 3.756 | 4.137 | 3.37 | 3.3 | 3.238 |
| Enthalpy (kJ/kg) | 1,014.38 | 1,013.91 | 2,955.16 | 44.96 | 44.63 | 86.04 | 234.12 | 86.04 | 41.82 | 48.17 | 47.750 | 13.124 | 1,014.380 | 2,955.157 | 44.65 | 44.2 | 190.183 |
| Density (kg/m$^3$) | 19.2 | 18.8 | 18.2 | 26.8 | 26.4 | 12.7 | 17.0 | 12.7 | 22.6 | 2.4 | 17.9 | 44.7 | 19.2 | 18.2 | 27.1 | 26.6 | 5.9 |
| V-L Molecular Weight | 20.399 | 20.399 | 18.015 | 20.460 | 20.536 | 10.019 | 10.019 | 10.019 | 20.006 | 43.533 | 16.615 | 16.615 | 20.399 | 18.015 | 20.112 | 20 | 4.694 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 12,902 | 12,902 | 24,116 | 84,369 | 84,652 | 57,768 | 253 | 1,442 | 16,065 | 2,023 | 14,042 | 14,042 | 27,953 | 28,104 | 31,587 | 31,587 | 18,999 |
| V-L Flowrate (lb/hr) | 263,189 | 263,189 | 434,462 | 1,726,146 | 1,738,377 | 578,766 | 2,538 | 14,451 | 321,400 | 88,085 | 233,309 | 233,309 | 570,209 | 506,295 | 635,261 | 635,261 | 89,187 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 418 | 418 | 550 | 95 | 95 | 87 | 177 | 87 | 95 | 95 | 95 | 95 | 418 | 550 | 95 | 94 | 87 |
| Pressure (psia) | 544.7 | 534.7 | 600.0 | 479.7 | 469.7 | 469.6 | 740.0 | 469.6 | 396.1 | 20.1 | 381.0 | 899.5 | 544.7 | 600.0 | 489.2 | 479.2 | 469.6 |
| Enthalpy (Btu/lb) | 436.1 | 435.9 | 1,270.5 | 19.3 | 19.2 | 37.0 | 100.7 | 37.0 | 18.0 | 20.7 | 20.5 | 5.6 | 436.1 | 1,270.5 | 19.2 | 19.0 | 81.8 |
| Density (lb/ft$^3$) | 1.196 | 1.173 | 1.135 | 1.675 | 1.648 | 0.792 | 1.063 | 0.792 | 1.410 | 0.148 | 1.116 | 2.789 | 1.196 | 1.135 | 1.690 | 1.658 | 0.371 |

# Exhibit 4-26 Case 4 Stream Table, Bituminous Coal to SNG and Ammonia with Carbon Sequestration (continued)

| | 35 | 36 | 37 | 38 | 39 | 40 | 41 | 42 | 43 | 44 | 45 | 46 | 47 | 48 | 49 | 50 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | |
| Ar | 0.0000 | 0.0072 | 0.0013 | 0.0010 | 0.0005 | 0.0035 | 0.0000 | 0.0005 | 0.0005 | 0.0005 | 0.0000 | 0.0020 | 0.0000 | 0.0035 | 0.0000 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0002 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.0000 | 0.0166 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0032 | 0.0477 | 0.0011 | 0.0349 | 0.0000 | 0.1032 | 0.0000 | 0.0057 | 0.0000 | 0.0000 |
| $CO_2$ | 0.0000 | 0.1443 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.9918 | 0.4296 | 0.5376 | 0.4594 | 0.0000 | 0.3331 | 0.9316 | 0.5770 | 0.0000 | 0.0000 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0017 | 0.0001 | 0.0012 | 0.0000 | 0.0004 | 0.0001 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.0000 | 0.1938 | 0.9905 | 0.7490 | 0.0030 | 0.7324 | 0.0047 | 0.0910 | 0.1023 | 0.0941 | 0.0000 | 0.0588 | 0.0005 | 0.2585 | 0.0000 | 0.0000 |
| $H_2O$ | 0.0000 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0327 | 0.0314 | 0.0323 | 0.0000 | 0.4947 | 0.0059 | 0.1538 | 1.0000 | 1.0000 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.3962 | 0.3265 | 0.3770 | 0.0000 | 0.0020 | 0.0619 | 0.0001 | 0.0000 | 0.0000 |
| $N_2$ | 1.0000 | 0.6196 | 0.0082 | 0.2500 | 0.0011 | 0.2451 | 0.0000 | 0.0006 | 0.0005 | 0.0006 | 0.0000 | 0.0008 | 0.0000 | 0.0014 | 0.0000 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0180 | 0.0000 | 0.0000 | 0.9954 | 0.0190 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0049 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 1,245 | 3,164 | 9,699 | 12,825 | 4,897 | 3,000 | 16,002 | 661 | 252 | 913 | 0 | 1,066 | 128 | 594 | 40,943 | 76,182 |
| V-L Flowrate (kg/hr) | 34,865 | 79,621 | 22,112 | 109,703 | 83,295 | 26,414 | 700,235 | 22,915 | 8,966 | 31,881 | 0 | 28,895 | 5,548 | 17,247 | 737,593 | 1,372,447 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10,895 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | |
| Temperature (°C) | 117 | 25 | 25 | 21 | -25 | -28 | 35 | 48 | 48 | 48 | 177 | 232 | 35 | 51 | 566 | 39 |
| Pressure (MPa, abs) | 3.241 | 0.138 | 3.238 | 13.614 | 13.048 | 3.238 | 15.272 | 0.163 | 0.163 | 0.163 | 0.119 | 0.085 | 3.272 | 0.103 | 12.512 | 0.827 |
| Enthalpy (kJ/kg) | 117.116 | 30.226 | 323.138 | 68.608 | -1,408.062 | -99.105 | -209.042 | 91.358 | 87.862 | 90.375 | --- | 1,165.597 | -1.733 | 229.322 | 3,513.235 | 164.155 |
| Density ($kg/m^3$) | 27.9 | 1.4 | 2.9 | 44.7 | 668.7 | 13.9 | 779.4 | 2.1 | 2.2 | 2.1 | 5,281.0 | 0.5 | 67.5 | 1.2 | 34.9 | 993.0 |
| V-L Molecular Weight | 28.013 | 25.166 | 2.280 | 8.554 | 17.010 | 8.805 | 43.759 | 34.655 | 35.629 | 34.923 | --- | 27.098 | 43.222 | 29.027 | 18.015 | 18.015 |
| | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 2,744 | 6,975 | 21,382 | 28,275 | 10,795 | 6,614 | 35,278 | 1,458 | 555 | 2,013 | 0 | 2,351 | 283 | 1,310 | 90,263 | 167,953 |
| V-L Flowrate (lb/hr) | 76,865 | 175,535 | 48,749 | 241,854 | 183,634 | 58,232 | 1,543,753 | 50,520 | 19,766 | 70,286 | 0 | 63,702 | 12,231 | 38,023 | 1,626,113 | 3,025,728 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 24,019 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | |
| Temperature (°F) | 242 | 77 | 77 | 70 | -13 | -19 | 95 | 119 | 119 | 119 | 351 | 450 | 95 | 123 | 1,050 | 102 |
| Pressure (psia) | 470.0 | 20.0 | 469.6 | 1,974.5 | 1,892.5 | 469.6 | 2,215.0 | 23.7 | 23.7 | 23.7 | 17.3 | 12.3 | 474.5 | 14.9 | 1,814.7 | 120.0 |
| Enthalpy (Btu/lb) | 50.4 | 13.0 | 138.9 | 29.5 | -605.4 | -42.6 | -89.9 | 39.3 | 37.8 | 38.9 | --- | 501.1 | -0.7 | 98.6 | 1,510.4 | 70.6 |
| Density ($lb/ft^3$) | 1.739 | 0.087 | 0.183 | 2.788 | 41.749 | 0.866 | 48.656 | 0.133 | 0.137 | 0.134 | 329.679 | 0.034 | 4.214 | 0.073 | 2.177 | 61.989 |

### 4.2.2 Key System Assumptions

System assumptions for Cases 3 and 4, Siemens gasifier using Illinois No. 6 coal with and without carbon sequestration, are compiled in Exhibit 4-27.

**Exhibit 4-27 Cases 3 and 4 Plant Study Configuration Matrix**

| Case | Case 3 | Case 4 |
|---|---|---|
| Gasifier Pressure, MPa (psia) | 4.2 (615) | 4.2 (615) |
| $O_2$:Coal Ratio, kg $O_2$/kg dried coal | 0.794 | 0.794 |
| Carbon Conversion, % | 99.5 | 99.5 |
| Syngas HHV at Gasifier Outlet, kJ/Nm$^3$ (Btu/scf) | 10,237 (275) | 10,237 (275) |
| Steam Cycle, MPa/°C (psig/°F) | 12.4/566/536 (1800/1050/996) | 12.4/566/537 (1800/1050/999) |
| Condenser Pressure, mm Hg (in Hg) | 51 (2.0) | 51 (2.0) |
| Combustion Turbine | N/A | N/A |
| Gasifier Technology | Siemens | Siemens |
| Oxidant | 99 vol% Oxygen | 99 vol% Oxygen |
| Coal | Illinois No. 6 | Illinois No. 6 |
| Coal Feed Moisture Content, % | 11.12 | 11.12 |
| COS Hydrolysis Reactor | Yes | Yes |
| Water Gas Shift | Yes | Yes |
| $H_2S$ Separation | Selexol (1$^{st}$ Stage) | Selexol (1$^{st}$ Stage) |
| Sulfur Removal, % | 99.4 | 99.9 |
| $CO_2$ Separation | Selexol (2$^{nd}$ Stage) | Selexol (2$^{nd}$ Stage) |
| $CO_2$ Sequestered, % | N/A | 69.0 |
| Sulfur Recovery | Claus Plant with Tail Gas Treatment / Elemental Sulfur | Claus Plant with Tail Gas Treatment / Elemental Sulfur |
| Methanation System | Based on Haldor Topsoe TREMP™ Process | Based on Haldor Topsoe TREMP™ Process |
| Ammonia Synthesis Loop | Based on Haldor Topsoe S-300 reactor with KM1R catalyst | Based on Haldor Topsoe S-300 reactor with KM1R catalyst |
| Particulate Control | Scrubber and AGR Absorber | Scrubber and AGR Absorber |
| Mercury Control | Carbon Bed | Carbon Bed |
| NOx Control | N/A | N/A |

**Balance of Plant and Sparing Philosophy**

The ammonia system feed and recycle compressors have 100 percent spares and the refrigeration unit is 2 x 50 percent. Consistent with other sections of the plant, non-rotating equipment is not spared in the ammonia plant.

The remaining balance of plant assumptions and sparing philosophy are common to all cases and are presented in Section 4.1.

### 4.2.3 Cases 3 and 4 Performance Results

The Siemens SNG and ammonia co-production plant without and with carbon sequestration and using Illinois No. 6 coal produces a net SNG output of 42 Bscf/year (90 percent CF) and 2,204 TPD of ammonia.

Overall performance for the two plants is summarized in Exhibit 4-28, which includes auxiliary power requirements and production values.  The ASU accounts for approximately 55 percent of the total auxiliary load in the non-sequestration case and 44 percent in the sequestration case, distributed between the main air compressor, the oxygen compressor, and ASU auxiliaries.  $CO_2$ compression accounts for about 18 percent of the total auxiliary load for the sequestration case. The AGR process accounts for about 11 percent and 9 percent of the auxiliary load for the non-sequestration and sequestration cases, respectively.  The ammonia production process accounts for about 16 percent and 13 percent of the auxiliary load for the non-sequestration and sequestration cases, respectively.  All other individual auxiliary loads are less than 3 percent of the total.

**Exhibit 4-28 Cases 3 and 4 Plant Performance Summary**

| POWER SUMMARY (Gross Power at Generator Terminals, kWe) | Case 3 | Case 4 |
|---|---|---|
| Steam Turbine Power | 292,300 | 292,300 |
| **TOTAL POWER, kWe** | **292,300** | **292,300** |
| **AUXILIARY LOAD SUMMARY, kWe** | | |
| Coal Handling | 700 | 700 |
| Coal Milling | 4,500 | 4,500 |
| Slag Handling | 1,140 | 1,140 |
| Air Separation Unit Auxiliaries | 1,000 | 1,000 |
| Air Separation Unit Main Air Compressor | 121,610 | 121,640 |
| Oxygen Compressor | 16,750 | 16,740 |
| Nitrogen Compressor | 9,570 | 9,570 |
| $CO_2$ Solid Feed System Compressors | 4,980 | 240 |
| SNG Compressors | 4,680 | 4,680 |
| Methanation Plant Recycle Compressor | 4,390 | 4,390 |
| Ammonia Plant Compressors | 26,290 | 26,610 |
| Ammonia Plant Refrigeration System | 15,590 | 15,580 |
| Gasifier Pilot Fuel Compressor and Incinerator Air Blower | 300 | 300 |
| $CO_2$ Compressor | N/A | 56,650 |
| Boiler Feedwater Pumps | 5,170 | 5,170 |
| Condensate Pump | 500 | 500 |
| Quench Water Pump | 640 | 640 |
| Circulating Water Pump | 6,200 | 6,220 |
| Ground Water Pumps | 670 | 680 |
| Cooling Tower Fans | 3,610 | 3,620 |
| Scrubber Pumps | 1,000 | 1,000 |
| Acid Gas Removal | 29,050 | 29,050 |
| Steam Turbine Auxiliaries | 100 | 100 |
| Claus Plant/TGTU Auxiliaries | 250 | 250 |
| Claus Plant TG Recycle Compressor | 310 | 320 |
| Miscellaneous Balance of Plant[2] | 3,000 | 3,000 |
| Transformer Losses | 1,600 | 1,740 |
| **TOTAL AUXILIARIES, kWe** | **263,600** | **316,030** |
| **NET POWER, kWe** | **28,700** | **-23,730** |
| SNG Production Rate, $MNm^3$/hr (Mscf/hr) | 150.9 (5,333) | 150.8 (5,328) |
| Ammonia Production Rate, kg/hr (lb/hr) | 83,324 (183,698) | 83,295 (183,634) |
| Conversion Efficiency ($HHV_{product}/HHV_{coal}$), % | 61.5% | 61.4% |
| Coal Feed Flow Rate, kg/hr (lb/hr) | 437,263 (964,000) | 437,263 (964,000) |
| Thermal Input[1], kWth | 3,295,885 | 3,295,885 |
| Condenser Cooling Duty, GJ/hr (MMBtu/hr) | 1,772 (1,680) | 1,772 (1,680) |
| Raw Water Withdrawal, $m^3$/min (gpm) | 28.1 (7,434) | 28.2 (7,458) |
| Raw Water Consumption, $m^3$/min (gpm) | 22.6 (5,979) | 22.7 (5,997) |

1 - HHV of Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)

2 - Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

**Environmental Performance**

The environmental targets for emissions of Hg, NOx, $SO_2$, and PM were presented in Section 2.5. A summary of the plant air emissions for Cases 3 and 4 is presented in Exhibit 4-29.

**Exhibit 4-29 Cases 3 and 4 Air Emissions**

| | kg/GJ ($lb/10^6$ Btu) | | Tonne/year (ton/year) 90% capacity factor | |
|---|---|---|---|---|
| | **Case 3** | **Case 4** | **Case 3** | **Case 4** |
| **$SO_2$** | 0.010 (0.023) | 0.0002 (0.0006) | 920 (1,014) | 23 (26) |
| **$NO_X$** | Negligible | Negligible | Negligible | Negligible |
| **Particulates** | 0.003 (0.0071) | 0.003 (0.0071) | 286 (315) | 286 (315) |
| **Hg** | 2.46E-7 (5.71E-7) | 2.46E-7 (5.71E-7) | 0.023 (0.025) | 0.023 (0.025) |
| **$CO_2$** | 63.0 (146.5) | 4.1 (9.5) | 5,891,839 (6,494,641) | 384,029 (423,320) |

The low level of $SO_2$ emissions in Case 4 is achieved by capture of the sulfur in the gas by the two-stage Selexol AGR process and co-sequestration with the $CO_2$ product. The clean syngas exiting the AGR process has a sulfur concentration of approximately 5 ppmv in both cases. The $H_2S$-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is treated using an amine based system to capture most of the remaining sulfur. The clean syngas from the TGTU is combined with a slipstream of clean syngas from the Selexol process, passed through an incinerator, and the hot, inert incinerator offgas is used to dry coal prior to being vented to the atmosphere. The higher sulfur emissions in Case 3 are due to the sulfur contained in the $CO_2$-rich stream from the Selexol process, which is vented rather than sequestered.

$NO_X$ emissions are negligible because no CT is used for power generation.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Ninety five percent of mercury is captured from the syngas by an activated carbon bed. $CO_2$ emissions represent the uncontrolled (Case 3) and controlled (Case 4) discharge from the process.

The carbon balance for the two cases is shown in Exhibit 4-30. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not used in the carbon capture equation below, but it is not neglected in the balance since Aspen accounts for air components throughout. Carbon leaves the plant as unburned carbon in the slag,

**Exhibit 4-30 Cases 3 and 4 Carbon Balance**

| Carbon In, kg/hr (lb/hr) | | | Carbon Out, kg/hr (lb/hr) | | |
|---|---|---|---|---|---|
| | Case 3 | Case 4 | | Case 3 | Case 4 |
| **Coal** | 278,732 (614,499) | 278,732 (614,499) | **Slag** | 1,394 (3,072) | 1,394 (3,072) |
| **Air ($CO_2$)** | 228 (503) | 228 (503) | **SNG Purification Off-Gas** | 10,810 (23,832) | N/A |
| | | | **ASU Vent** | 203 (447) | 203 (447) |
| | | | **SNG** | 72,874 (160,660) | 72,809 (160,517) |
| | | | **$CO_2$ Selexol Vent** | 160,170 (353,115) | N/A |
| | | | **$CO_2$ Lock Hopper Vent** | 21,082 (46,477) | 840 (1,852) |
| | | | **$CO_2$ Product** | N/A | 191,257 (421,650) |
| | | | **Stack** | 6,294 (13,876) | 6,324 (13,942) |
| | | | **PSA Purge Gas** | 6,133 (13,521) | 6,132 (13,519) |
| **Total** | 278,961 (615,003) | 278,960 (615,003) | **Total** | 278,961 (615,003) | 278,960 (615,003) |

$CO_2$ in the stack gas, waste heat boiler gas, ASU vent gas, SNG product, and the $CO_2$ product (either vented or sequestered). The carbon capture efficiency is presented in two distinct ways. The first way defines capture as the amount of carbon in the $CO_2$ product for sequestration relative to the amount of carbon in the coal less carbon contained in the slag, and is represented by the following fraction:

(Carbon in Product for Sequestration)/[(Carbon in the Coal)-(Carbon in Slag)] or
Non sequestration (Case 3)
(421,650)/(614,499-3,072)*100 = 69.0% (Case 4)

The second way does not penalize the production facility for the carbon converted to SNG product, as the end use of the SNG is unknown. In a carbon constrained scenario, the SNG may or may not be used in a scenario where CCS is implemented. For this method, the burden of carbon mitigation falls on the end-user. The following fraction represents this scenario:

[(Carbon in Product for Sequestration) + (Carbon in SNG Product)]/[(Carbon in the Coal)-(Carbon in Slag)] or
(0 + 160,660)/(614,499 – 3,072)*100 = 26.3% (Case 3)
(421,650 + 160,517)/(614,499-3,072)*100 = 95.2% (Case 4)

Exhibit 4-31 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur in the SNG product, sulfur emitted in the stack gas, and sulfur co-sequestered with the $CO_2$ product (Case 4 only). Sulfur in the slag is considered negligible. The total sulfur capture is represented by the following fraction:

(Sulfur byproduct + Sulfur in $CO_2$ product)/Sulfur in the coal or
(24,016 + 0)/24,162 = 99.4 percent (Case 3)
(24,019 + 128)/24,162 = 99.9 percent (Case 4)

**Exhibit 4-31 Cases 3 and 4 Sulfur Balance**

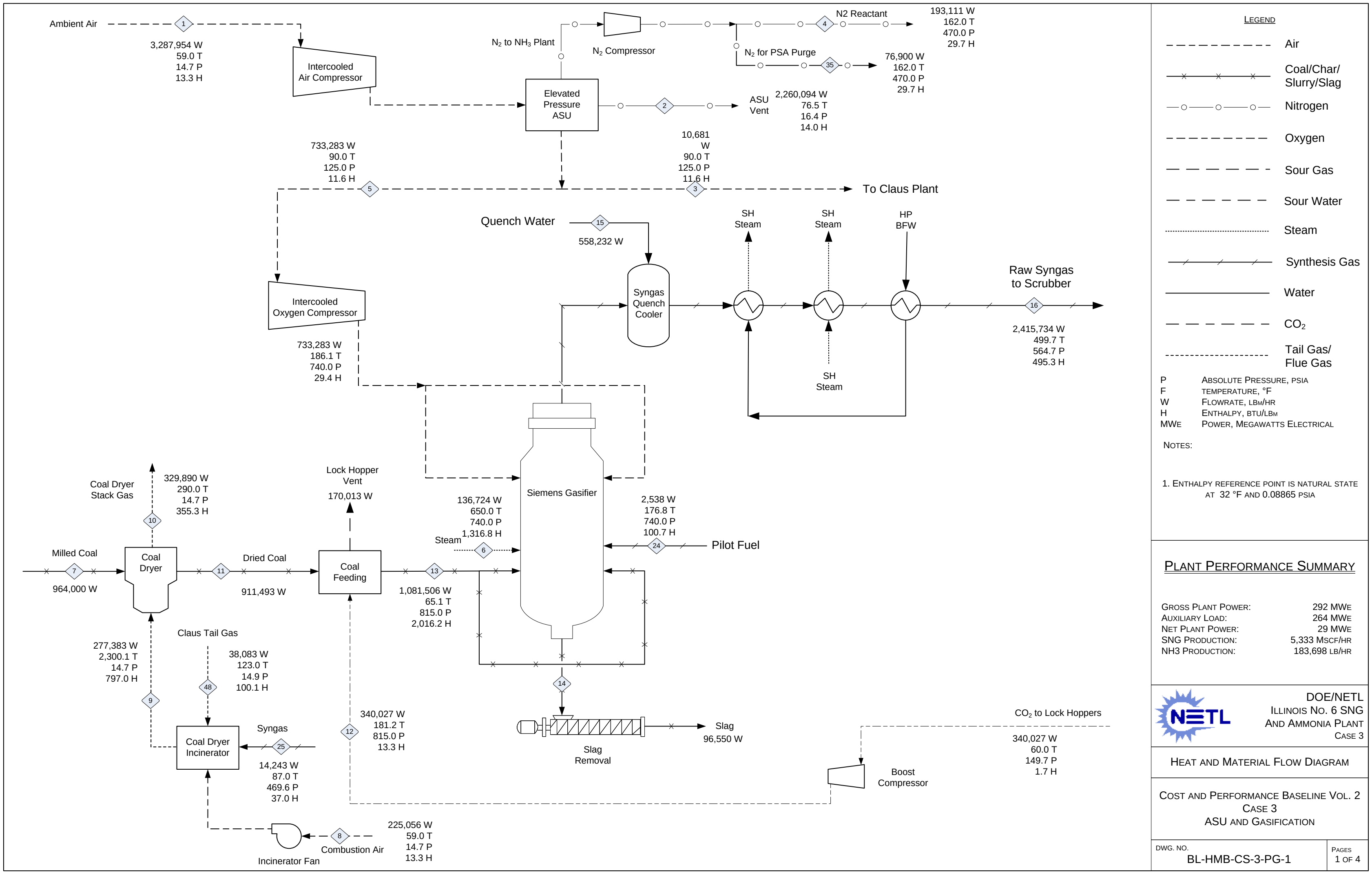
| Sulfur In, kg/hr (lb/hr) | | | Sulfur Out, kg/hr (lb/hr) | | |
|---|---|---|---|---|---|
| | **Case 3** | **Case 4** | | **Case 3** | **Case 4** |
| **Coal** | 10,960 (24,162) | 10,960 (24,162) | **Elemental Sulfur** | 10,893 (24,016) | 10,895 (24,019) |
| | | | **SNG** | 4 (9) | 4 (9) |
| | | | **Ammonia** | 0 (0) | 0 (0) |
| | | | **$CO_2$ Vent Streams/Stack** | 61 (134) | 2 (4) |
| | | | **$CO_2$ Product** | N/A | 58 (128) |
| | | | **PSA Purge Gas** | 1 (3) | 1 (3) |
| **Total** | 10,960 (24,162) | 10,960 (24,162) | **Total** | 10,960 (24,162) | 10,960 (24,162) |

Exhibit 4-32 shows the overall water balance for the plant. Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as syngas condensate, and is re-used as internal recycle. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is defined as the water removed from the ground or diverted from a surface-water source for use in the plant and was assumed to be provided 50 percent by a POTW and 50 percent from groundwater. Raw water withdrawal can be represented by the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, quench system makeup, and slag handling makeup. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source balance.

**Exhibit 4-32 Cases 3 and 4 Water Balance**

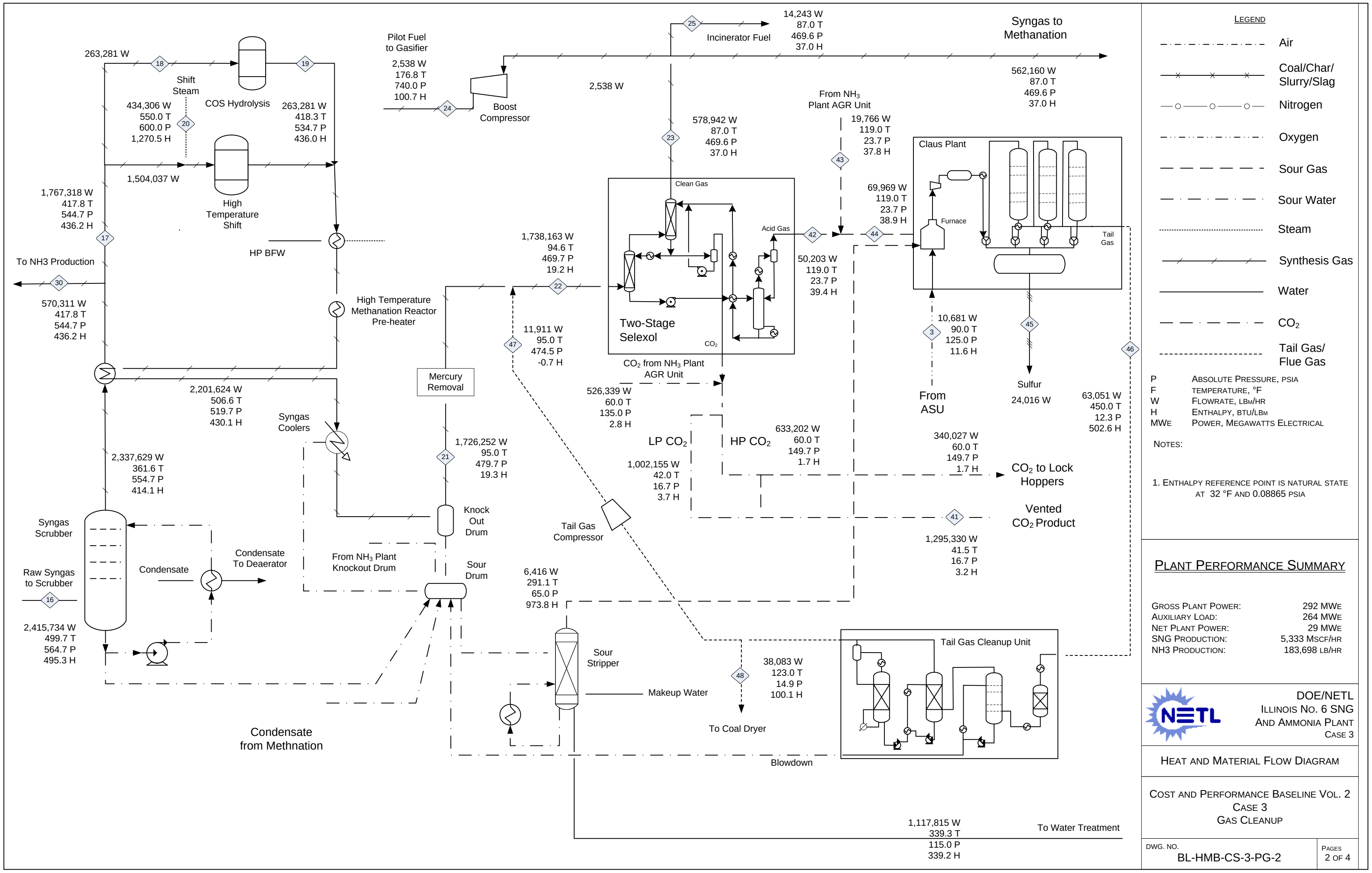
| | Case 3 | Case 4 |
|---|---|---|
| **Water Demand, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.95 (251)** | **0.95 (251)** |
| **Quench/Wash** | **4.2 (1,116)** | **4.2 (1,116)** |
| **Condenser Makeup** | **8.5 (2,258)** | **8.5 (2,258)** |
| Gasifier Steam | 1.0 (273) | 1.0 (273) |
| Shift Steam | 7.1 (1,881) | 7.1 (1,882) |
| BFW Makeup | 0.39 (103) | 0.39 (103) |
| **Cooling Tower** | **24.1 (6,371)** | **24.2 (6,396)** |
| **Total** | **37.8 (9,996)** | **37.9 (10,021)** |
| **Internal Recycle, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.95 (251)** | **0.95 (251)** |
| **Quench/Wash** | **4.2 (1,116)** | **4.2 (1,116)** |
| **Condenser Makeup** | **0.0 (0)** | **0.0 (0)** |
| **Cooling Tower** | **4.5 (1,195)** | **4.5 (1,196)** |
| Water from Coal Drying | 0.0 (0) | 0.0 (0) |
| BFW Blowdown | 0.39 (103) | 0.39 (103) |
| SWS Blowdown | 0.85 (224) | 0.85 (224) |
| SWS Excess | 3.3 (868) | 3.3 (869) |
| **Total** | **9.7 (2,563)** | **9.7 (2,563)** |
| **Raw Water Withdrawal, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.0 (0)** | **0.0 (0)** |
| **Quench/Wash** | **0.0 (0)** | **0.0 (0)** |
| **Condenser Makeup** | **8.5 (2,258)** | **8.5 (2,258)** |
| Gasifier Steam | 1.0 (273) | 1.0 (273) |
| Shift Steam | 7.1 (1,881) | 7.1 (1,882) |
| BFW Makeup | 0.39 (103) | 0.39 (103) |
| **Cooling Tower** | **19.6 (5,176)** | **19.7 (5,200)** |
| **Total** | **28.1 (7,434)** | **28.2 (7,458)** |
| **Process Water Discharge, $m^3$/min (gpm)** | | |
| **SWS Blowdown** | **0.08 (22)** | **0.08 (22)** |
| **Cooling Tower Blowdown** | **5.4 (1,433)** | **5.4 (1,438)** |
| **Total** | **5.5 (1,455)** | **5.5 (1,461)** |
| **Raw Water Consumption, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.0 (0)** | **0.0 (0)** |
| **Quench/Wash** | **0.0 (0)** | **0.0 (0)** |
| **SWS Blowdown** | **-0.08 (-22)** | **-0.08 (-22)** |
| **Condenser Makeup** | **8.5 (2,258)** | **8.5 (2,258)** |
| **Cooling Tower** | **14.2 (3,743)** | **14.2 (3,761)** |
| **Total** | **22.6 (5,979)** | **22.7 (5,997)** |

# Exhibit 4-33 Case 3 Gasification and ASU Heat and Mass Balance

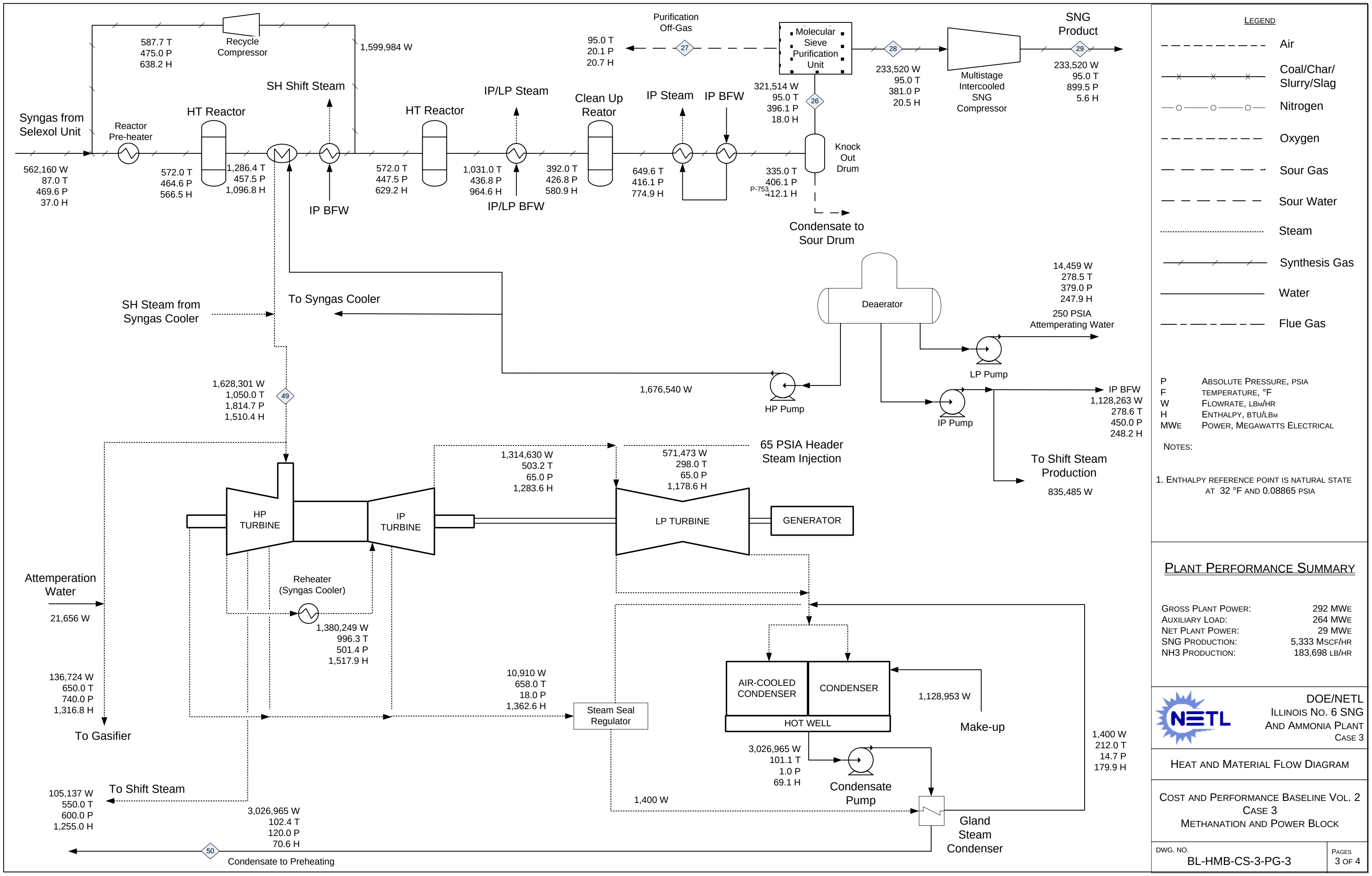



ENG UNITS

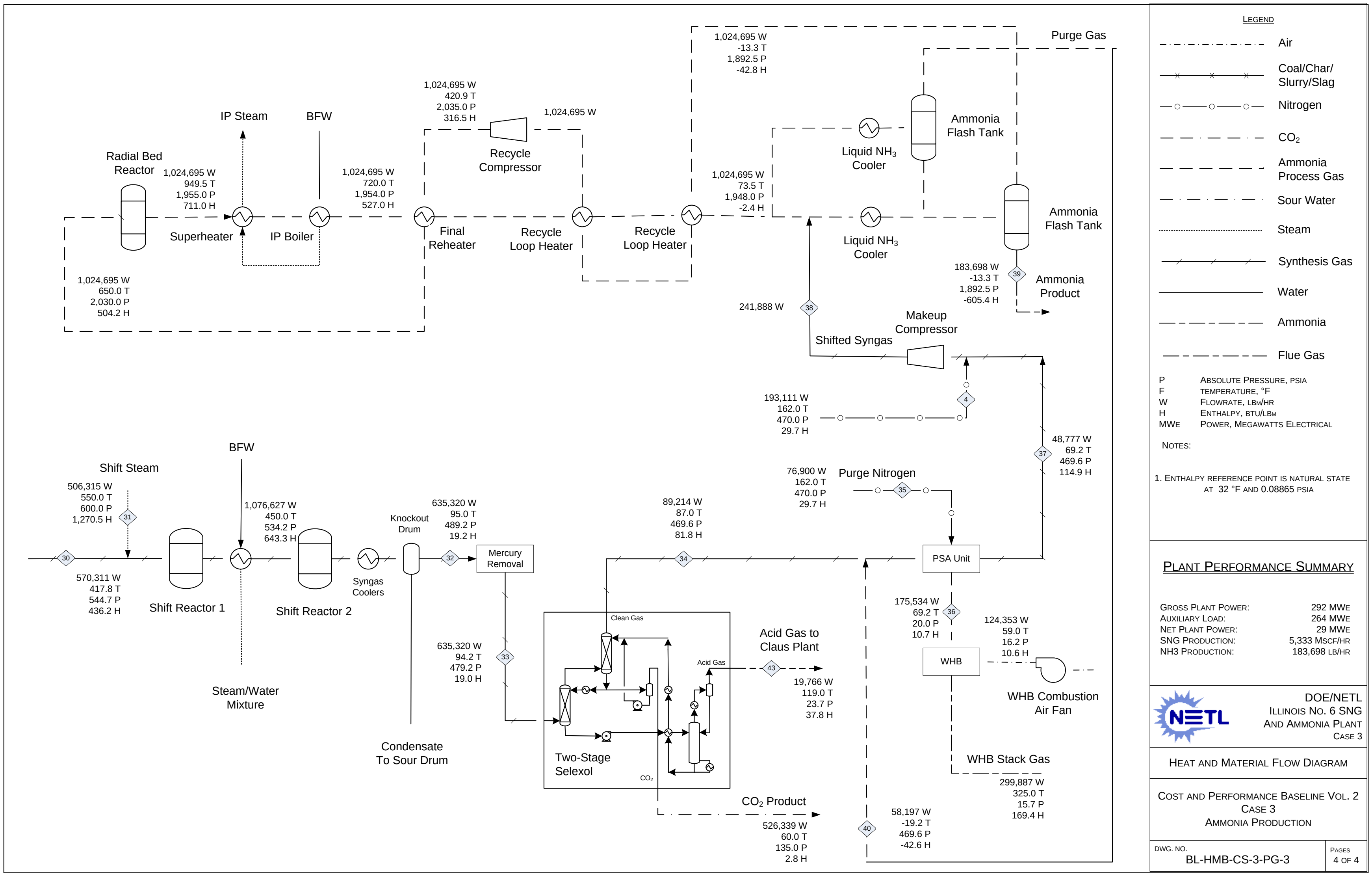
# Exhibit 4-34 Case 3 Gas Cleanup Heat and Mass Balance



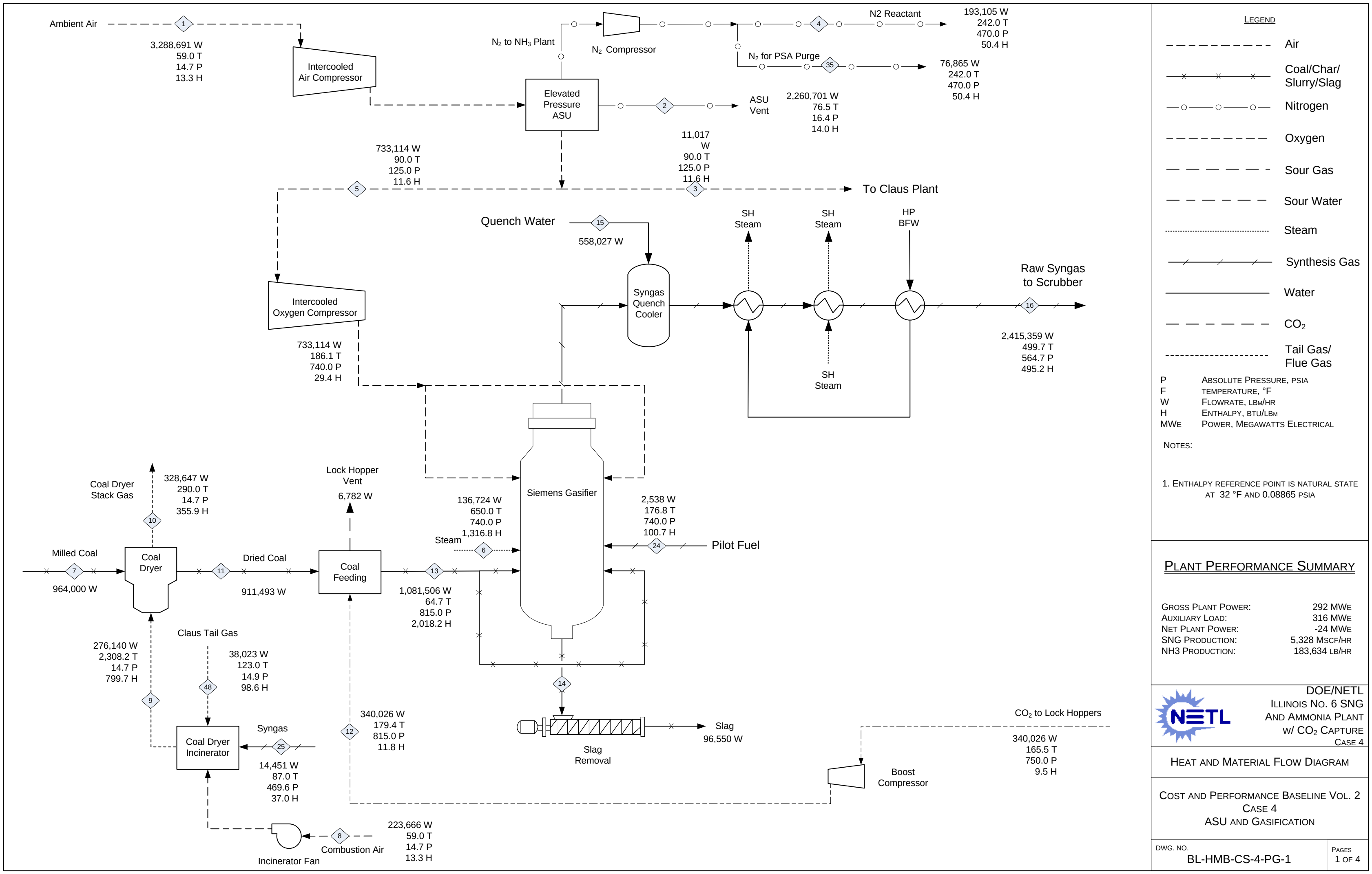
## Exhibit 4-35 Case 3 Methanation and Power Block Heat and Mass Balance



## Exhibit 4-36 Case 3 Ammonia Production Heat and Mass Balance

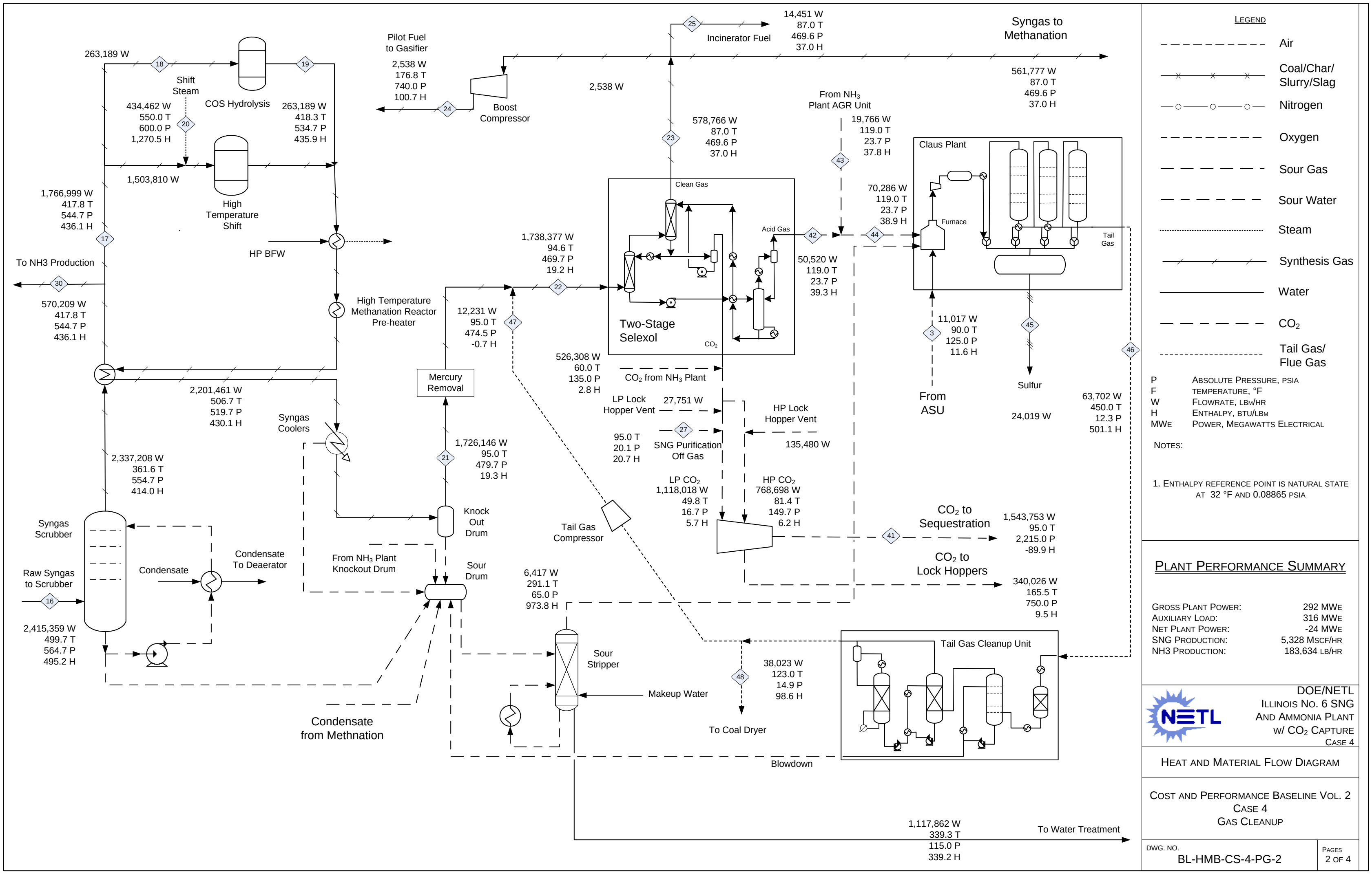


## Exhibit 4-37 Case 4 Gasification and ASU Heat and Mass Balance

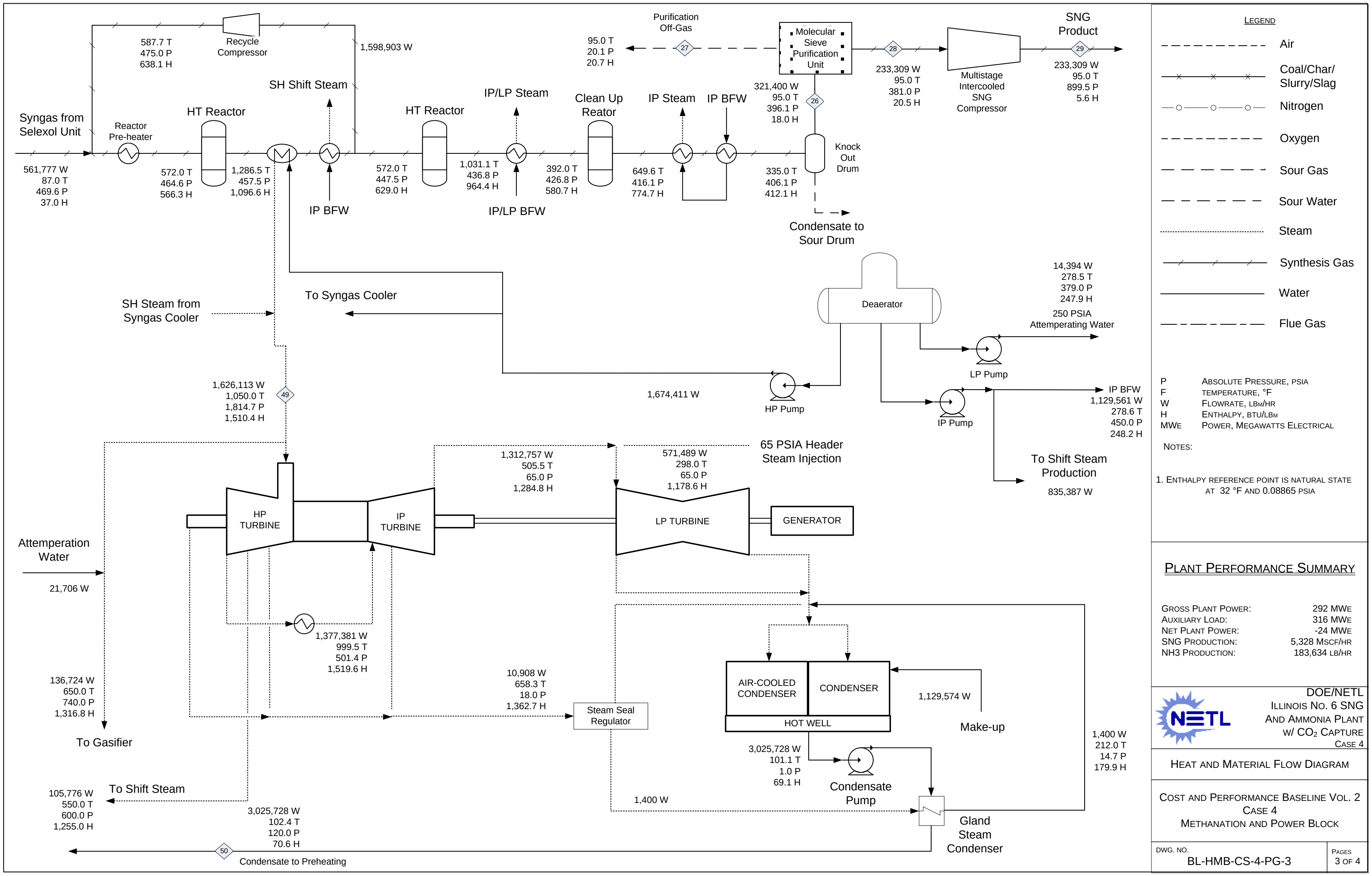


ENG UNITS

# Exhibit 4-38 Case 4 Gas Cleanup Heat and Mass Balance



## Exhibit 4-39 Case 4 Methanation and Power Block Heat and Mass Balance



# Exhibit 4-40 Case 4 Ammonia Production Heat and Mass Balance

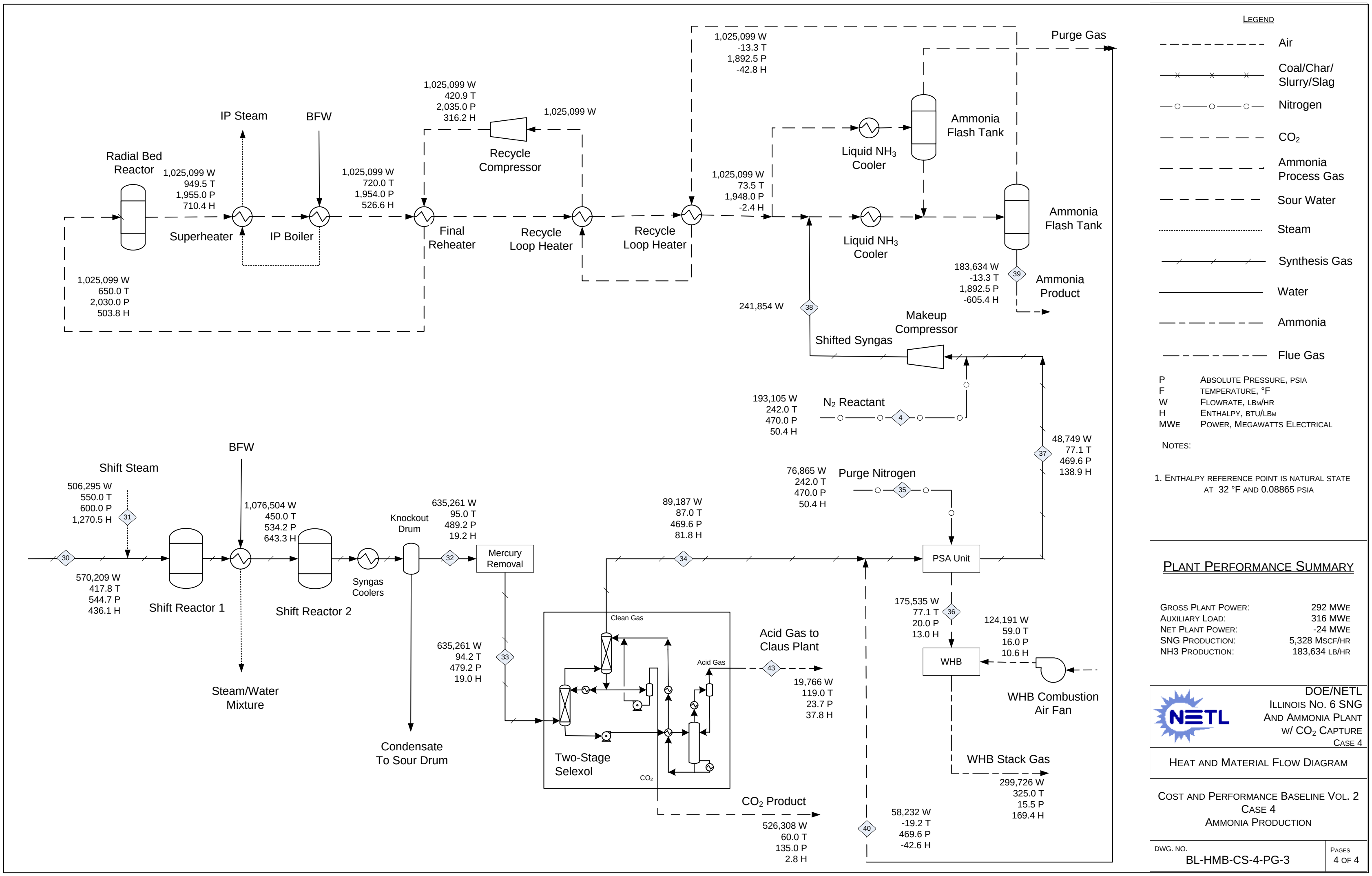


**Exhibit 4-41 Cases 3 and 4 Energy Balance**

| | HHV | | Sensible + Latent | | Power | | Total | |
|---|---|---|---|---|---|---|---|---|
| | **Case 3** | **Case 4** | **Case 3** | **Case 4** | **Case 3** | **Case 4** | **Case 3** | **Case 4** |
| **Heat In, GJ/hr (MMBtu/hr)** | | | | | | | | |
| Coal | 11,865 (11,246) | 11,865 (11,246) | 12.1 (11.5) | 12.1 (11.5) | 0 (0) | 0 (0) | 11,877 (11,257) | 11,877 (11,257) |
| ASU Air | 0 (0) | 0 (0) | 50.7 (48.0) | 50.7 (48.0) | 0 (0) | 0 (0) | 51 (48) | 51 (48) |
| Raw Water Makeup | 0 (0) | 0 (0) | 105.8 (100.3) | 106.2 (100.6) | 0 (0) | 0 (0) | 106 (100) | 106 (101) |
| Auxiliary Power | 0 (0) | 0 (0) | 0.0 (0.0) | 0.0 (0.0) | 949 (899) | 1,138 (1,078) | 949 (899) | 1,138 (1,078) |
| **Totals** | **11,865 (11,246)** | **11,865 (11,246)** | **168.6 (159.8)** | **168.9 (160.1)** | **949 (899)** | **1,138 (1,078)** | **12,983 (12,305)** | **13,172 (12,484)** |
| **Heat Out, GJ/hr (MMBtu/hr)** | | | | | | | | |
| ASU/$NH_3$ Intercoolers | 0 (0) | 0 (0) | 405 (384) | 412 (390) | 0 (0) | 0 (0) | 405 (384) | 412 (390) |
| ASU Vent | 0 (0) | 0 (0) | 33 (32) | 33 (32) | 0 (0) | 0 (0) | 33 (32) | 33 (32) |
| Slag | 46 (43) | 46 (43) | 79 (75) | 79 (75) | 0 (0) | 0 (0) | 125 (118) | 125 (118) |
| Coal Drying Stack Gas | 0 (0) | 0 (0) | 123.7 (117.2) | 123.4 (117.0) | 0 (0) | 0 (0) | 124 (117) | 123 (117) |
| Sulfur | 101 (96) | 101 (96) | 1.2 (1.2) | 1.2 (1.2) | 0 (0) | 0 (0) | 102 (97) | 102 (97) |
| $CO_2$ Intercoolers | 0 (0) | 0 (0) | 0.0 (0.0) | 345 (327) | 0 (0) | 0 (0) | 0 (0) | 345 (327) |
| $CO_2$ | 0 (0) | 0 (0) | 4.4 (4.2) | -146 (-139) | 0 (0) | 0 (0) | 4 (4) | -146 (-139) |
| Cooling Tower Blowdown | 0 (0) | 0 (0) | 40 (38) | 40 (38) | 0 (0) | 0 (0) | 40 (38) | 40 (38) |
| SNG | 5,409 (5,127) | 5,404 (5,122) | 1.4 (1.3) | 1.4 (1.3) | 0 (0) | 0 (0) | 5,410 (5,128) | 5,405 (5,123) |
| SNG Purification Off-Gas | 0 (0) | 0 (0) | 2 (2) | 0 (0) | 0 (0) | 0 (0) | 2 (2) | 0 (0) |
| Ammonia | 1,871 (1,773) | 1,870 (1,773) | -117 (-111) | -117 (-111) | 0 (0) | 0 (0) | 1,754 (1,662) | 1,753 (1,661) |
| Ammonia WHB Off-Gas | 0 (0) | 0 (0) | 54 (51) | 54 (51) | 0 (0) | 0 (0) | 54 (51) | 54 (51) |
| Condenser | 0 (0) | 0 (0) | 1,777 (1,684) | 1,776 (1,684) | 0 (0) | 0 (0) | 1,777 (1,684) | 1,776 (1,684) |
| Auxiliary Cooling Loads | 0 (0) | 0 (0) | 700 (664) | 707 (670) | 0 (0) | 0 (0) | 700 (664) | 707 (670) |
| Process Losses* | 0 (0) | 0 (0) | 1,400 (1,327) | 1,390 (1,317) | 0 (0) | 0 (0) | 1,400 (1,327) | 1,390 (1,317) |
| Power | 0 (0) | 0 (0) | 0.0 (0.0) | 0.0 (0.0) | 1,052 (997) | 1,052 (997) | 1,052 (997) | 1,052 (997) |
| **Totals** | **7,426 (7,039)** | **7,421 (7,034)** | **4,504 (4,269)** | **4,699 (4,454)** | **1,052 (997)** | **1,052 (997)** | **12,983 (12,305)** | **13,172 (12,484)** |

* Includes other energy losses not explicitly accounted for in the model.

### 4.2.4 Cases 3 and 4 Equipment Lists

Major equipment items for the Siemens gasifier SNG and ammonia co-production plant without and with carbon sequestration using Illinois No. 6 coal are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 4.2.5. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

## ACCOUNT 1 COAL HANDLING

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Bottom Trestle Dumper and Receiving Hoppers | N/A | 2 | 0 | 181 tonne (200 ton) | 181 tonne (200 ton) |
| 2 | Feeder | Belt | 2 | 0 | 572 tonne/hr (630 tph) | 572 tonne/hr (630 tph) |
| 3 | Conveyor No. 1 | Belt | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 4 | Transfer Tower No. 1 | Enclosed | 1 | 0 | N/A | N/A |
| 5 | Conveyor No. 2 | Belt | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 6 | As-Received Coal Sampling System | Two-stage | 1 | 0 | N/A | N/A |
| 7 | Stacker/Reclaimer | Traveling, linear | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 8 | Reclaim Hopper | N/A | 2 | 1 | 91 tonne (100 ton) | 91 tonne (100 ton) |
| 9 | Feeder | Vibratory | 2 | 1 | 363 tonne/hr (400 tph) | 363 tonne/hr (400 tph) |
| 10 | Conveyor No. 3 | Belt w/ tripper | 1 | 0 | 726 tonne/hr (800 tph) | 726 tonne/hr (800 tph) |
| 11 | Crusher Tower | N/A | 1 | 0 | N/A | N/A |
| 12 | Coal Surge Bin w/ Vent Filter | Dual outlet | 2 | 0 | 363 tonne (400 ton) | 363 tonne (400 ton) |
| 13 | Crusher | Impactor reduction | 2 | 0 | 8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0) | 8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0) |
| 14 | As-Fired Coal Sampling System | Swing hammer | 1 | 1 | N/A | N/A |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 15 | Conveyor No. 4 | Belt w/tripper | 1 | 0 | 726 tonne/hr (800 tph) | 726 tonne/hr (800 tph) |
| 16 | Transfer Tower No. 2 | Enclosed | 1 | 0 | N/A | N/A |
| 17 | Conveyor No. 5 | Belt w/ tripper | 1 | 0 | 726 tonne/hr (800 tph) | 726 tonne/hr (800 tph) |
| 18 | Coal Silo w/ Vent Filter and Slide Gates | Field erected | 3 | 0 | 1,633 tonne (1,800 ton) | 1,633 tonne (1,800 ton) |

**ACCOUNT 2 COAL PREPARATION AND FEED**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Feeder | Vibratory | 6 | 0 | 82 tonne/hr (90 tph) | 82 tonne/hr (90 tph) |
| 2 | Conveyor No. 6 | Belt w/tripper | 3 | 0 | 163 tonne/hr (180 tph) | 163 tonne/hr (180 tph) |
| 3 | Roller Mill Feed Hopper | Dual Outlet | 3 | 0 | 318 tonne (350 ton) | 318 tonne (350 ton) |
| 4 | Weigh Feeder | Belt | 6 | 0 | 82 tonne/hr (90 tph) | 82 tonne/hr (90 tph) |
| 5 | Pulverizer | Rotary | 6 | 0 | 82 tonne/hr (90 tph) | 82 tonne/hr (90 tph) |

**ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Demineralized Water Storage Tank | Vertical, cylindrical, outdoor | 3 | 0 | 726,799 liters (192,000 gal) | 726,799 liters (192,000 gal) |
| 2 | Condensate Pumps | Vertical canned | 2 | 1 | 12,681 lpm @ 91 m $H_2O$ (3,350 gpm @ 300 ft $H_2O$) | 12,681 lpm @ 91 m $H_2O$ (3,350 gpm @ 300 ft $H_2O$) |
| 3 | Deaerator | Horizontal spray type | 2 | 0 | 920,793 kg/hr (2,030,000 lb/hr) | 920,793 kg/hr (2,030,000 lb/hr) |
| 4 | Intermediate Pressure Feedwater Pump | Horizontal centrifugal, single stage | 2 | 1 | 8,479 lpm @ 341 m $H_2O$ (2,240 gpm @ 1120 ft $H_2O$) | 8,517 lpm @ 341 m $H_2O$ (2,250 gpm @ 1120 ft $H_2O$) |
| 5 | High Pressure Feedwater Pump | Barrel type, multi-stage, centrifugal | 2 | 1 | HP water: 7,268 lpm @ 1,676 m $H_2O$ (1,920 gpm @ 5,500 ft $H_2O$) | HP water: 7,230 lpm @ 1,676 m $H_2O$ (1,910 gpm @ 5,500 ft $H_2O$) |
| 6 | Auxiliary Boiler | Shop fabricated, water tube | 1 | 0 | 18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F) | 18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F) |
| 7 | Service Air Compressors | Flooded Screw | 3 | 1 | 28 $m^3$/min @ 0.7 MPa (1,000 scfm @ 100 psig) | 28 $m^3$/min @ 0.7 MPa (1,000 scfm @ 100 psig) |
| 8 | Instrument Air Dryers | Duplex, regenerative | 3 | 1 | 28 $m^3$/min (1,000 scfm) | 28 $m^3$/min (1,000 scfm) |
| 9 | Closed Cycle Cooling Heat Exchangers | Plate and frame | 2 | 0 | 608 GJ/hr (576 MMBtu/hr) each | 615 GJ/hr (583 MMBtu/hr) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 10 | Closed Cycle Cooling Water Pumps | Horizontal centrifugal | 2 | 1 | 218,040 lpm @ 21 m $H_2O$ (57,600 gpm @ 70 ft $H_2O$) | 220,690 lpm @ 21 m $H_2O$ (58,300 gpm @ 70 ft $H_2O$) |
| 11 | Engine-Driven Fire Pump | Vertical turbine, diesel engine | 1 | 1 | 3,785 lpm @ 107 m $H_2O$ (1,000 gpm @ 350 ft $H_2O$) | 3,785 lpm @ 107 m $H_2O$ (1,000 gpm @ 350 ft $H_2O$) |
| 12 | Fire Service Booster Pump | Two-stage horizontal centrifugal | 1 | 1 | 2,650 lpm @ 76 m $H_2O$ (700 gpm @ 250 ft $H_2O$) | 2,650 lpm @ 76 m $H_2O$ (700 gpm @ 250 ft $H_2O$) |
| 13 | Raw Water Pumps | Stainless steel, single suction | 2 | 1 | 7,798 lpm @ 18 m $H_2O$ (2,060 gpm @ 60 ft $H_2O$) | 7,836 lpm @ 18 m $H_2O$ (2,070 gpm @ 60 ft $H_2O$) |
| 14 | Ground Water Pumps | Stainless steel, single suction | 6 | 1 | 2,612 lpm @ 268 m $H_2O$ (690 gpm @ 880 ft $H_2O$) | 2,612 lpm @ 268 m $H_2O$ (690 gpm @ 880 ft $H_2O$) |
| 15 | Filtered Water Pumps | Stainless steel, single suction | 2 | 1 | 7,647 lpm @ 49 m $H_2O$ (2,020 gpm @ 160 ft $H_2O$) | 7,647 lpm @ 49 m $H_2O$ (2,020 gpm @ 160 ft $H_2O$) |
| 16 | Filtered Water Tank | Vertical, cylindrical | 2 | 0 | 3,660,493 liter (967,000 gal) | 3,660,493 liter (967,000 gal) |
| 17 | Makeup Water Demineralizer | Anion, cation, and mixed bed | 5 | 0 | 1,703 lpm (450 gpm) | 1,703 lpm (450 gpm) |
| 18 | Liquid Waste Treatment System | | 1 | 0 | 10 years, 24-hour storm | 10 years, 24-hour storm |

## ACCOUNT 4 GASIFIER, ASU, AND ACCESSORIES INCLUDING LOW TEMPERATURE HEAT RECOVERY

| Equipment No. | Description | Type | Operating Qty. | Spare s | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Gasifier | Pressurized dry-feed, entrained bed | 6 | 0 | 1,905 tonne/day, 4.2 MPa (2,100 tpd, 615 psia) | 1,905 tonne/day, 4.2 MPa (2,100 tpd, 615 psia) |
| 2 | Synthesis Gas Cooler | Convective spiral-wound tube boiler | 6 | 0 | 200,941 kg/hr (443,000 lb/hr) | 200,941 kg/hr (443,000 lb/hr) |
| 3 | Syngas Scrubber Including Sour Water Stripper | Vertical upflow | 6 | 0 | 200,941 kg/hr (443,000 lb/hr) | 200,941 kg/hr (443,000 lb/hr) |
| 4 | Raw Gas Coolers | Shell and tube with condensate drain | 12 | 0 | 414,583 kg/hr (914,000 lb/hr) | 414,583 kg/hr (914,000 lb/hr) |
| 5 | Raw Gas Knockout Drum | Vertical with mist eliminator | 3 | 0 | 287,124 kg/hr, 35°C, 3.3 MPa (633,000 lb/hr, 95°F, 485 psia) | 287,124 kg/hr, 35°C, 3.3 MPa (633,000 lb/hr, 95°F, 485 psia) |
| 6 | Flare Stack | Self-supporting, carbon steel, stainless steel top, pilot ignition | 3 | 0 | 401,883 kg/hr (886,000 lb/hr) syngas | 401,883 kg/hr (886,000 lb/hr) syngas |
| 7 | ASU Main Air Compressor | Centrifugal, multi-stage | 3 | 0 | 7,476 $m^3$/min @ 1.3 MPa (264,000 scfm @ 190 psia) | 7,476 $m^3$/min @ 1.3 MPa (264,000 scfm @ 190 psia) |
| 8 | Cold Box | Vendor design | 3 | 0 | 2,994 tonne/day (3,300 tpd) of 99% purity oxygen | 2,994 tonne/day (3,300 tpd) of 99% purity oxygen |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 9 | Oxygen Compressor | Centrifugal, multi-stage | 3 | 0 | 1,501 $m^3$/min (53,000 scfm) Suction - 0.9 MPa 130 psia) Discharge - 5.1 MPa (740 psia) | 1,501 $m^3$/min (53,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia) |
| 10 | Coal Feed System $CO_2$ Compressor | Centrifugal, multi-stage | 3 | 0 | 510 $m^3$/min (18,000 scfm) Suction - 1.0 MPa (150 psia) Discharge - 5.7 MPa (820 psia) | 510 $m^3$/min (18,000 scfm) Suction - 5.2 MPa (750 psia) Discharge - 5.7 MPa (820 psia) |

**ACCOUNT 5A SYNGAS CLEANUP**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Mercury Adsorber | Sulfated carbon bed | 3 | 0 | 392,811 kg/hr (866,000 lb/hr) 35°C (95°F) 3.3 MPa (480 psia) | 392,811 kg/hr (866,000 lb/hr) 35°C (95°F) 3.3 MPa (480 psia) |
| 2 | Zinc Oxide Guard Bed | Fixed bed | 1 | 0 | 335 $Nm^3$/min (11,827 acfm) | 335 $Nm^3$/min (11,817 acfm) |
| 3 | Sulfur Plant | Claus type | 2 | 0 | 144 tonne/day (159 tpd) | 144 tonne/day (159 tpd) |
| 4 | SNG Plant COS Hydrolysis Reactor | Fixed bed, catalytic | 2 | 0 | 65,771 kg/hr (145,000 lb/hr) 216°C (420°F) 3.7 MPa (540 psia) | 65,771 kg/hr (145,000 lb/hr) 216°C (420°F) 3.7 MPa (540 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 5 | SNG Plant Water Gas Shift Reactor | Fixed bed, catalytic | 2 | 0 | 483,529 kg/hr (1,066,000 lb/hr) 232°C (450°F) 3.7 MPa (540 psia) | 483,529 kg/hr 1,066,000 lb/hr) 232°C (450°F) 3.7 MPa (540 psia) |
| 6 | Shift Reactor Heat Recovery Exchanger | Shell and Tube | 2 each | 0 | Exchanger 1: 34 GJ/hr (32 MMBtu/hr) Exchanger 2: 61 GJ/hr (57 MMBtu/hr) Exchanger 3: 17 GJ/hr (16 MMBtu/hr) | Exchanger 1: 34 GJ/hr (32 MMBtu/hr) Exchanger 2: 61 GJ/hr (57 MMBtu/hr) Exchanger 3: 17 GJ/hr (16 MMBtu/hr) |
| 7 | Ammonia Plant Water Gas Shift Reactors | Fixed bed, catalytic | 2 | 0 | 268,527 kg/hr (592,000 lb/hr) 249°C (480°F) 3.7 MPa (540 psia) | 268,527 kg/hr (592,000 lb/hr) 249°C (480°F) 3.7 MPa (540 psia) |
| 8 | Ammonia Plant Shift Reactor Heat Recovery Exchanger | Shell and Tube | 1 | 0 | 227 GJ/hr (215 MMBtu/hr) | 227 GJ/hr (215 MMBtu/hr) |
| 9 | Acid Gas Removal Plant | Two-stage Selexol | 3 | 0 | 394,625 kg/hr (870,000 lb/hr) 35°C (95°F) 3.2 MPa (470 psia) | 394,625 kg/hr (870,000 lb/hr) 35°C (95°F) 3.2 MPa (470 psia) |
| 10 | Tail Gas Treatment Unit | Proprietary amine, absorber/stripper | 2 | 0 | 12,561 kg/hr (27,692 lb/hr) 49°C (120°F) 0.1 MPa (10.6 psia) | 12,625 kg/hr (27,833 lb/hr) 49°C (120°F) 0.1 MPa (10.6 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 11 | Tail Gas Treatment Incinerator | N/A | 2 | 0 | 69,200 kg/hr (152,561 lb/hr) 24°C (75°F) 0.1 MPa (14.9 psia) | 68,890 kg/hr (151,877 lb/hr) 24°C (74°F) 0.1 MPa (14.9 psia) |
| 12 | Tail Gas Recycle Compressor | Centrifugal | 2 | 0 | 3,060 kg/hr @ 3.3 MPa (6,746 lb/hr @ 475 psia) | 3,139 kg/hr @ 3.3 MPa (6,920 lb/hr @ 475 psia) |

**ACCOUNT 5A SNG PRODUCTION/METHANATION SYSTEM**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Methanation Reactor Preheater | Shell and Tube | 2 | 0 | 106 GJ/hr (101 MMBtu/hr) | 106 GJ/hr (100 MMBtu/hr) |
| 2 | Methanation Reactor 1 | Fixed Bed, catalytic | 2 | 0 | 539,321 kg/hr (1,189,000 lb/hr) 699°C (1,290°F) 3.2 MPa (460 psia) | 538,868 kg/hr (1,188,000 lb/hr) 699°C (1290°F) 3.2 MPa (460 psia) |
| 3 | Methanation Reactor Intercooler 1 | Shell and Tube | 2 | 0 | 462 GJ/hr (437 MMBtu/hr) | 461 GJ/hr (437 MMBtu/hr) |
| 4 | Methanation Reactor Intercooler 2 | Shell and Tube | 2 | 0 | 125 GJ/hr (119 MMBtu/hr) | 125 GJ/hr (119 MMBtu/hr) |
| 5 | Methanation Reactor 2 | Fixed Bed, catalytic | 2 | 0 | 140,160 kg/hr (309,000 lb/hr) 554°C (1,030°F) 3.1 MPa (450 psia) | 140,160 kg/hr (309,000 lb/hr) 554°C (1,030°F) 3.1 MPa (450 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 6 | Methanation Reactor Intercooler 3 | Shell and Tube | 2 | 0 | 33 GJ/hr (31 MMBtu/hr) | 33 GJ/hr (31 MMBtu/hr) |
| 7 | Methanation Reactor 3 | Fixed Bed, catalytic | 2 | 0 | 140,160 kg/hr (309,000 lb/hr) 343°C (650°F) 3.0 MPa (430 psia) | 140,160 kg/hr (309,000 lb/hr) 343°C (650°F) 3.0 MPa (430 psia) |
| 8 | Methanation Reactor Intercooler 4 | Shell and Tube | 2 | 0 | 45 GJ/hr (43 MMBtu/hr) | 45 GJ/hr (43 MMBtu/hr) |
| 9 | Methanation Reactor Intercooler 5 | Shell and Tube | 2 | 0 | 73 GJ/hr (70 MMBtu/hr) | 73 GJ/hr (69 MMBtu/hr) |
| 10 | SNG Purification Condenser 1 | Shell and Tube | 2 | 0 | 62 GJ/hr (59 MMBtu/hr) | 62 GJ/hr (59 MMBtu/hr) |
| 11 | SNG Purification Condenser 2 | Shell and Tube | 2 | 0 | 37 GJ/hr (35 MMBtu/hr) | 37 GJ/hr (35 MMBtu/hr) |
| 12 | Methanation Recycle Compressor | Centrifugal | 2 | 1 | 9,781 $m^3$/min (345,400 scfm) Suction - 3.1 MPa (448 psia) Discharge - 3.3 MPa (475 psia) | 9,772 $m^3$/min (345,100 scfm) Suction - 3.1 MPa (448 psia) Discharge - 3.3 MPa (475 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 13 | Molecular Sieve Purification Reactor | Fixed bed | 8 | 0 | 22,226 kg/hr (49,000 lb/hr)<br>129°C (265°F)<br>2.8 MPa (401 psia) | 22,226 kg/hr (49,000 lb/hr)<br>129°C (265°F)<br>2.8 MPa (401 psia) |
| 14 | SNG Product Compressor | Centrifugal, Multi-staged | 2 | 1 | 1,385 $m^3$/min<br>(48,900 scfm)<br>2.6 MPa (381 psia)<br>6.2 MPa (900 psia) | 1,382 $m^3$/min<br>(48,800 scfm)<br>2.6 MPa (381 psia)<br>6.2 MPa (900 psia) |

**ACCOUNT 5A AMMONIA SYNTHESIS LOOP**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Ammonia Converter Preheater 1 | Shell and Tube | 1 | 0 | 511,199 kg/hr<br>(1,127,000 lb/hr)<br>76°C (169°F)<br>13.4 MPa (1,949 psia)<br>107 MMBtu/hr | 511,652 kg/hr<br>(1,128,000 lb/hr)<br>76°C (169°F)<br>13.4 MPa (1,949 psia)<br>107 MMBtu/hr |
| 2 | Ammonia Converter Preheater 2 | Shell and Tube | 1 | 0 | 511,199 kg/hr<br>(1,127,000 lb/hr)<br>250°C (482°F)<br>13.4 MPa (1,949 psia)<br>278 MMBtu/hr | 511,652 kg/hr<br>(1,128,000 lb/hr)<br>250°C (482°F)<br>13.4 MPa (1,949 psia)<br>278 MMBtu/hr |
| 3 | Ammonia Converter Preheater 3 | Shell and Tube | 1 | 0 | 511,199 kg/hr<br>(1,127,000 lb/hr)<br>382°C (720°F)<br>14.0 MPa (2,035 psia)<br>212 MMBtu/hr | 511,652 kg/hr<br>(1,128,000 lb/hr)<br>382°C (720°F)<br>14.0 MPa (2,035 psia)<br>211 MMBtu/hr |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 4 | Ammonia Synthesis Loop Superheater | Shell and Tube | 1 | 0 | 511,199 kg/hr (1,127,000 lb/hr)<br>510°C (949°F)<br>13.5 MPa (1,955 psia)<br>154 MMBtu/hr | 511,652 kg/hr (1,128,000 lb/hr)<br>510°C (949°F)<br>13.5 MPa (1,955 psia)<br>153 MMBtu/hr |
| 5 | Ammonia Synthesis Loop BFW Heater | Shell and Tube | 1 | 0 | 511,199 kg/hr (1,127,000 lb/hr)<br>416°C (780°F)<br>13.5 MPa (1,955 psia)<br>54 MMBtu/hr | 511,652 kg/hr (1,128,000 lb/hr)<br>416°C (780°F)<br>13.5 MPa (1,955 psia)<br>54 MMBtu/hr |
| 6 | Ammonia Converter | Haldor Topsoe S-300 Radial Flow | 1 | 0 | 511,199 kg/hr 1,127,000 lb/hr)<br>510°C (949°F)<br>14.0 MPa (2,030 psia) | 511,652 kg/hr 1,128,000 lb/hr)<br>510°C (949°F)<br>14.0 MPa (2,030 psia) |
| 7 | Refrigeration System | Ammonia-Based | 2 | 0 | 26,025 kW refrigeration (7,400 tons refrigeration) | 26,025 kW refrigeration (7,400 tons refrigeration) |
| 8 | Primary Ammonia Separator | Refrigerated | 1 | 0 | 602,824 kg/hr (1,329,000 lb/hr)<br>-25°C (-13°F)<br>13.3 MPa (1,928 psia) | 603,278 kg/hr (1,330,000 lb/hr)<br>-25°C (-13°F)<br>13.3 MPa (1,928 psia) |
| 9 | Purge Separator | Refrigerated | 1 | 0 | 35,834 kg/hr (79,000 lb/hr)<br>-25°C (-13°F)<br>13.4 MPa (1,948 psia) | 35,834 kg/hr (79,000 lb/hr)<br>-25°C (-13°F)<br>13.4 MPa (1,948 psia) |
| 10 | Ammonia System Feed Compressor | Centrifugal, Multi-Staged | 1 | 1 | 5,573 $m^3$/min (196,800 scfm)<br>Suction - 3.24MPa (470 psia)<br>Discharge - 13.6 MPa (1,975 psia) | 5,570 $m^3$/min (196,700 scfm)<br>Suction - 3.24MPa (470 psia)<br>Discharge - 13.6 MPa (1,975 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 11 | Ammonia System Recycle Compressor | Centrifugal | 1 | 1 | 22,999 $m^3$/min (812,200 scfm) Suction - 13.0 MPa (1,892 psia) Discharge - 14.0 MPa (2,035 psia) | 22,990 $m^3$/min (811,900 scfm) Suction - 13.0 MPa (1,892 psia) Discharge - 14.0 MPa (2,035 psia) |
| 12 | Pressure Swing Absorption System | Fixed bed | 1 | 0 | 8 MM SCMD (280 MM SCFD) 21°C (69°F) 3.2 MPa (470 psia) | 8 MM SCMD (280 MM SCFD) 25°C (77°F) 3.2 MPa (470 psia) |
| 13 | Ammonia Storage Vessel | Cylindrical, Refrigerated | 2 | 0 | 32,683 tonnes (36,000 tons) Stored at -23°C (-10°F) 0.1 MPa (15 psia) | 32,683 tonnes (36,000 tons) Stored at -23°C (-10°F) 0.1 MPa (15 psia) |
| 14 | Waste Heat Boiler | Shop fabricated, water tube | 1 | 0 | 149,685 kg/hr (330,000 lb/hr) 1012°C (1,853°F) 0.1 MPa (20 psia) | 149,685 kg/hr (330,000 lb/hr) 1014°C (1,857°F) 0.1 MPa (20 psia) |

### ACCOUNT 5B $CO_2$ COMPRESSION

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | $CO_2$ Compressor | Integrally geared, multi-stage centrifugal | 4 | 0 | N/A | 1,739 $m^3$/min @ 15.3 MPa (61,400 scfm @ 2,215 psia) |

### ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Combustion Turbine | N/A | N/A | N/A | N/A | N/A |
| 2 | Combustion Turbine Generator | N/A | N/A | N/A | N/A | N/A |

### ACCOUNT 7 DUCTING & STACK

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Stack | CS plate, type 409SS liner | 1 | 0 | 76 m (250 ft) high x 3.0 m (10 ft) diameter | 76 m (250 ft) high x 2.4 m (8 ft) diameter |

## ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Steam Turbine | Commercially available | 1 | 0 | 308 MW 12.4 MPa/566°C (1800 psig/ 1050°F) | 308 MW 12.4 MPa/566°C (1800 psig/ 1050°F) |
| 2 | Steam Turbine Generator | Hydrogen cooled, static excitation | 1 | 0 | 340 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase | 340 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase |
| 3 | Surface Condenser | Single pass, divided waterbox including vacuum pumps | 1 | 0 | 1,952 GJ/hr (1,850 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F) | 1,952 GJ/hr (1,850 MMBtu/hr), Condensing temperature 16°C (60°F), Ambient temperature 11°C (20°F) |
| 4 | Air-cooled Condenser | --- | N/A | N/A | N/A | N/A |

## ACCOUNT 9 COOLING WATER SYSTEM

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Circulating Water Pumps | Vertical, wet pit | 2 | 1 | 620,808 lpm @ 30 m (164,000 gpm @ 100 ft) | 624,593 lpm @ 30 m (165,000 gpm @ 100 ft) |
| 2 | Cooling Tower | Evaporative, mechanical draft, multi-cell | 1 | 0 | 11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 3,461 GJ/hr (3,280 MMBtu/hr) heat duty | 11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 3,482 GJ/hr (3,300 MMBtu/hr) heat duty |

## ACCOUNT 10 SLAG RECOVERY AND HANDLING

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Slag Quench Tank | Water bath | 6 | 0 | 151,416 liters (40,000 gal) | 151,416 liters (40,000 gal) |
| 2 | Slag Crusher | Roll | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 3 | Slag Depressurizer | Proprietary | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 4 | Slag Receiving Tank | Horizontal, weir | 6 | 0 | 90,850 liters (24,000 gal) | 90,850 liters (24,000 gal) |
| 5 | Black Water Overflow Tank | Shop fabricated | 6 | 0 | 41,640 liters (11,000 gal) | 41,640 liters (11,000 gal) |
| 6 | Slag Conveyor | Drag chain | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 7 | Slag Separation Screen | Vibrating | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 8 | Coarse Slag Conveyor | Belt/bucket | 6 | 0 | 8 tonne/hr (9 tph) | 8 tonne/hr (9 tph) |
| 9 | Fine Ash Settling Tank | Vertical, gravity | 6 | 0 | 132,489 liters (35,000 gal) | 132,489 liters (35,000 gal) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 10 | Fine Ash Recycle Pumps | Horizontal centrifugal | 6 | 3 | 38 lpm @ 14 m $H_2O$ (10 gpm @ 46 ft $H_2O$) | 38 lpm @ 14 m $H_2O$ (10 gpm @ 46 ft $H_2O$) |
| 11 | Grey Water Storage Tank | Field erected | 6 | 0 | 41,640 liters (11,000 gal) | 41,640 liters (11,000 gal) |
| 12 | Grey Water Pumps | Centrifugal | 6 | 3 | 151 lpm @ 433 m $H_2O$ (40 gpm @ 1,420 ft $H_2O$) | 151 lpm @ 433 m $H_2O$ (40 gpm @ 1,420 ft $H_2O$) |
| 13 | Slag Storage Bin | Vertical, field erected | 6 | 0 | 544 tonne (600 tons) | 544 tonne (600 tons) |
| 14 | Unloading Equipment | Telescoping chute | 2 | 0 | 100 tonne/hr (110 tph) | 100 tonne/hr (110 tph) |

**ACCOUNT 11 ACCESSORY ELECTRIC PLANT**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | STG Step-up Transformer | Oil-filled | 1 | 0 | 24 kV/345 kV, 340 MVA, 3-ph, 60 Hz | 24 kV/345 kV, 340 MVA, 3-ph, 60 Hz |
| 2 | Auxiliary Transformer | Oil-filled | 1 | 1 | 24 kV/4.16 kV, 291 MVA, 3-ph, 60 Hz | 24 kV/4.16 kV, 349 MVA, 3-ph, 60 Hz |
| 3 | Low Voltage Transformer | Dry ventilated | 1 | 1 | 4.16 kV/480 V, 44 MVA, 3-ph, 60 Hz | 4.16 kV/480 V, 52 MVA, 3-ph, 60 Hz |
| 4 | STG Isolated Phase Bus Duct and Tap Bus | Aluminum, self-cooled | 1 | 0 | 24 kV, 3-ph, 60 Hz | 24 kV, 3-ph, 60 Hz |
| 5 | Medium Voltage Switchgear | Metal clad | 1 | 1 | 24 kV, 3-ph, 60 Hz | 4.16 kV, 3-ph, 60 Hz |
| 6 | Low Voltage Switchgear | Metal enclosed | 1 | 1 | 4.16 kV, 3-ph, 60 Hz | 480 V, 3-ph, 60 Hz |
| 7 | Emergency Diesel Generator | Sized for emergency shutdown | 1 | 0 | 750 kW, 480 V, 3-ph, 60 Hz | 750 kW, 480 V, 3-ph, 60 Hz |

## ACCOUNT 12 INSTRUMENTATION AND CONTROLS

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 3 Design Condition | Case 4 Design Condition |
|---|---|---|---|---|---|---|
| 1 | DCS - Main Control | Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W) | 1 | 0 | Operator stations/printers and engineering stations/printers | Operator stations/printers and engineering stations/printers |
| 2 | DCS - Processor | Microprocessor with redundant input/output | 1 | 0 | N/A | N/A |
| 3 | DCS - Data Highway | Fiber optic | 1 | 0 | Fully redundant, 25% spare | Fully redundant, 25% spare |

### 4.2.5 Cases 3 and 4 Cost Estimating

The cost estimating methodology was described previously in Section 2.8.

The TPC organized by cost account, owner's costs, TOC, and initial and annual O&M costs for the SNG and ammonia co-production plant without sequestration using Illinois No. 6 coal (Case 3) are shown in Exhibit 4-42 and Exhibit 4-43, respectively. The same data for the SNG and ammonia co-production plant with sequestration using Illinois No. 6 coal (Case 4) are shown in Exhibit 4-44 and Exhibit 4-45.

The estimated TOC of the co-production plant without carbon sequestration is 3.74 billion dollars and with carbon sequestration is 3.83 billion dollars. Project and process contingencies represent 11.7 and 3.3 percent for both the non-sequestration and sequestration cases. The FYCOP for SNG is $15.82/MMBtu for the non-sequestration case and $17.85/MMBtu for the sequestration case, with a consequential selling price of ammonia of $799/ton and $828/ton for the non-sequestration and sequestration cases respectively, as shown in Exhibit ES-6.

# Exhibit 4-42 Case 3 Total Plant Cost Details

| | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | Client: | USDOE/NETL | | | | | | | | Report Date: | 2010-Feb-18 |
| | Project: | Cost & Performance Baseline for Fossil Energy Plants Vol.2 | | | | | | | | | |
| | | | | TOTAL PLANT COST SUMMARY | | | | | | | |
| | Case: | Case 3 - Siemens Quench SNG & Ammonia Production w/o CO2 | | | | | | | | | |
| | Plant Size: | 28.7 | MW,net | Estimate Type: | | Conceptual | | | Cost Base (Jun) | 2007 | ($x1000) |
| Acct | | Equipment | Material | Labor | | Sales | Bare Erected | Eng'g CM | Contingencies | | Total Plant Cost |
| No. | Item/Description | Cost | Cost | Direct | Indirect | Tax | Cost $ | H.O.& Fee | Process | Project | $ |
| 1 | COAL & SORBENT HANDLING | | | | | | | | | | |
| 1.1 | Coal Receive & Unload | $5,727 | $0 | $2,798 | $0 | $0 | $8,525 | $764 | $0 | $1,858 | $11,146 |
| 1.2 | Coal Stackout & Reclaim | $7,400 | $0 | $1,794 | $0 | $0 | $9,194 | $806 | $0 | $2,000 | $12,000 |
| 1.3 | Coal Conveyors & Yd Crush | $6,880 | $0 | $1,775 | $0 | $0 | $8,655 | $760 | $0 | $1,883 | $11,298 |
| 1.4 | Other Coal Handling | $1,800 | $0 | $411 | $0 | $0 | $2,211 | $193 | $0 | $481 | $2,885 |
| 1.5 | Sorbent Receive & Unload | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.6 | Sorbent Stackout & Reclaim | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.7 | Sorbent Conveyors | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.8 | Other Sorbent Handling | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.9 | Coal & Sorbent Hnd.Foundations | $0 | $4,052 | $10,133 | $0 | $0 | $14,185 | $1,360 | $0 | $3,109 | $18,653 |
| | **SUBTOTAL 1.** | **$21,806** | **$4,052** | **$16,911** | **$0** | **$0** | **$42,769** | **$3,882** | **$0** | **$9,330** | **$55,982** |
| 2 | COAL & SORBENT PREP & FEED | | | | | | | | | | |
| 2.1 | Coal Crushing & Drying | $92,894 | $0 | $13,536 | $0 | $0 | $106,430 | $9,666 | $0 | $23,219 | $139,315 |
| 2.2 | Prepared Coal Storage & Feed | $4,400 | $1,053 | $690 | $0 | $0 | $6,143 | $525 | $0 | $1,334 | $8,002 |
| 2.3 | Dry Coal Injection System | $144,803 | $1,681 | $13,448 | $0 | $0 | $159,932 | $13,775 | $0 | $34,741 | $208,448 |
| 2.4 | Misc.Coal Prep & Feed | $2,420 | $1,761 | $5,279 | $0 | $0 | $9,459 | $869 | $0 | $2,066 | $12,394 |
| 2.5 | Sorbent Prep Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.6 | Sorbent Storage & Feed | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.7 | Sorbent Injection System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.8 | Booster Air Supply System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.9 | Coal & Sorbent Feed Foundation | $0 | $9,405 | $7,721 | $0 | $0 | $17,126 | $1,586 | $0 | $3,742 | $22,454 |
| | **SUBTOTAL 2.** | **$244,517** | **$13,899** | **$40,674** | **$0** | **$0** | **$299,090** | **$26,421** | **$0** | **$65,102** | **$390,613** |
| 3 | FEEDWATER & MISC. BOP SYSTEMS | | | | | | | | | | |
| 3.1 | Feedwater System | $3,508 | $1,630 | $1,977 | $0 | $0 | $7,115 | $643 | $0 | $1,552 | $9,310 |
| 3.2 | Water Makeup & Pretreating | $854 | $89 | $477 | $0 | $0 | $1,421 | $135 | $0 | $467 | $2,023 |
| 3.3 | Other Feedwater Subsystems | $3,808 | $1,151 | $1,884 | $0 | $0 | $6,843 | $621 | $0 | $1,493 | $8,958 |
| 3.4 | Service Water Systems | $489 | $1,006 | $3,493 | $0 | $0 | $4,988 | $487 | $0 | $1,643 | $7,118 |
| 3.5 | Other Boiler Plant Systems | $2,623 | $1,016 | $2,519 | $0 | $0 | $6,158 | $584 | $0 | $1,348 | $8,091 |
| 3.6 | FO Supply Sys & Nat Gas | $102 | $210 | $188 | $0 | $0 | $500 | $48 | $0 | $110 | $658 |
| 3.7 | Waste Treatment Equipment | $1,194 | $0 | $728 | $0 | $0 | $1,922 | $187 | $0 | $633 | $2,742 |
| 3.8 | Misc. Power Plant Equipment | $1,721 | $230 | $883 | $0 | $0 | $2,834 | $274 | $0 | $932 | $4,040 |
| | **SUBTOTAL 3.** | **$14,298** | **$5,334** | **$12,150** | **$0** | **$0** | **$31,782** | **$2,980** | **$0** | **$8,177** | **$42,939** |
| 4 | GASIFIER & ACCESSORIES | | | | | | | | | | |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries | $326,034 | $0 | $150,944 | $0 | $0 | $476,977 | $42,364 | $71,547 | $88,633 | $679,521 |
| 4.2 | Syngas Cooling | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.3 | ASU/Oxidant Compression | $250,140 | $0 | w/equip. | $0 | $0 | $250,140 | $24,246 | $0 | $27,439 | $301,825 |
| 4.4 | LT Heat Recovery & FG Saturation | $55,465 | $0 | $20,551 | $0 | $0 | $76,016 | $7,282 | $0 | $16,660 | $99,958 |
| 4.5 | Misc. Gasification Equipment | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.6 | Flare Stack System | $0 | $2,332 | $949 | $0 | $0 | $3,282 | $315 | $0 | $719 | $4,315 |
| 4.7 | CO2 Solid Feed System Compressors | $10,514 | $2,523 | $3,785 | $0 | $0 | $16,822 | $1,682 | $0 | $3,701 | $22,205 |
| 4.8 | Major Component Rigging | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.9 | Gasification Foundations | $0 | $21,135 | $12,059 | $0 | $0 | $33,194 | $3,039 | $0 | $9,058 | $45,291 |
| | **SUBTOTAL 4.** | **$642,153** | **$25,990** | **$188,288** | **$0** | **$0** | **$856,431** | **$78,928** | **$71,547** | **$146,210** | **$1,153,115** |

# Exhibit 4-42 Case 3 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 5A | GAS CLEANUP & PIPING | | | | | | | | | | |
| 5A.1 | Double Stage Selexol | $128,808 | $0 | $109,297 | $0 | $0 | $238,105 | $23,027 | $47,621 | $61,751 | $370,504 |
| 5A.2 | Elemental Sulfur Plant | $16,279 | $3,245 | $21,003 | $0 | $0 | $40,527 | $3,937 | $0 | $8,893 | $53,357 |
| 5A.3 | Mercury Removal | $2,499 | $0 | $1,902 | $0 | $0 | $4,401 | $425 | $220 | $1,009 | $6,056 |
| 5A.4a | Shift Reactors | $11,793 | $0 | $4,747 | $0 | $0 | $16,540 | $1,586 | $0 | $3,625 | $21,751 |
| 5A.4b | COS Hydrolysis | $1,494 | $0 | $1,951 | $0 | $0 | $3,445 | $335 | $0 | $756 | $4,536 |
| 5A.5 | Methanation | $21,900 | $8,760 | $13,140 | $0 | $0 | $43,800 | $4,380 | $4,380 | $10,512 | $63,072 |
| 5A.6 | SNG Purification & Compression | $17,050 | $6,820 | $10,230 | $0 | $0 | $34,100 | $3,410 | $0 | $7,502 | $45,012 |
| 5A.7 | Fuel Gas Piping | $0 | $1,749 | $1,225 | $0 | $0 | $2,974 | $276 | $0 | $650 | $3,900 |
| 5A.8 | Ammonia Production | $171,774 | $44,276 | $88,551 | $0 | $0 | $304,601 | $30,460 | $0 | $50,259 | $385,320 |
| 5A.9 | Process Interconnects | $0 | $18,400 | $27,600 | $0 | $0 | $46,000 | $4,600 | $0 | $10,120 | $60,720 |
| 5A.10 | HGCU Foundations | $0 | $1,771 | $1,142 | $0 | $0 | $2,913 | $268 | $0 | $954 | $4,135 |
| 5A.11 | Zinc Oxide Guard Bed | $664 | $0 | $122 | $0 | $0 | $786 | $79 | $0 | $173 | $1,038 |
| | **SUBTOTAL 5A.** | **$372,262** | **$85,021** | **$280,911** | **$0** | **$0** | **$738,194** | **$72,782** | **$52,221** | **$156,204** | **$1,019,401** |
| 5B | CO2 COMPRESSION | | | | | | | | | | |
| 5B.1 | CO2 Removal System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 5B.2 | CO2 Compression & Drying | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 5B.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 6 | COMBUSTION TURBINE/ACCESSORIES | | | | | | | | | | |
| 6.1 | Combustion Turbine Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.2 | Syngas Expander | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.3 | Compressed Air Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.9 | Combustion Turbine Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 6.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 7 | HRSG, DUCTING & STACK | | | | | | | | | | |
| 7.1 | Heat Recovery Steam Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.2 | HRSG Accessories | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $503 | $359 | $0 | $0 | $863 | $76 | $0 | $188 | $1,126 |
| 7.4 | Stack | $1,814 | $0 | $682 | $0 | $0 | $2,496 | $239 | $0 | $274 | $3,009 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $364 | $349 | $0 | $0 | $713 | $66 | $0 | $234 | $1,013 |
| | **SUBTOTAL 7.** | **$1,814** | **$867** | **$1,390** | **$0** | **$0** | **$4,071** | **$381** | **$0** | **$695** | **$5,147** |
| 8 | STEAM TURBINE GENERATOR | | | | | | | | | | |
| 8.1 | Steam TG & Accessories | $30,291 | $0 | $5,224 | $0 | $0 | $35,514 | $3,408 | $0 | $3,892 | $42,814 |
| 8.2 | Turbine Plant Auxiliaries | $163 | $0 | $482 | $0 | $0 | $645 | $63 | $0 | $71 | $780 |
| 8.3 | Condenser & Auxiliaries | $5,594 | $0 | $1,578 | $0 | $0 | $7,172 | $685 | $0 | $786 | $8,643 |
| 8.4 | Steam Piping | $5,482 | $0 | $3,856 | $0 | $0 | $9,338 | $802 | $0 | $2,535 | $12,675 |
| 8.9 | TG Foundations | $0 | $1,043 | $1,764 | $0 | $0 | $2,807 | $266 | $0 | $922 | $3,995 |
| | **SUBTOTAL 8.** | **$41,530** | **$1,043** | **$12,905** | **$0** | **$0** | **$55,477** | **$5,225** | **$0** | **$8,206** | **$68,908** |

## Exhibit 4-42 Case 3 Total Plant Cost Details (continued)

| Acct | | Equipment | Material | Labor | | Sales | Bare Erected | Eng'g CM | Contingencies | | Total Plant Cost |
|---|---|---|---|---|---|---|---|---|---|---|---|
| No. | Item/Description | Cost | Cost | Direct | Indirect | Tax | Cost $ | H.O.& Fee | Process | Project | $ |
| 9 | COOLING WATER SYSTEM | | | | | | | | | | |
| 9.1 | Cooling Towers | $8,777 | $0 | $1,597 | $0 | $0 | $10,373 | $988 | $0 | $1,704 | $13,065 |
| 9.2 | Circulating Water Pumps | $2,333 | $0 | $85 | $0 | $0 | $2,417 | $203 | $0 | $393 | $3,013 |
| 9.3 | Circ.Water System Auxiliaries | $185 | $0 | $26 | $0 | $0 | $212 | $20 | $0 | $35 | $267 |
| 9.4 | Circ.Water Piping | $0 | $7,740 | $2,007 | $0 | $0 | $9,747 | $881 | $0 | $2,125 | $12,753 |
| 9.5 | Make-up Water System | $450 | $0 | $643 | $0 | $0 | $1,093 | $105 | $0 | $239 | $1,437 |
| 9.6 | Component Cooling Water Sys | $914 | $1,093 | $778 | $0 | $0 | $2,784 | $261 | $0 | $609 | $3,654 |
| 9.9 | Circ.Water System Foundations | $0 | $2,834 | $4,818 | $0 | $0 | $7,652 | $725 | $0 | $2,513 | $10,891 |
| | **SUBTOTAL 9.** | **$12,658** | **$11,667** | **$9,953** | **$0** | **$0** | **$34,278** | **$3,183** | **$0** | **$7,619** | **$45,080** |
| 10 | ASH/SPENT SORBENT HANDLING SYS | | | | | | | | | | |
| 10.1 | Slag Dewatering & Cooling | $36,250 | $0 | $35,754 | $0 | $0 | $72,004 | $6,977 | $0 | $7,898 | $86,879 |
| 10.2 | Gasifier Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.3 | Cleanup Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.4 | High Temperature Ash Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.5 | Other Ash Recovery Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.6 | Ash Storage Silos | $817 | $0 | $889 | $0 | $0 | $1,707 | $166 | $0 | $281 | $2,153 |
| 10.7 | Ash Transport & Feed Equipment | $1,096 | $0 | $265 | $0 | $0 | $1,361 | $127 | $0 | $223 | $1,711 |
| 10.8 | Misc. Ash Handling Equipment | $1,693 | $2,075 | $620 | $0 | $0 | $4,388 | $418 | $0 | $721 | $5,526 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $72 | $91 | $0 | $0 | $163 | $15 | $0 | $54 | $232 |
| | **SUBTOTAL 10.** | **$39,857** | **$2,147** | **$37,618** | **$0** | **$0** | **$79,623** | **$7,702** | **$0** | **$9,177** | **$96,502** |
| 11 | ACCESSORY ELECTRIC PLANT | | | | | | | | | | |
| 11.1 | Generator Equipment | $420 | $0 | $416 | $0 | $0 | $836 | $80 | $0 | $92 | $1,007 |
| 11.2 | Station Service Equipment | $5,754 | $0 | $519 | $0 | $0 | $6,272 | $578 | $0 | $685 | $7,536 |
| 11.3 | Switchgear & Motor Control | $10,637 | $0 | $1,935 | $0 | $0 | $12,572 | $1,166 | $0 | $2,061 | $15,799 |
| 11.4 | Conduit & Cable Tray | $0 | $4,941 | $16,302 | $0 | $0 | $21,243 | $2,055 | $0 | $5,824 | $29,122 |
| 11.5 | Wire & Cable | $0 | $9,441 | $6,203 | $0 | $0 | $15,645 | $1,137 | $0 | $4,195 | $20,977 |
| 11.6 | Protective Equipment | $0 | $855 | $3,111 | $0 | $0 | $3,965 | $387 | $0 | $653 | $5,005 |
| 11.7 | Standby Equipment | $330 | $0 | $322 | $0 | $0 | $651 | $62 | $0 | $107 | $821 |
| 11.8 | Main Power Transformers | $9,274 | $0 | $182 | $0 | $0 | $9,456 | $718 | $0 | $1,526 | $11,700 |
| 11.9 | Electrical Foundations | $0 | $197 | $518 | $0 | $0 | $716 | $68 | $0 | $235 | $1,019 |
| | **SUBTOTAL 11.** | **$26,415** | **$15,435** | **$29,506** | **$0** | **$0** | **$71,356** | **$6,251** | **$0** | **$15,378** | **$92,985** |
| 12 | INSTRUMENTATION & CONTROL | | | | | | | | | | |
| 12.1 | IGCC Control Equipment | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.2 | Combustion Turbine Control | w/6.1 | $0 | w/6.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.3 | Steam Turbine Control | w/8.1 | $0 | w/8.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.4 | Other Major Component Control | $536 | $0 | $358 | $0 | $0 | $893 | $85 | $45 | $153 | $1,176 |
| 12.5 | Signal Processing Equipment | w/12.7 | $0 | w/12.7 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.6 | Control Boards,Panels & Racks | $356 | $0 | $228 | $0 | $0 | $584 | $55 | $29 | $134 | $802 |
| 12.7 | Computer & Accessories | $5,694 | $0 | $182 | $0 | $0 | $5,876 | $539 | $294 | $671 | $7,380 |
| 12.8 | Instrument Wiring & Tubing | $0 | $2,884 | $5,896 | $0 | $0 | $8,780 | $745 | $439 | $2,491 | $12,454 |
| 12.9 | Other I & C Equipment | $4,245 | $0 | $2,061 | $0 | $0 | $6,306 | $593 | $315 | $1,082 | $8,297 |
| | **SUBTOTAL 12.** | **$10,830** | **$2,884** | **$8,725** | **$0** | **$0** | **$22,439** | **$2,017** | **$1,122** | **$4,531** | **$30,110** |

## Exhibit 4-42 Case 3 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 13 | IMPROVEMENTS TO SITE | | | | | | | | | | |
| 13.1 | Site Preparation | $0 | $147 | $3,142 | $0 | $0 | $3,289 | $326 | $0 | $1,085 | $4,700 |
| 13.2 | Site Improvements | $0 | $2,614 | $3,474 | $0 | $0 | $6,089 | $601 | $0 | $2,007 | $8,696 |
| 13.3 | Site Facilities | $4,685 | $0 | $4,944 | $0 | $0 | $9,629 | $949 | $0 | $3,173 | $13,751 |
| | **SUBTOTAL 13.** | **$4,685** | **$2,761** | **$11,560** | **$0** | **$0** | **$19,006** | **$1,877** | **$0** | **$6,265** | **$27,148** |
| 14 | BUILDINGS & STRUCTURES | | | | | | | | | | |
| 14.1 | Combustion Turbine Area | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 14.2 | Steam Turbine Building | $0 | $2,530 | $3,604 | $0 | $0 | $6,134 | $564 | $0 | $1,005 | $7,703 |
| 14.3 | Administration Building | $0 | $870 | $630 | $0 | $0 | $1,500 | $134 | $0 | $245 | $1,879 |
| 14.4 | Circulation Water Pumphouse | $0 | $160 | $90 | $0 | $0 | $250 | $22 | $0 | $41 | $312 |
| 14.5 | Water Treatment Buildings | $0 | $785 | $766 | $0 | $0 | $1,552 | $140 | $0 | $254 | $1,945 |
| 14.6 | Machine Shop | $0 | $445 | $305 | $0 | $0 | $750 | $67 | $0 | $123 | $939 |
| 14.7 | Warehouse | $0 | $762 | $487 | $0 | $0 | $1,249 | $111 | $0 | $204 | $1,564 |
| 14.8 | Other Buildings & Structures | $0 | $528 | $411 | $0 | $0 | $940 | $84 | $0 | $205 | $1,228 |
| 14.9 | Waste Treating Building & Str. | $0 | $1,299 | $2,483 | $0 | $0 | $3,782 | $353 | $0 | $827 | $4,961 |
| | **SUBTOTAL 14.** | **$0** | **$7,379** | **$8,777** | **$0** | **$0** | **$16,156** | **$1,474** | **$0** | **$2,902** | **$20,532** |
| | | | | | | | | | | | |
| | **TOTAL COST** | **$1,432,827** | **$178,481** | **$659,367** | **$0** | **$0** | **$2,270,675** | **$213,101** | **$124,890** | **$439,797** | **$3,048,463** |
| | | | | | | | | | | | |
| | **Owner's Costs** | | | | | | | | | | |
| | **Preproduction Costs** | | | | | | | | | | |
| | 6 Months All Labor | | | | | | | | | | $23,796 |
| | 1 Month Maintenance Materials | | | | | | | | | | $4,247 |
| | 1 Month Non-fuel Consumables | | | | | | | | | | $819 |
| | 1 Month Waste Disposal | | | | | | | | | | $577 |
| | 25% of 1 Months Fuel Cost at 100% CF | | | | | | | | | | $3,359 |
| | 2% of TPC | | | | | | | | | | $60,969 |
| | **Total** | | | | | | | | | | **$93,767** |
| | **Inventory Capital** | | | | | | | | | | |
| | 60 day supply of fuel and consumables at 100% CF | | | | | | | | | | $28,158 |
| | 0.5% of TPC (spare parts) | | | | | | | | | | $15,242 |
| | **Total** | | | | | | | | | | **$43,400** |
| | | | | | | | | | | | |
| | **Initial Cost for Catalyst and Chemicals** | | | | | | | | | | $16,303 |
| | **Land** | | | | | | | | | | $900 |
| | **Other Owner's Costs** | | | | | | | | | | $457,269 |
| | **Financing Costs** | | | | | | | | | | $82,309 |
| | **Total Overnight Costs (TOC)** | | | | | | | | | | **$3,742,411** |
| | TASC Multiplier | | | | | | | | | | 1.201 |
| | **Total As-Spent Cost (TASC)** | | | | | | | | | | **$4,494,636** |

# Exhibit 4-43 Case 3 Initial and Annual O&M Costs

| | | | | | |
|---|---|---|---|---|---|
| INITIAL & ANNUAL O&M EXPENSES | | | | Cost Base (Jun): | 2007 |
| Case 3 - Siemens Quench SNG & Ammonia Production w/o CO2 | | | | MWe-net: | 29 |
| SNG (MMbtu/hr): 5,126 | nmonia (lb/hr): 183,698 | | | Capacity Factor (%): | 90 |
| OPERATING & MAINTENANCE LABOR | | | | | |
| Operating Labor | | | | | |
| Operating Labor Rate (base): | 34.65 $/hour | | | | |
| Operating Labor Burden: | 30.00 % of base | | | | |
| Labor O-H Charge Rate: | 25.00 % of labor | | | | |
| | | | Total | | |
| Skilled Operator | 2.0 | | 2.0 | | |
| Operator | 13.0 | | 13.0 | | |
| Foreman | 1.0 | | 1.0 | | |
| Lab Tech's, etc. | 3.0 | | 3.0 | | |
| TOTAL-O.J.'s | 19.0 | | 19.0 | | |
| | | | | | Annual Cost $ |
| Annual Operating Labor Cost | | | | | $7,497,290 |
| Maintenance Labor Cost | | | | | $30,575,935 |
| Administrative & Support Labor | | | | | $9,518,306 |
| Property Taxes and Insurance | | | | | $60,969,265 |
| **TOTAL FIXED OPERATING COSTS** | | | | | **$108,560,796** |
| VARIABLE OPERATING COSTS | | | | | |
| **Maintenance Material Cost** | | | | | **$45,863,903** |
| Consumables | Consumption Initial Fill | Consumption /Day | Unit Cost | Initial Fill Cost | |
| **Water(/1000 gallons)** | 0 | 5,352 | 1.08 | **$0** | **$1,901,899** |
| **Chemicals** | | | | | |
| MU & WT Chem. (lb) | 0 | 31,889 | 0.17 | $0 | $1,812,954 |
| Carbon (Mercury Removal) (lb) | 234,445 | 401 | 1.05 | $246,207 | $138,492 |
| COS Catalyst (m3) | 88 | 0.1 | 2,397.36 | $210,486 | $59,199 |
| Water Gas Shift Catalyst (ft3) | 8,501 | 7.5 | 498.83 | $4,240,751 | $1,222,008 |
| ZnO Sorbent (ton) | 113 | 0.4 | 12,574.00 | $1,419,605 | $1,597,055 |
| Methanation Catalyst (ft3) | 9,191 | 6.99 | 440.00 | $4,044,122 | $1,011,030 |
| Selexol Solution (gal) | 449,546 | 143 | 13.40 | $6,023,125 | $630,285 |
| SCR Catalyst (m3) | 0 | 0 | 0.00 | $0 | $0 |
| Aqueous Ammonia (ton) | 0 | 0 | 0.00 | $0 | $0 |
| Ammonia Catalyst (ft3) | 69.74 | 0.6 | 1,701.00 | $118,628 | $314,201 |
| Claus Catalyst (ft3) | w/equip | 3.75 | 131.27 | $0 | $161,850 |
| **Subtotal Chemicals** | | | | **$16,302,923** | **$6,947,076** |
| **Other** | | | | | |
| Supplemental Fuel (MBtu) | 0 | 0 | 0.00 | $0 | $0 |
| Supplemental Electricity (for consumption) (MWh) | 0 | 0 | 61.60 | $0 | $0 |
| Gases, N2 etc. (/100scf) | 0 | 0 | 0.00 | $0 | $0 |
| L.P. Steam (/1000 pounds) | 0 | 0 | 0.00 | $0 | $0 |
| **Subtotal Other** | | | | **$0** | **$0** |
| **Waste Disposal** | | | | | |
| Spent Mercury Catalyst (lb.) | 0 | 401 | 0.42 | $0 | $55,002 |
| Spent ZnO Sorbent (ton) | 0 | 0.39 | 16.23 | $0 | $2,061 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 |
| Slag (ton) | 0 | 1,159 | 16.23 | $0 | $6,175,272 |
| **Subtotal-Waste Disposal** | | | | **$0** | **$6,232,335** |
| **By-products & Emissions** | | | | | |
| Sulfur (tons) | 0 | 288 | 0.00 | $0 | $0 |
| Supplemental Electricity (for sale) (MWh) | 0 | 689 | 58.00 | $0 | -$13,123,706 |
| **Subtotal By-Products** | | | | **$0** | **-$13,123,706** |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$16,302,923** | **$47,821,506** |
| **Fuel (ton)** | 0 | 11,568 | 38.18 | **$0** | **$145,104,505** |

# Exhibit 4-44 Case 4 Total Plant Cost Details

| | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | Client: | USDOE/NETL | | | | | | | | Report Date: | 2010-Feb-18 |
| | Project: | Cost & Performance Baseline for Fossil Energy Plants Vol.2 | | | | | | | | | |
| | TOTAL PLANT COST SUMMARY | | | | | | | | | | |
| | Case: | Case 4 - Siemens Quench SNG & Ammonia Production w/ CO2 | | | | | | | | | |
| | Plant Size: | -23.73 | MW,net | Estimate Type: | | Conceptual | | | Cost Base (Jun) | 2007 | ($x1000) |
| Acct | | Equipment | Material | Labor | | Sales | Bare Erected | Eng'g CM | Contingencies | | Total Plant Cost |
| No. | Item/Description | Cost | Cost | Direct | Indirect | Tax | Cost $ | H.O.& Fee | Process | Project | $ |
| 1 | COAL & SORBENT HANDLING | | | | | | | | | | |
| 1.1 | Coal Receive & Unload | $5,727 | $0 | $2,798 | $0 | $0 | $8,525 | $764 | $0 | $1,858 | $11,146 |
| 1.2 | Coal Stackout & Reclaim | $7,400 | $0 | $1,794 | $0 | $0 | $9,194 | $806 | $0 | $2,000 | $12,000 |
| 1.3 | Coal Conveyors & Yd Crush | $6,880 | $0 | $1,775 | $0 | $0 | $8,655 | $760 | $0 | $1,883 | $11,298 |
| 1.4 | Other Coal Handling | $1,800 | $0 | $411 | $0 | $0 | $2,211 | $193 | $0 | $481 | $2,885 |
| 1.5 | Sorbent Receive & Unload | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.6 | Sorbent Stackout & Reclaim | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.7 | Sorbent Conveyors | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.8 | Other Sorbent Handling | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.9 | Coal & Sorbent Hnd.Foundations | $0 | $4,052 | $10,133 | $0 | $0 | $14,185 | $1,360 | $0 | $3,109 | $18,653 |
| | **SUBTOTAL 1.** | **$21,806** | **$4,052** | **$16,911** | **$0** | **$0** | **$42,769** | **$3,882** | **$0** | **$9,330** | **$55,982** |
| 2 | COAL & SORBENT PREP & FEED | | | | | | | | | | |
| 2.1 | Coal Crushing & Drying | $92,894 | $0 | $13,536 | $0 | $0 | $106,430 | $9,666 | $0 | $23,219 | $139,315 |
| 2.2 | Prepared Coal Storage & Feed | $4,400 | $1,053 | $690 | $0 | $0 | $6,143 | $525 | $0 | $1,334 | $8,002 |
| 2.3 | Dry Coal Injection System | $144,803 | $1,681 | $13,448 | $0 | $0 | $159,932 | $13,775 | $0 | $34,741 | $208,448 |
| 2.4 | Misc.Coal Prep & Feed | $2,420 | $1,761 | $5,279 | $0 | $0 | $9,459 | $869 | $0 | $2,066 | $12,394 |
| 2.5 | Sorbent Prep Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.6 | Sorbent Storage & Feed | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.7 | Sorbent Injection System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.8 | Booster Air Supply System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.9 | Coal & Sorbent Feed Foundation | $0 | $9,405 | $7,721 | $0 | $0 | $17,126 | $1,586 | $0 | $3,742 | $22,454 |
| | **SUBTOTAL 2.** | **$244,517** | **$13,899** | **$40,674** | **$0** | **$0** | **$299,090** | **$26,421** | **$0** | **$65,102** | **$390,613** |
| 3 | FEEDWATER & MISC. BOP SYSTEMS | | | | | | | | | | |
| 3.1 | Feedwater System | $3,504 | $1,629 | $1,975 | $0 | $0 | $7,109 | $642 | $0 | $1,550 | $9,301 |
| 3.2 | Water Makeup & Pretreating | $856 | $89 | $478 | $0 | $0 | $1,424 | $136 | $0 | $468 | $2,028 |
| 3.3 | Other Feedwater Subsystems | $3,805 | $1,150 | $1,882 | $0 | $0 | $6,837 | $621 | $0 | $1,492 | $8,949 |
| 3.4 | Service Water Systems | $490 | $1,009 | $3,501 | $0 | $0 | $5,000 | $488 | $0 | $1,646 | $7,134 |
| 3.5 | Other Boiler Plant Systems | $2,629 | $1,019 | $2,525 | $0 | $0 | $6,172 | $585 | $0 | $1,352 | $8,109 |
| 3.6 | FO Supply Sys & Nat Gas | $102 | $210 | $188 | $0 | $0 | $500 | $48 | $0 | $110 | $658 |
| 3.7 | Waste Treatment Equipment | $1,197 | $0 | $730 | $0 | $0 | $1,927 | $188 | $0 | $634 | $2,749 |
| 3.8 | Misc. Power Plant Equipment | $1,721 | $230 | $883 | $0 | $0 | $2,834 | $274 | $0 | $932 | $4,040 |
| | **SUBTOTAL 3.** | **$14,303** | **$5,336** | **$12,163** | **$0** | **$0** | **$31,803** | **$2,982** | **$0** | **$8,184** | **$42,968** |
| 4 | GASIFIER & ACCESSORIES | | | | | | | | | | |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries | $326,007 | $0 | $150,932 | $0 | $0 | $476,939 | $42,361 | $71,541 | $88,626 | $679,467 |
| 4.2 | Syngas Cooling | w/4.1 | $0 | w/ 4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.3 | ASU/Oxidant Compression | $250,140 | $0 | w/equip. | $0 | $0 | $250,140 | $24,246 | $0 | $27,439 | $301,825 |
| 4.4 | LT Heat Recovery & FG Saturation | $55,463 | $0 | $20,550 | $0 | $0 | $76,013 | $7,281 | $0 | $16,659 | $99,953 |
| 4.5 | Misc. Gasification Equipment | w/4.1 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.6 | Flare Stack System | $0 | $2,332 | $949 | $0 | $0 | $3,280 | $315 | $0 | $719 | $4,314 |
| 4.7 | CO2 Solid Feed System Compressors | $6,692 | $1,606 | $2,409 | $0 | $0 | $10,707 | $1,071 | $0 | $2,356 | $14,134 |
| 4.8 | Major Component Rigging | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.9 | Gasification Foundations | $0 | $21,135 | $12,059 | $0 | $0 | $33,194 | $3,039 | $0 | $9,058 | $45,291 |
| | **SUBTOTAL 4.** | **$638,302** | **$25,072** | **$186,899** | **$0** | **$0** | **$850,274** | **$78,312** | **$71,541** | **$144,856** | **$1,144,983** |

## Exhibit 4-44 Case 4 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 5A | GAS CLEANUP & PIPING | | | | | | | | | | |
| 5A.1 | Double Stage Selexol | $128,808 | $0 | $109,297 | $0 | $0 | $238,105 | $23,027 | $47,621 | $61,751 | $370,504 |
| 5A.2 | Elemental Sulfur Plant | $16,281 | $3,245 | $21,005 | $0 | $0 | $40,531 | $3,937 | $0 | $8,894 | $53,361 |
| 5A.3 | Mercury Removal | $2,499 | $0 | $1,901 | $0 | $0 | $4,400 | $425 | $220 | $1,009 | $6,054 |
| 5A.4a | Shift Reactors | $11,791 | $0 | $4,746 | $0 | $0 | $16,537 | $1,585 | $0 | $3,625 | $21,747 |
| 5A.4b | COS Hydrolysis | $1,493 | $0 | $1,950 | $0 | $0 | $3,443 | $335 | $0 | $756 | $4,533 |
| 5A.5 | Methanation | $21,900 | $8,760 | $13,140 | $0 | $0 | $43,800 | $4,380 | $4,380 | $10,512 | $63,072 |
| 5A.6 | SNG Purification & Compression | $17,050 | $6,820 | $10,230 | $0 | $0 | $34,100 | $3,410 | $0 | $7,502 | $45,012 |
| 5A.7 | Fuel Gas Piping | $0 | $1,749 | $1,224 | $0 | $0 | $2,973 | $276 | $0 | $650 | $3,898 |
| 5A.8 | Ammonia Production | $171,733 | $44,261 | $88,523 | $0 | $0 | $304,517 | $30,452 | $0 | $50,245 | $385,214 |
| 5A.9 | Process Interconnects | $0 | $18,400 | $27,600 | $0 | $0 | $46,000 | $4,600 | $0 | $10,120 | $60,720 |
| 5A.10 | HGCU Foundations | $0 | $1,771 | $1,142 | $0 | $0 | $2,912 | $268 | $0 | $954 | $4,134 |
| 5A.11 | Zinc Oxide Guard Bed | $664 | $0 | $122 | $0 | $0 | $786 | $79 | $0 | $173 | $1,038 |
| | **SUBTOTAL 5A.** | **$372,219** | **$85,005** | **$280,881** | **$0** | **$0** | **$738,105** | **$72,773** | **$52,221** | **$156,189** | **$1,019,288** |
| 5B | CO2 COMPRESSION | | | | | | | | | | |
| 5B.1 | CO2 Removal System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 5B.2 | CO2 Compression & Drying | $38,073 | $0 | $22,420 | $0 | $0 | $60,493 | $5,824 | $0 | $13,263 | $79,581 |
| | **SUBTOTAL 5B.** | **$38,073** | **$0** | **$22,420** | **$0** | **$0** | **$60,493** | **$5,824** | **$0** | **$13,263** | **$79,581** |
| 6 | COMBUSTION TURBINE/ACCESSORIES | | | | | | | | | | |
| 6.1 | Combustion Turbine Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.2 | Syngas Expander | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.3 | Compressed Air Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.9 | Combustion Turbine Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 6.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 7 | HRSG, DUCTING & STACK | | | | | | | | | | |
| 7.1 | Heat Recovery Steam Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.2 | HRSG Accessories | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.4 | Stack | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 7.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 8 | STEAM TURBINE GENERATOR | | | | | | | | | | |
| 8.1 | Steam TG & Accessories | $30,291 | $0 | $5,224 | $0 | $0 | $35,514 | $3,408 | $0 | $3,892 | $42,814 |
| 8.2 | Turbine Plant Auxiliaries | $163 | $0 | $482 | $0 | $0 | $645 | $63 | $0 | $71 | $779 |
| 8.3 | Condenser & Auxiliaries | $5,594 | $0 | $1,580 | $0 | $0 | $7,174 | $685 | $0 | $786 | $8,645 |
| 8.4 | Steam Piping | $5,477 | $0 | $3,853 | $0 | $0 | $9,329 | $801 | $0 | $2,533 | $12,663 |
| 8.9 | TG Foundations | $0 | $1,043 | $1,764 | $0 | $0 | $2,807 | $266 | $0 | $922 | $3,995 |
| | **SUBTOTAL 8.** | **$41,524** | **$1,043** | **$12,902** | **$0** | **$0** | **$55,470** | **$5,224** | **$0** | **$8,204** | **$68,897** |

# Exhibit 4-44 Case 4 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 9 | COOLING WATER SYSTEM | | | | | | | | | | |
| 9.1 | Cooling Towers | $7,345 | $0 | $1,603 | $0 | $0 | $8,949 | $853 | $0 | $1,470 | $11,272 |
| 9.2 | Circulating Water Pumps | $2,343 | $0 | $85 | $0 | $0 | $2,427 | $204 | $0 | $395 | $3,026 |
| 9.3 | Circ.Water System Auxiliaries | $186 | $0 | $27 | $0 | $0 | $213 | $20 | $0 | $35 | $268 |
| 9.4 | Circ.Water Piping | $0 | $7,768 | $2,014 | $0 | $0 | $9,782 | $884 | $0 | $2,133 | $12,800 |
| 9.5 | Make-up Water System | $451 | $0 | $644 | $0 | $0 | $1,095 | $105 | $0 | $240 | $1,440 |
| 9.6 | Component Cooling Water Sys | $917 | $1,097 | $780 | $0 | $0 | $2,794 | $262 | $0 | $611 | $3,667 |
| 9.9 | Circ.Water System Foundations | $0 | $2,845 | $4,836 | $0 | $0 | $7,680 | $728 | $0 | $2,523 | $10,931 |
| | **SUBTOTAL 9.** | **$11,242** | **$11,710** | **$9,989** | **$0** | **$0** | **$32,940** | **$3,056** | **$0** | **$7,407** | **$43,403** |
| 10 | ASH/SPENT SORBENT HANDLING SYS | | | | | | | | | | |
| 10.1 | Slag Dewatering & Cooling | $36,250 | $0 | $35,754 | $0 | $0 | $72,004 | $6,977 | $0 | $7,898 | $86,879 |
| 10.2 | Gasifier Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.3 | Cleanup Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.4 | High Temperature Ash Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.5 | Other Ash Recovery Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.6 | Ash Storage Silos | $817 | $0 | $889 | $0 | $0 | $1,707 | $166 | $0 | $281 | $2,153 |
| 10.7 | Ash Transport & Feed Equipment | $1,096 | $0 | $265 | $0 | $0 | $1,361 | $127 | $0 | $223 | $1,711 |
| 10.8 | Misc. Ash Handling Equipment | $1,693 | $2,075 | $620 | $0 | $0 | $4,388 | $418 | $0 | $721 | $5,526 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $72 | $91 | $0 | $0 | $163 | $15 | $0 | $54 | $232 |
| | **SUBTOTAL 10.** | **$39,857** | **$2,147** | **$37,618** | **$0** | **$0** | **$79,623** | **$7,702** | **$0** | **$9,177** | **$96,502** |
| 11 | ACCESSORY ELECTRIC PLANT | | | | | | | | | | |
| 11.1 | Generator Equipment | $420 | $0 | $416 | $0 | $0 | $836 | $80 | $0 | $92 | $1,007 |
| 11.2 | Station Service Equipment | $6,219 | $0 | $560 | $0 | $0 | $6,780 | $625 | $0 | $740 | $8,145 |
| 11.3 | Switchgear & Motor Control | $11,498 | $0 | $2,091 | $0 | $0 | $13,589 | $1,260 | $0 | $2,227 | $17,076 |
| 11.4 | Conduit & Cable Tray | $0 | $5,341 | $17,620 | $0 | $0 | $22,961 | $2,221 | $0 | $6,295 | $31,477 |
| 11.5 | Wire & Cable | $0 | $10,205 | $6,705 | $0 | $0 | $16,910 | $1,228 | $0 | $4,535 | $22,673 |
| 11.6 | Protective Equipment | $0 | $855 | $3,111 | $0 | $0 | $3,965 | $387 | $0 | $653 | $5,005 |
| 11.7 | Standby Equipment | $330 | $0 | $322 | $0 | $0 | $651 | $62 | $0 | $107 | $821 |
| 11.8 | Main Power Transformers | $9,977 | $0 | $182 | $0 | $0 | $10,159 | $771 | $0 | $1,639 | $12,569 |
| 11.9 | Electrical Foundations | $0 | $197 | $518 | $0 | $0 | $716 | $68 | $0 | $235 | $1,019 |
| | **SUBTOTAL 11.** | **$28,443** | **$16,598** | **$31,524** | **$0** | **$0** | **$76,565** | **$6,703** | **$0** | **$16,524** | **$99,792** |
| 12 | INSTRUMENTATION & CONTROL | | | | | | | | | | |
| 12.1 | IGCC Control Equipment | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.2 | Combustion Turbine Control | w/6.1 | $0 | w/6.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.3 | Steam Turbine Control | w/8.1 | $0 | w/8.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.4 | Other Major Component Control | $532 | $0 | $355 | $0 | $0 | $888 | $84 | $44 | $152 | $1,169 |
| 12.5 | Signal Processing Equipment | w/12.7 | $0 | w/12.7 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.6 | Control Boards,Panels & Racks | $353 | $0 | $227 | $0 | $0 | $580 | $55 | $29 | $133 | $797 |
| 12.7 | Computer & Accessories | $5,657 | $0 | $181 | $0 | $0 | $5,838 | $536 | $292 | $667 | $7,333 |
| 12.8 | Instrument Wiring & Tubing | $0 | $2,866 | $5,858 | $0 | $0 | $8,724 | $740 | $436 | $2,475 | $12,375 |
| 12.9 | Other I & C Equipment | $4,218 | $0 | $2,048 | $0 | $0 | $6,266 | $590 | $313 | $1,075 | $8,244 |
| | **SUBTOTAL 12.** | **$10,761** | **$2,866** | **$8,669** | **$0** | **$0** | **$22,296** | **$2,004** | **$1,115** | **$4,502** | **$29,917** |

# Exhibit 4-44 Case 4 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 13 | IMPROVEMENTS TO SITE | | | | | | | | | | |
| 13.1 | Site Preparation | $0 | $147 | $3,142 | $0 | $0 | $3,289 | $326 | $0 | $1,085 | $4,700 |
| 13.2 | Site Improvements | $0 | $2,614 | $3,474 | $0 | $0 | $6,089 | $601 | $0 | $2,007 | $8,696 |
| 13.3 | Site Facilities | $4,685 | $0 | $4,944 | $0 | $0 | $9,629 | $949 | $0 | $3,173 | $13,751 |
| | **SUBTOTAL 13.** | **$4,685** | **$2,761** | **$11,560** | **$0** | **$0** | **$19,006** | **$1,877** | **$0** | **$6,265** | **$27,148** |
| 14 | BUILDINGS & STRUCTURES | | | | | | | | | | |
| 14.1 | Combustion Turbine Area | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 14.2 | Steam Turbine Building | $0 | $2,530 | $3,604 | $0 | $0 | $6,134 | $564 | $0 | $1,005 | $7,703 |
| 14.3 | Administration Building | $0 | $870 | $630 | $0 | $0 | $1,500 | $134 | $0 | $245 | $1,879 |
| 14.4 | Circulation Water Pumphouse | $0 | $160 | $90 | $0 | $0 | $250 | $22 | $0 | $41 | $312 |
| 14.5 | Water Treatment Buildings | $0 | $787 | $768 | $0 | $0 | $1,555 | $140 | $0 | $254 | $1,950 |
| 14.6 | Machine Shop | $0 | $445 | $305 | $0 | $0 | $750 | $67 | $0 | $123 | $939 |
| 14.7 | Warehouse | $0 | $762 | $487 | $0 | $0 | $1,249 | $111 | $0 | $204 | $1,564 |
| 14.8 | Other Buildings & Structures | $0 | $528 | $411 | $0 | $0 | $940 | $84 | $0 | $205 | $1,228 |
| 14.9 | Waste Treating Building & Str. | $0 | $1,299 | $2,483 | $0 | $0 | $3,782 | $353 | $0 | $827 | $4,961 |
| | **SUBTOTAL 14.** | **$0** | **$7,381** | **$8,778** | **$0** | **$0** | **$16,159** | **$1,474** | **$0** | **$2,903** | **$20,536** |
| | | | | | | | | | | | |
| | **TOTAL COST** | **$1,465,733** | **$177,872** | **$680,989** | **$0** | **$0** | **$2,324,594** | **$218,235** | **$124,877** | **$451,906** | **$3,119,611** |
| | | | | | | | | | | | |
| | | | | | | | | | | | |
| | **Owner's Costs** | | | | | | | | | | |
| | **Preproduction Costs** | | | | | | | | | | |
| | 6 Months All Labor | | | | | | | | | | $24,443 |
| | 1 Month Maintenance Materials | | | | | | | | | | $4,391 |
| | 1 Month Non-fuel Consumables | | | | | | | | | | $1,890 |
| | 1 Month Waste Disposal | | | | | | | | | | $577 |
| | 25% of 1 Months Fuel Cost at 100% CF | | | | | | | | | | $3,359 |
| | 2% of TPC | | | | | | | | | | $62,392 |
| | **Total** | | | | | | | | | | **$97,052** |
| | **Inventory Capital** | | | | | | | | | | |
| | 60 day supply of fuel and consumables at 100% CF | | | | | | | | | | $28,163 |
| | 0.5% of TPC (spare parts) | | | | | | | | | | $15,598 |
| | **Total** | | | | | | | | | | **$43,761** |
| | | | | | | | | | | | |
| | **Initial Cost for Catalyst and Chemicals** | | | | | | | | | | $16,321 |
| | **Land** | | | | | | | | | | $900 |
| | **Other Owner's Costs** | | | | | | | | | | $467,942 |
| | **Financing Costs** | | | | | | | | | | $84,230 |
| | **Total Overnight Costs (TOC)** | | | | | | | | | | **$3,829,817** |
| | TASC Multiplier | | | | | | | | | | 1.201 |
| | **Total As-Spent Cost (TASC)** | | | | | | | | | | **$4,599,610** |

# Exhibit 4-45 Case 4 Initial and Annual O&M Costs

| INITIAL & ANNUAL O&M EXPENSES | | | | | | Cost Base (Jun): | 2007 |
|---|---|---|---|---|---|---|---|
| Case 4 - Siemens Quench SNG & Ammonia Production w/ CO2 | | | | | | MWe-net: | -24 |
| SNG (MMbtu/hr): | 5,122 | nmonia (lb/hr): | 183,634 | | | Capacity Factor (%): | 90 |
| OPERATING & MAINTENANCE LABOR | | | | | | | |
| Operating Labor | | | | | | | |
| Operating Labor Rate (base): | 34.65 | $/hour | | | | | |
| Operating Labor Burden: | 30.00 | % of base | | | | | |
| Labor O-H Charge Rate: | 25.00 | % of labor | | | | | |
| | | | Total | | | | |
| Skilled Operator | 2.0 | | 2.0 | | | | |
| Operator | 13.0 | | 13.0 | | | | |
| Foreman | 1.0 | | 1.0 | | | | |
| Lab Tech's, etc. | 3.0 | | 3.0 | | | | |
| TOTAL-O.J.'s | 19.0 | | 19.0 | | | | |
| | | | | | Annual Cost | | |
| | | | | | $ | | |
| Annual Operating Labor Cost | | | | | $7,497,290 | | |
| Maintenance Labor Cost | | | | | $31,612,112 | | |
| Administrative & Support Labor | | | | | $9,777,351 | | |
| Property Taxes and Insurance | | | | | $62,392,230 | | |
| **TOTAL FIXED OPERATING COSTS** | | | | | **$111,278,983** | | |
| VARIABLE OPERATING COSTS | | | | | | | |
| **Maintenance Material Cost** | | | | | **$47,418,169** | | |
| Consumables | Consumption Initial Fill | Consumption /Day | Unit Cost | Initial Fill Cost | | | |
| **Water(/1000 gallons)** | 0 | 5,370 | 1.08 | **$0** | **$1,908,039** | | |
| **Chemicals** | | | | | | | |
| MU & WT Chem.(lb) | 0 | 31,991 | 0.17 | $0 | $1,818,807 | | |
| Carbon (Mercury Removal) (lb) | 234,364 | 401 | 1.05 | $246,122 | $138,444 | | |
| COS Catalyst (m3) | 88 | 0.1 | 2,397.36 | $210,355 | $59,162 | | |
| Water Gas Shift Catalyst (ft3) | 8,499 | 7.5 | 498.83 | $4,239,732 | $1,221,706 | | |
| ZnO Sorbent (ton) | 115 | 0.4 | 12,574.00 | $1,442,149 | $1,622,417 | | |
| Methanation Catalyst (ft3) | 9,183 | 6.99 | 440.00 | $4,040,718 | $1,010,180 | | |
| Selexol Solution (gal) | 449,563 | 143 | 13.40 | $6,023,353 | $630,308 | | |
| SCR Catalyst (m3) | 0 | 0 | 0.00 | $0 | $0 | | |
| Aqueous Ammonia (ton) | 0 | 0 | 0.00 | $0 | $0 | | |
| Ammonia Catalyst (ft3) | 69.74 | 0.6 | 1,701.00 | $118,628 | $314,201 | | |
| Claus Catalyst (ft3) | w/equip | 3.75 | 131.27 | $0 | $161,869 | | |
| **Subtotal Chemicals** | | | | **$16,321,057** | **$6,977,094** | | |
| **Other** | | | | | | | |
| Supplemental Fuel (MBtu) | 0 | 0 | 0.00 | $0 | $0 | | |
| Supplemental Electricity (for consumption) (MV | 0 | 570 | 61.60 | $0 | $11,524,579 | | |
| Gases, N2 etc.(/100scf) | 0 | 0 | 0.00 | $0 | $0 | | |
| L.P. Steam (/1000 pounds) | 0 | 0 | 0.00 | $0 | $0 | | |
| **Subtotal Other** | | | | **$0** | **$11,524,579** | | |
| **Waste Disposal** | | | | | | | |
| Spent Mercury Catalyst (lb.) | 0 | 401 | 0.42 | $0 | $54,983 | | |
| Spent ZnO Sorbent (ton) | 0 | 0.39 | 16.23 | $0 | $2,094 | | |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | | |
| Slag (ton) | 0 | 1,159 | 16.23 | $0 | $6,175,271 | | |
| **Subtotal-Waste Disposal** | | | | **$0** | **$6,232,347** | | |
| **By-products & Emissions** | | | | | | | |
| Sulfur (tons) | 0 | 288 | 0.00 | $0 | $0 | | |
| Supplemental Electricity (for sale) (MWh) | 0 | 0 | 58.00 | $0 | $0 | | |
| **Subtotal By-Products** | | | | **$0** | **$0** | | |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$16,321,057** | **$74,060,228** | | |
| **Fuel (ton)** | 0 | 11,568 | 38.18 | **$0** | **$145,104,475** | | |

# 5. LOW RANK COAL CASES

## 5.1 SYNTHETIC NATURAL GAS PRODUCTION USING ROSEBUD PRB COAL

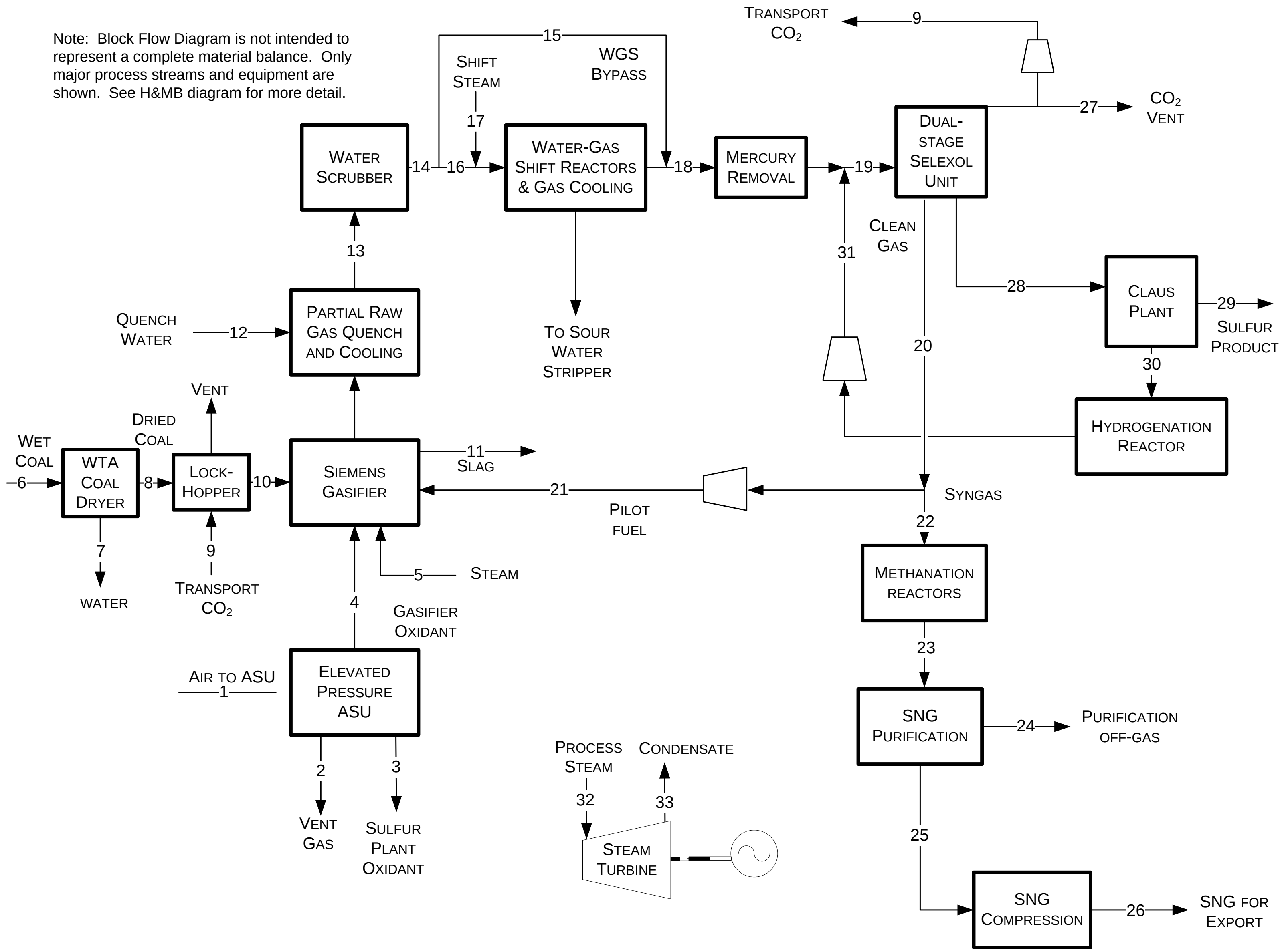
### 5.1.1 Process Description

In this section the overall Siemens gasification process for SNG production using Montana Rosebud PRB coal is described. The plant configurations are similar to Cases 1 and 2, with the exception of the coal type utilized and coal drying method. Only the major process areas that differ from Cases 1 and 2 are described below. Since the carbon content of the subbituminous coal used is lower than that of the bituminous coal, the coal feed rate is significantly increased. The process BFDs for the PRB coal non-sequestration and sequestration cases are shown in Exhibit 5-1 and Exhibit 5-3. The associated stream tables are shown in Exhibit 5-2 for the non-sequestration case and Exhibit 5-4 for the sequestration case.

**Coal Drying**

The WTA coal drying process includes a water-heated-, horizontal-, rotary-kiln coal pre-heater, a fluidized bed coal dryer and a water-cooled-, horizontal-, rotary-kiln coal cooler. The moisture driven from the coal in the fluid bed dryer passes through a baghouse for particulate removal and then is split into two streams. The smaller of the two streams is compressed and used as the fluidizing medium in the coal dryer. The bulk of the removed moisture is compressed to approximately 0.77 MPa (112 psia) and the temperature is raised to about 406°C (763°F) in the process. The high temperature vapor passes through internal coils in the dryer to provide the heat to drive off the coal moisture and then exits the dryer as liquid water. The warm water is used in the coal pre-heater before being used as cooling tower makeup water (stream 7). The vapor compressor consumes the vast majority of the WTA process auxiliary load.

**Exhibit 5-1 Case 5 Block Flow Diagram, Subbituminous Rosebud PRB Coal to SNG without Carbon Sequestration**



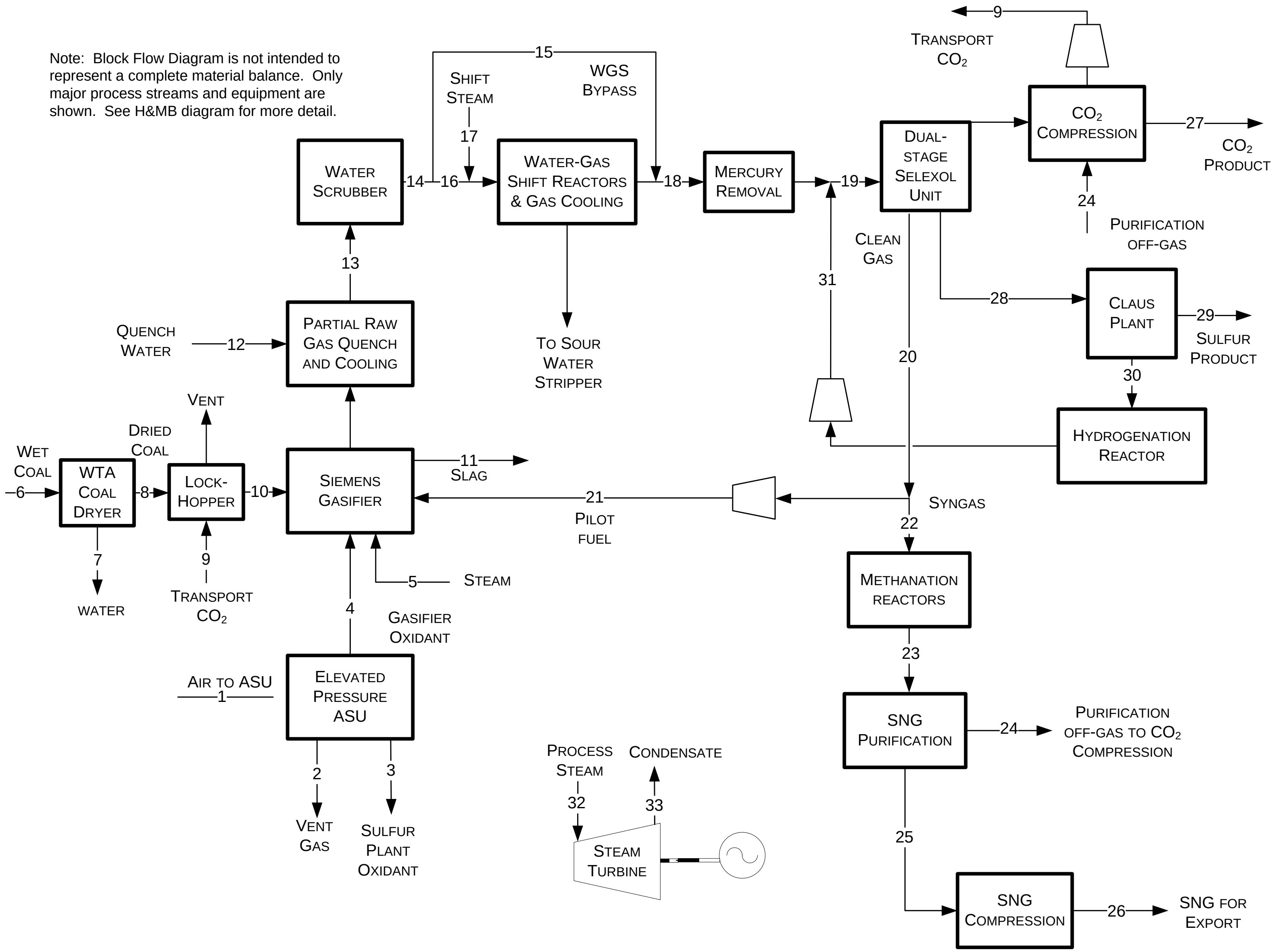
# Exhibit 5-2 Case 5 Stream Table, Subbituminous Rosebud Coal to SNG without Carbon Sequestration

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | |
| Ar | 0.0093 | 0.0091 | 0.0101 | 0.0101 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0020 | 0.0021 | 0.0021 | 0.0021 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0106 | 0.0014 | 0.0000 | 0.0000 | 0.4342 | 0.4497 | 0.4497 | 0.4497 | 0.0000 |
| $CO_2$ | 0.0003 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.9765 | 0.1307 | 0.0000 | 0.0000 | 0.0560 | 0.0580 | 0.0580 | 0.0580 | 0.0000 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0002 | 0.0002 | 0.0002 | 0.0002 | 0.0000 |
| $H_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0113 | 0.6295 | 0.0000 | 0.0000 | 0.1836 | 0.1901 | 0.1901 | 0.1901 | 0.0000 |
| $H_2O$ | 0.0064 | 0.0026 | 0.0000 | 0.0000 | 1.0000 | 0.0000 | 1.0000 | 0.0000 | 0.0013 | 0.0985 | 0.0000 | 1.0000 | 0.3188 | 0.2945 | 0.2945 | 0.2945 | 1.0000 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0022 | 0.0023 | 0.0023 | 0.0023 | 0.0000 |
| $N_2$ | 0.7759 | 0.9803 | 0.0004 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0096 | 0.0000 | 0.0000 | 0.0029 | 0.0030 | 0.0030 | 0.0030 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.2081 | 0.0076 | 0.9895 | 0.9895 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1302 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 51,342 | 40,633 | 74 | 10,414 | 2,754 | 0 | 6,669 | 0 | 4,090 | 15,275 | 0 | 12,518 | 52,503 | 50,698 | 6,569 | 44,129 | 15,436 |
| V-L Flowrate (kg/hr) | 1,483,532 | 1,143,127 | 2,361 | 334,053 | 49,619 | 0 | 120,139 | 0 | 177,211 | 202,764 | 0 | 225,516 | 1,102,187 | 1,069,647 | 138,592 | 931,055 | 278,087 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 571,223 | 0 | 451,084 | 0 | 336,925 | 48,199 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 6 | 26 | 32 | 32 | 343 | 6 | 34 | 71 | 76 | 18 | 1,427 | 216 | 260 | 214 | 214 | 214 | 288 |
| Pressure (MPa, abs) | 0.09 | 0.11 | 0.86 | 0.86 | 5.10 | 0.10 | 0.60 | 0.09 | 5.62 | 5.62 | 4.24 | 8.27 | 3.89 | 3.756 | 3.76 | 3.76 | 4.14 |
| Enthalpy (kJ/kg) | 15.71 | 30.53 | 26.94 | 26.94 | 3,177.81 | --- | 152.37 | --- | 22.14 | 3,807.88 | --- | 1,017.87 | 1,091.95 | 959.483 | 959.48 | 959.48 | 3,070.04 |
| Density ($kg/m^3$) | 1.1 | 1.3 | 11.0 | 11.0 | 20.1 | --- | 985.3 | --- | 103.2 | 33.8 | --- | 782.0 | 18.6 | 19.8 | 19.8 | 19.8 | 18.2 |
| V-L Molecular Weight | 28.895 | 28.133 | 32.078 | 32.078 | 18.015 | --- | 18.015 | --- | **43.330** | **13.275** | --- | 18.015 | 20.993 | 21.099 | 21.099 | 21.099 | 18.015 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 113,190 | 89,581 | 162 | 22,959 | 6,072 | 0 | 14,702 | 0 | 9,017 | 33,675 | 0 | 27,598 | 115,750 | 111,769 | 14,482 | 97,287 | 34,031 |
| V-L Flowrate (lb/hr) | 3,270,629 | 2,520,163 | 5,205 | 736,460 | 109,392 | 0 | 264,861 | 0 | 390,684 | 447,019 | 0 | 497,178 | 2,429,905 | 2,358,168 | 305,544 | 2,052,624 | 613,077 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 1,259,331 | 0 | 994,470 | 0 | 742,793 | 106,261 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 42 | 78 | 90 | 90 | 650 | 42 | 92 | 160 | 168 | 64 | 2,600 | 420 | 500 | 417 | 417 | 417 | 550 |
| Pressure (psia) | 13.0 | 16.4 | 125.0 | 125.0 | 740.0 | 13.8 | 86.7 | 13.5 | 815.0 | 815.0 | 614.7 | 1,200.0 | 564.7 | 544.7 | 544.7 | 544.7 | 600.0 |
| Enthalpy (Btu/lb) | 6.8 | 13.1 | 11.6 | 11.6 | 1,366.2 | --- | 65.5 | --- | 9.5 | 1,637.1 | --- | 437.6 | 469.5 | 412.5 | 412.5 | 412.5 | 1,319.9 |
| Density ($lb/ft^3$) | 0.070 | 0.080 | 0.685 | 0.685 | 1.257 | --- | 61.508 | --- | 6.441 | 2.111 | --- | 48.817 | 1.162 | 1.238 | 1.238 | 1.238 | 1.135 |
| | A - Reference conditions are 32.02 F & 0.089 PSIA | | | | | | | | | | | | | | | | |

## Exhibit 5-2 Case 5 Stream Table, Subbituminous Rosebud Coal to SNG without Carbon Sequestration (continued)

| | 18 | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | |
| Ar | 0.0021 | 0.0021 | 0.0031 | 0.0031 | 0.0031 | 0.0109 | 0.0000 | 0.0126 | 0.0126 | 0.0000 | 0.0007 | 0.0000 | 0.0016 | 0.0022 | 0.0000 | 0.0000 |
| $CH_4$ | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.8283 | 0.0000 | 0.9595 | 0.9595 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.1606 | 0.1589 | 0.2361 | 0.2361 | 0.2361 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0024 | 0.0594 | 0.0000 | 0.0851 | 0.0073 | 0.0000 | 0.0000 |
| $CO_2$ | 0.3487 | 0.3537 | 0.0405 | 0.0405 | 0.0405 | 0.1364 | 0.9832 | 0.0024 | 0.0024 | 0.9909 | 0.5950 | 0.0000 | 0.4846 | 0.7989 | 0.0000 | 0.0000 |
| COS | 0.0001 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0018 | 0.0000 | 0.0003 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.4812 | 0.4779 | 0.7156 | 0.7156 | 0.7156 | 0.0064 | 0.0000 | 0.0075 | 0.0075 | 0.0024 | 0.1138 | 0.0000 | 0.0549 | 0.1856 | 0.0000 | 0.0000 |
| $H_2O$ | 0.0018 | 0.0017 | 0.0001 | 0.0001 | 0.0001 | 0.0023 | 0.0168 | 0.0000 | 0.0000 | 0.0042 | 0.0413 | 0.0000 | 0.3702 | 0.0009 | 1.0000 | 1.0000 |
| $H_2S$ | 0.0025 | 0.0025 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1873 | 0.0000 | 0.0014 | 0.0041 | 0.0000 | 0.0000 |
| $N_2$ | 0.0030 | 0.0030 | 0.0045 | 0.0045 | 0.0045 | 0.0156 | 0.0000 | 0.0180 | 0.0180 | 0.0000 | 0.0005 | 0.0000 | 0.0007 | 0.0010 | 0.0000 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0012 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 50,536 | 51,104 | 33,915 | 149 | 33,766 | 9,681 | 1,323 | 8,358 | 8,358 | 12,415 | 684 | 0 | 803 | 568 | 43,734 | 60,796 |
| V-L Flowrate (kg/hr) | 1,066,602 | 1,087,054 | 342,281 | 1,508 | 340,773 | 195,747 | 57,660 | 138,086 | 138,086 | 543,371 | 24,191 | 0 | 24,674 | 20,452 | 787,876 | 1,095,261 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4,095 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | |
| Temperature (°C) | 35 | 34 | 31 | 80 | 31 | 35 | 35 | 35 | 35 | 5 | 48 | 175 | 232 | 21 | 566 | 33 |
| Pressure (MPa, abs) | 3.34 | 3.27 | 3.24 | 5.10 | 3.24 | 2.77 | 0.14 | 2.627 | 6.202 | 0.115 | 0.16 | 0.1 | 0.085 | 3.341 | 12.512 | 0.827 |
| Enthalpy (kJ/kg) | 42.82 | 41.44 | 85.34 | 232.30 | 85.34 | 40.71 | 46.43 | 47.903 | 12.710 | 7.664 | 101.63 | --- | 845.516 | -8.120 | 3,495.174 | 120.842 |
| Density ($kg/m^3$) | 28.0 | 27.8 | 12.8 | 17.2 | 12.8 | 23.2 | 2.4 | 17.8 | 44.5 | 2.2 | 2.2 | 5,285.9 | 0.6 | 56.5 | 34.9 | 995.1 |
| V-L Molecular Weight | 21.106 | 21.271 | 10.092 | 10.092 | 10.092 | 20.220 | 43.574 | 16.522 | 16.522 | 43.766 | 35.361 | --- | 30.719 | 36.012 | 18.015 | 18.015 |
| | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 111,414 | 112,666 | 74,770 | 329 | 74,440 | 21,342 | 2,917 | 18,425 | 18,425 | 27,371 | 1,508 | 0 | 1,771 | 1,252 | 96,416 | 134,033 |
| V-L Flowrate (lb/hr) | 2,351,455 | 2,396,544 | 754,601 | 3,325 | 751,276 | 431,548 | 127,119 | 304,428 | 304,428 | 1,197,927 | 53,332 | 0 | 54,397 | 45,090 | 1,736,970 | 2,414,637 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 9,028 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | |
| Temperature (°F) | 95 | 94 | 87 | 177 | 87 | 95 | 95 | 95 | 95 | 42 | 119 | 347 | 450 | 69 | 1,050 | 91 |
| Pressure (psia) | 484.7 | 474.7 | 469.6 | 740.0 | 469.6 | 401.1 | 20.1 | 381.0 | 899.5 | 16.7 | 23.7 | 17.3 | 12.3 | 484.5 | 1,814.7 | 120.0 |
| Enthalpy (Btu/lb) | 18.4 | 17.8 | 36.7 | 99.9 | 36.7 | 17.5 | 20.0 | 20.6 | 5.5 | 3.3 | 43.7 | --- | 363.5 | -3.5 | 1,502.7 | 52.0 |
| Density ($lb/ft^3$) | 1.751 | 1.733 | 0.798 | 1.071 | 0.798 | 1.446 | 0.148 | 1.110 | 2.778 | 0.137 | 0.136 | 330 | 0.039 | 3.528 | 2.177 | 62.123 |

**Exhibit 5-3 Case 6 Block Flow Diagram, Subbituminous Rosebud Coal to SNG with Carbon Sequestration**



## Exhibit 5-4 Case 6 Stream Table, Subbituminous Rosebud PRB Coal to SNG with Carbon Sequestration

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | |
| Ar | 0.0093 | 0.0091 | 0.0101 | 0.0101 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0020 | 0.0021 | 0.0021 | 0.0021 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0041 | 0.0005 | 0.0000 | 0.0000 | 0.4343 | 0.4497 | 0.4497 | 0.4497 | 0.0000 |
| $CO_2$ | 0.0003 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.9914 | 0.1316 | 0.0000 | 0.0000 | 0.0561 | 0.0581 | 0.0581 | 0.0581 | 0.0000 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0002 | 0.0002 | 0.0002 | 0.0002 | 0.0000 |
| $H_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0043 | 0.6293 | 0.0000 | 0.0000 | 0.1833 | 0.1898 | 0.1898 | 0.1898 | 0.0000 |
| $H_2O$ | 0.0064 | 0.0026 | 0.0000 | 0.0000 | 1.0000 | 0.0000 | 1.0000 | 0.0000 | 0.0001 | 0.0985 | 0.0000 | 1.0000 | 0.3189 | 0.2946 | 0.2946 | 0.2946 | 1.0000 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0022 | 0.0023 | 0.0023 | 0.0023 | 0.0000 |
| $N_2$ | 0.7759 | 0.9803 | 0.0004 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0096 | 0.0000 | 0.0000 | 0.0029 | 0.0030 | 0.0030 | 0.0030 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.2081 | 0.0076 | 0.9895 | 0.9895 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1304 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 0.0000 | 1.0000 | | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 51,336 | 40,628 | 73 | 10,413 | 2,754 | 0 | 6,669 | 0 | 4,049 | 15,254 | 0 | 12,514 | 52,479 | 50,673 | 6,564 | 44,109 | 15,439 |
| V-L Flowrate (kg/hr) | 1,483,363 | 1,142,996 | 2,343 | 334,032 | 49,619 | 0 | 120,139 | 0 | 177,211 | 202,764 | 0 | 225,439 | 1,102,088 | 1,069,598 | 138,555 | 931,043 | 278,144 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 571,223 | 0 | 451,084 | 0 | 336,925 | 48,199 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 6 | 26 | 32 | 32 | 343 | 6 | 34 | 71 | 75 | 18 | 1,427 | 216 | 260 | 214 | 214 | 214 | 288 |
| Pressure (MPa, abs) | 0.09 | 0.11 | 0.86 | 0.86 | 5.10 | 0.09 | 0.55 | 0.09 | 5.62 | 5.62 | 4.24 | 8.27 | 3.89 | 3.756 | 3.76 | 3.76 | 4.14 |
| Enthalpy (kJ/kg) | 15.71 | 30.53 | 26.94 | 26.94 | 3,177.81 | --- | 152.81 | --- | 18.45 | 3,812.95 | --- | 1,017.87 | 1,091.75 | 959.191 | 959.19 | 959.19 | 3,070.04 |
| Density ($kg/m^3$) | 1.1 | 1.3 | 11.0 | 11.0 | 20.1 | --- | 985.2 | --- | 105.5 | 33.9 | --- | 782.0 | 18.6 | 19.8 | 19.8 | 19.8 | 18.2 |
| V-L Molecular Weight | 28.895 | 28.133 | 32.078 | 32.078 | 18.015 | --- | 18.015 | --- | 43.764 | 13.292 | --- | 18.015 | 21.001 | 21.108 | 21.108 | 21.108 | 18.015 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 113,177 | 89,570 | 161 | 22,957 | 6,072 | 0 | 14,702 | 0 | 8,927 | 33,630 | 0 | 27,588 | 115,695 | 111,715 | 14,472 | 97,244 | 34,038 |
| V-L Flowrate (lb/hr) | 3,270,256 | 2,519,876 | 5,166 | 736,415 | 109,392 | 0 | 264,860 | 0 | 390,684 | 447,019 | 0 | 497,007 | 2,429,689 | 2,358,061 | 305,462 | 2,052,598 | 613,203 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 1,259,331 | 0 | 994,470 | 0 | 742,793 | 106,261 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 42 | 78 | 90 | 90 | 650 | 42 | 93 | 160 | 166 | 64 | 2,600 | 420 | 500 | 417 | 417 | 417 | 550 |
| Pressure (psia) | 13.0 | 16.4 | 125.0 | 125.0 | 740.0 | 13.0 | 80.3 | 12.7 | 815.0 | 815.0 | 614.7 | 1,200.0 | 564.7 | 544.7 | 544.7 | 544.7 | 600.0 |
| Enthalpy (Btu/lb) | 6.8 | 13.1 | 11.6 | 11.6 | 1,366.2 | --- | 65.7 | --- | 7.9 | 1,639.3 | --- | 437.6 | 469.4 | 412.4 | 412.4 | 412.4 | 1,319.9 |
| Density ($lb/ft^3$) | 0.070 | 0.080 | 0.685 | 0.685 | 1.257 | --- | 61.501 | --- | 6.587 | 2.116 | --- | 48.817 | 1.162 | 1.238 | 1.238 | 1.238 | 1.135 |
| | A - Reference conditions are 32.02 F & 0.089 PSIA | | | | | | | | | | | | | | | | |

## Exhibit 5-4 Case 6 Stream Table, Subbituminous Rosebud PRB Coal to SNG with Carbon Sequestration (continued)

| | 18 | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | |
| Ar | 0.0021 | 0.0021 | 0.0031 | 0.0031 | 0.0031 | 0.0109 | 0.0000 | 0.0126 | 0.0126 | 0.0001 | 0.0007 | 0.0000 | 0.0016 | 0.0022 | 0.0000 | 0.0000 |
| $CH_4$ | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.8276 | 0.0000 | 0.9590 | 0.9590 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.1606 | 0.1589 | 0.2361 | 0.2361 | 0.2361 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0041 | 0.0594 | 0.0000 | 0.0846 | 0.0074 | 0.0000 | 0.0000 |
| $CO_2$ | 0.3490 | 0.3540 | 0.0406 | 0.0406 | 0.0406 | 0.1367 | 0.9833 | 0.0024 | 0.0024 | 0.9914 | 0.5957 | 0.0000 | 0.4861 | 0.7981 | 0.0000 | 0.0000 |
| COS | 0.0001 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0018 | 0.0000 | 0.0003 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.4810 | 0.4777 | 0.7155 | 0.7155 | 0.7155 | 0.0069 | 0.0000 | 0.0080 | 0.0080 | 0.0043 | 0.1139 | 0.0000 | 0.0552 | 0.1871 | 0.0000 | 0.0000 |
| $H_2O$ | 0.0017 | 0.0017 | 0.0001 | 0.0001 | 0.0001 | 0.0023 | 0.0167 | 0.0000 | 0.0000 | 0.0001 | 0.0410 | 0.0000 | 0.3696 | 0.0009 | 1.0000 | 1.0000 |
| $H_2S$ | 0.0025 | 0.0025 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1869 | 0.0000 | 0.0014 | 0.0033 | 0.0000 | 0.0000 |
| $N_2$ | 0.0030 | 0.0030 | 0.0045 | 0.0045 | 0.0045 | 0.0155 | 0.0000 | 0.0180 | 0.0180 | 0.0000 | 0.0005 | 0.0000 | 0.0007 | 0.0009 | 0.0000 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0006 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 50,513 | 51,082 | 33,890 | 149 | 33,740 | 9,680 | 1,326 | 8,354 | 8,354 | 15,652 | 683 | 0 | 802 | 569 | 43,615 | 60,862 |
| V-L Flowrate (kg/hr) | 1,066,572 | 1,087,014 | 342,095 | 1,508 | 340,587 | 195,732 | 57,763 | 137,969 | 137,969 | 684,975 | 24,177 | 0 | 24,640 | 20,441 | 785,734 | 1,096,442 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4,098 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | |
| Temperature (°C) | 35 | 34 | 31 | 80 | 31 | 35 | 35 | 35 | 35 | 35 | 48 | 175 | 232 | 21 | 566 | 33 |
| Pressure (MPa, abs) | 3.38 | 3.31 | 3.24 | 5.10 | 3.24 | 2.77 | 0.14 | 2.627 | 6.202 | 15.272 | 0.163 | 0.12 | 0.1 | 3.341 | 12.512 | 0.827 |
| Enthalpy (kJ/kg) | 42.63 | 41.26 | 85.32 | 232.24 | 85.32 | 40.72 | 46.40 | 47.945 | 12.775 | -208.887 | 101.201 | --- | 844.6 | -8.013 | 3,513.235 | 138.901 |
| Density ($kg/m^3$) | 28.4 | 28.1 | 12.8 | 17.2 | 12.8 | 23.2 | 2.4 | 17.8 | 44.5 | 778.9 | 2.2 | 5,286.0 | 0.6 | 56.4 | 34.9 | 995.1 |
| V-L Molecular Weight | 21.115 | 21.280 | 10.094 | 10.094 | 10.094 | 20.221 | 43.574 | 16.515 | 16.515 | 43.764 | 35.372 | --- | 31 | 35.955 | 18.015 | 18.015 |
| | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 111,362 | 112,615 | 74,714 | 329 | 74,384 | 21,340 | 2,922 | 18,418 | 18,418 | 34,506 | 1,507 | 0 | 1,768 | 1,253 | 96,154 | 134,177 |
| V-L Flowrate (lb/hr) | 2,351,389 | 2,396,455 | 754,191 | 3,325 | 750,866 | 431,516 | 127,346 | 304,170 | 304,170 | 1,510,111 | 53,301 | 0 | 54,322 | 45,065 | 1,732,248 | 2,417,240 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 9,034 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | |
| Temperature (°F) | 95 | 94 | 87 | 177 | 87 | 95 | 95 | 95 | 95 | 95 | 119 | 347 | 450 | 69 | 1,050 | 91 |
| Pressure (psia) | 489.7 | 479.7 | 469.6 | 740.0 | 469.6 | 401.1 | 20.1 | 381.0 | 899.5 | 2,215.0 | 23.7 | 17.3 | 12.3 | 484.5 | 1,814.7 | 120.0 |
| Enthalpy (Btu/lb) | 18.3 | 17.7 | 36.7 | 99.8 | 36.7 | 17.5 | 19.9 | 20.6 | 5.5 | -89.8 | 43.5 | --- | 363.1 | -3.4 | 1,510.4 | 59.7 |
| Density ($lb/ft^3$) | 1.770 | 1.752 | 0.799 | 1.072 | 0.799 | 1.446 | 0.148 | 1.110 | 2.776 | 48.625 | 0.136 | 329.997 | 0.039 | 3.520 | 2.177 | 62.123 |

### 5.1.2 Key System Assumptions

System assumptions for Cases 5 and 6, Siemens gasifier using Rosebud PRB coal with and without carbon sequestration, are compiled in Exhibit 5-5.

**Exhibit 5-5 Cases 5 and 6 Plant Study Configuration Matrix**

| Case | Case 5 | Case 6 |
|---|---|---|
| Gasifier Pressure, MPa (psia) | 4.2 (615) | 4.2 (615) |
| $O_2$:Coal Ratio, kg $O_2$/kg dry coal | 0.731 | 0.731 |
| Carbon Conversion, % | 99.9 | 99.9 |
| Syngas HHV at Gasifier Outlet, kJ/Nm$^3$ (Btu/scf) | 9,746 (263) | 9,746 (263) |
| Steam Cycle, MPa/°C/°C (psig/°F/°F) | 12.4/566/539 (1800/1050/1002) | 12.4/566/539 (1800/1050/1002) |
| Condenser Pressure, mm Hg (in Hg) | 36 (1.4) | 36 (1.4) |
| CT | N/A | N/A |
| Gasifier Technology | Siemens | Siemens |
| Oxidant | 99 vol% Oxygen | 99 vol% Oxygen |
| Coal | Rosebud PRB | Rosebud PRB |
| Coal Feed Moisture Content, % | 25.77 | 25.77 |
| COS Hydrolysis Reactor | No | No |
| WGS | Yes | Yes |
| $H_2S$ Separation | Selexol (1$^{st}$ Stage) | Selexol (1$^{st}$ Stage) |
| Sulfur Removal, % | 98.5 | 99.9 |
| $CO_2$ Separation | Selexol (2$^{nd}$ Stage) | Selexol (2$^{nd}$ Stage) |
| $CO_2$ Sequestration, % | N/A | 65.8 |
| Sulfur Recovery | Claus Plant with Tail Gas Treatment / Elemental Sulfur | Claus Plant with Tail Gas Treatment / Elemental Sulfur |
| Methanation System | Haldor Topsoe TREMP™ Process | Haldor Topsoe TREMP™ Process |
| Particulate Control | Scrubber and AGR Absorber | Scrubber and AGR Absorber |
| Mercury Control | Carbon Bed | Carbon Bed |
| NOx Control | N/A | N/A |

**Balance of Plant and Sparing Philosophy**

The balance of plant assumptions and sparing philosophy are common to all cases and are presented in Section 4.1.3.

### 5.1.3 Cases 5 and 6 Performance Results

The Siemens SNG plant without and with carbon sequestration and using Rosebud PRB coal produces a net output of 61 Bscf/year at 100 percent capacity.

Overall performance for the two plants is summarized in Exhibit 5-6, which includes auxiliary power requirements and production values. The ASU accounts for approximately 59 percent of the total auxiliary load in non-sequestration case and 49 percent in the sequestration case, distributed between the main air compressor, the oxygen compressor, and ASU auxiliaries. $CO_2$ compression accounts for about 19 percent in the sequestration case. The AGR process accounts for approximately 12 percent and 10 percent of the auxiliary load for the non-sequestration and sequestration cases, respectively. The WTA coal drying process accounts for 8 percent in the non-sequestration case and 7 percent in the sequestration case. All other individual auxiliary loads are less than 3 percent of the total. Both Case 5 and Case 6 are net power producers, but the net power production of Case 6 is very small. This is due to the auxiliary power required for $CO_2$ compression.

**Exhibit 5-6 Cases 5 and 6 Plant Performance Summary**

| **POWER SUMMARY (Gross Power at Generator Terminals, kWe)** | **Case 5** | **Case 6** |
|---|---|---|
| Steam Turbine Power | 302,000 | 302,000 |
| **TOTAL POWER, kWe** | **302,000** | **302,000** |
| **AUXILIARY LOAD SUMMARY, kWe** | | |
| Coal Handling | 840 | 840 |
| Coal Milling | 5,870 | 5,870 |
| Slag Handling | 1,250 | 1,250 |
| WTA Coal Dryer Compressor | 20,050 | 20,060 |
| WTA Coal Dryer Auxiliaries | 880 | 880 |
| Air Separation Unit Auxiliaries | 1,000 | 1,000 |
| Air Separation Unit Main Air Compressor | 124,170 | 124,160 |
| Oxygen Compressor | 21,250 | 21,250 |
| $CO_2$ Compressor | 0 | 56,730 |
| $CO_2$ Solid Feed System Compressor | 5,580 | 270 |
| SNG Compressors | 6,040 | 6,040 |
| Methanation Plant Recycle Compressor | 4,720 | 4,650 |
| Gasifier Pilot Fuel Compressor | 60 | 60 |
| Boiler Feedwater Pumps | 5,020 | 5,020 |
| Condensate Pump | 400 | 400 |
| Quench Water Pump | 570 | 570 |
| Circulating Water Pump | 4,550 | 5,190 |
| Ground Water Pumps | 440 | 500 |
| Cooling Tower Fans | 2,970 | 3,380 |
| Air Cooled Condenser Fans | 3,640 | 3,630 |
| Scrubber Pumps | 980 | 980 |
| Acid Gas Removal | 30,240 | 30,250 |
| Steam Turbine Auxiliaries | 100 | 100 |
| Claus Plant/TGTU Auxiliaries | 250 | 250 |
| Claus Plant TG Recycle Compressor | 2,120 | 2,130 |
| Miscellaneous Balance of Plant[2] | 3,000 | 3,000 |
| Transformer Losses | 1,580 | 1,730 |
| **TOTAL AUXILIARIES, kWe** | **247,570** | **300,190** |
| **NET POWER, kWe** | **54,430** | **1,810** |
| SNG Production Rate, $MNm^3$/hr (Mscf/hr) | 197.9 (6,992) | 197.8 (6,989) |
| Capacity Factor | 90% | 90% |
| Net Exported Power Efficiency (HHV) | 1.7% | 0.1% |
| SNG Conversion Efficiency ($HHV_{product}/HHV_{coal}$), % | 63.1% | 63.1% |
| Coal to SNG Product Yield, $MNm^3_{SNG}/tonne_{coal}$ ($Mscf_{SNG}/ton_{coal}$) | 0.33 (11.1) | 0.33 (11.1) |
| Coal Feed Flow Rate, kg/hr (lb/hr) | 571,223 (1,259,331) | 571,223 (1,259,331) |
| Thermal Input,[1] kWth | 3,160,745 | 3,160,745 |
| Condenser Cooling Duty, GJ/hr (MMBtu/hr) | 1,614 (1,530) | 1,625 (1,540) |
| Raw Water Withdrawal, $m^3$/min (gpm) | 18.4 (4,853) | 20.9 (5,509) |
| Raw Water Consumption, $m^3$/min (gpm) | 14.3 (3,783) | 16.2 (4,291) |

1 - HHV of Rosebud PRB coal is 19,920 kJ/kg (8,564 Btu/lb)

2 - Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

**Environmental Performance**

The environmental targets for emissions of Hg, NOx, $SO_2$, and PM were presented in Section 2.5. A summary of the plant air emissions for Cases 5 and 6 is presented in Exhibit 5-7.

**Exhibit 5-7 Cases 5 and 6 Air Emissions**

| | kg/GJ ($lb/10^6$ Btu) | | Tonne/year (ton/year) 90% capacity factor | |
|---|---|---|---|---|
| | **Case 5** | **Case 6** | **Case 5** | **Case 6** |
| **$SO_2$** | 0.009 (0.022) | 0.000 (0.000) | 840 (926) | 0 (0) |
| **$NO_X$** | Negligible | Negligible | Negligible | Negligible |
| **Particulates** | 0.003 (0.0071) | 0.003 (0.0071) | 274 (302) | 274 (302) |
| **Hg** | 1.51E-7 (3.51E-7) | 1.51E-7 (3.51E-7) | 0.014 (0.015) | 0.014 (0.015) |
| **$CO_2$** | 60.3 (140.3) | 0.3 (0.7) | 5,412,955 (5,966,762) | 27,781 (30,624) |

The low level of $SO_2$ emissions in Case 6 is achieved by capture of the sulfur in the gas by the two-stage Selexol AGR process and co-sequestration with the $CO_2$ product. Because the WTA coal drying process does not include combustion of SNG, there are no $SO_2$ emissions associated with drying. Just as in the non-capture cases, the $SO_2$ emissions are significantly less than the environmental targets of Section 2.5. The clean syngas exiting the AGR process has a sulfur concentration of approximately 2 ppmv in both cases. The $H_2S$-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The tail gas is hydrogenated and recycled to the AGR unit. The higher sulfur emissions in Case 5 are due to the sulfur contained in the $CO_2$-rich stream from the Selexol process, which is vented rather than sequestered.

$NO_X$ emissions are negligible because no CT is used for power generation.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Ninety five percent of mercury is captured from the syngas by an activated carbon bed.

$CO_2$ emissions represent the uncontrolled (Case 5) and controlled (Case 6) discharge from the process. The carbon balance for the two cases is shown in Exhibit 5-8.

The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not used in the carbon capture equation below, but it is not neglected in the balance since Aspen accounts for air components throughout. Carbon leaves the plant as unburned carbon in the slag, ASU vent gas, SNG product, and the $CO_2$ product (either vented or sequestered). The carbon capture efficiency is presented in two distinct ways. The first way defines capture as the amount of carbon in the $CO_2$ product for sequestration relative to the amount of carbon in the coal less carbon contained in the slag, and is represented by the following fraction:

(Carbon in Product for Sequestration)/[(Carbon in the Coal)-(Carbon in Slag)] or

Non sequestration (Case 5)

(412,575)/(630,524-3,153)*100 = 65.8% (Case 6)

**Exhibit 5-8 Cases 5 and 6 Carbon Balance**

| Carbon In, kg/hr (lb/hr) | | | Carbon Out, kg/hr (lb/hr) | | |
|---|---|---|---|---|---|
| | **Case 5** | **Case 6** | | **Case 5** | **Case 6** |
| **Coal** | 286,001 (630,524) | 286,001 (630,524) | **Slag** | 1,430 (3,153) | 1,430 (3,153) |
| **Air ($CO_2$)** | 202 (446) | 202 (446) | **SNG Purification Off-Gas** | 15,627 (34,452) | 0 (0) |
| | | | **ASU Vent** | 202 (446) | 202 (446) |
| | | | **SNG** | 96,555 (212,866) | 96,464 (212,667) |
| | | | **$CO_2$ Vent** | 148,142 (326,598) | 187,141 (412,575) |
| | | | **$CO_2$ Feed Vent** | 24,247 (53,455) | 966 (2,129) |
| **Total** | 286,203 (630,969) | 286,203 (630,969) | **Total** | 286,203 (630,969) | 286,203 (630,969) |

The second way does not penalize the production facility for the carbon converted to SNG product, as the end use of the SNG is unknown. In a carbon constrained scenario, the SNG may or may not be used in a scenario where CCS is implemented. For this method, the burden of carbon mitigation falls on the end-user. This method is represented by the following fraction:

[(Carbon in Product for Sequestration) + (Carbon in SNG Product)]/[(Carbon in the Coal)-(Carbon in Slag)] or

(0 + 212,866)/(630,524 – 3,153)*100 = 33.9% (Case 5)

(412,575 + 212,866)/(630,524-3,153)*100 = 99.7% (Case 6)

Exhibit 5-9 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the stack gas, sulfur co-sequestered with the $CO_2$ product, and sulfur remaining in the SNG product. Sulfur in the slag is considered negligible. The total sulfur capture is represented by the following fraction:

(Sulfur byproduct + Sulfur in $CO_2$ product)/Sulfur in the coal or
(9,028 + 0)/9,161 = 98.5% (Case 5)
(9,033 + 123)/9,161 = 99.9% (Case 6)

**Exhibit 5-9 Cases 5 and 6 Sulfur Balance**

| Sulfur In, kg/hr (lb/hr) | | | Sulfur Out, kg/hr (lb/hr) | | |
|---|---|---|---|---|---|
| | **Case 5** | **Case 6** | | **Case 5** | **Case 6** |
| **Coal** | 4,155 (9,161) | 4,155 (9,161) | **Elemental Sulfur** | 4,095 (9,028) | 4,097 (9,033) |
| | | | **SNG** | 2 (4) | 2 (4) |
| | | | **$CO_2$ Vent Streams/Stack** | 58 (128) | 0.3 (1) |
| | | | **$CO_2$ Product** | N/A | 56 (123) |
| **Total** | 4,155 (9,161) | 4,155 (9,161) | **Total** | 4,155 (9,161) | 4,155 (9,161) |

Exhibit 5-10 shows the overall water balance for the plant. Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as syngas condensate, and is re-used as internal recycle. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is defined as the water removed from the ground or diverted from a surface-water source for use in the plant and was assumed to be provided 50 percent by a POTW and 50 percent from groundwater. Raw water withdrawal can be represented by the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, quench system makeup, and slag handling makeup. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source.

**Exhibit 5-10 Cases 5 and 6 Water Balance**

| | Case 5 | Case 6 |
|---|---|---|
| **Water Demand, $m^3$/min (gpm)** | | |
| **Slag Handling** | **1.04 (276)** | **1.04 (276)** |
| **Quench/Wash** | **3.8 (994)** | **3.8 (994)** |
| **Condenser Makeup** | **5.7 (1,507)** | **5.7 (1,507)** |
| Gasifier Steam | 0.83 (219) | 0.83 (219) |
| Shift Steam | 4.6 (1,226) | 4.6 (1,226) |
| BFW Makeup | 0.23 (62) | 0.23 (62) |
| **Cooling Tower** | **17.7 (4,679)** | **20.2 (5,334)** |
| **Total** | **28.2 (7,456)** | **30.7 (8,111)** |
| **Internal Recycle, $m^3$/min (gpm)** | | |
| **Slag Handling** | **1.04 (276)** | **1.04 (276)** |
| **Quench/Wash** | **3.8 (994)** | **3.8 (994)** |
| **Condenser Makeup** | **0.0 (0)** | **0.0 (0)** |
| **Cooling Tower** | **5.0 (1,332)** | **5.0 (1,332)** |
| Water from Coal Drying | 2.0 (530) | 2.0 (530) |
| BFW Blowdown | 0.23 (62) | 0.23 (62) |
| SWS Blowdown | 0.69 (183) | 0.69 (183) |
| SWS Excess | 2.1 (558) | 2.1 (558) |
| **Total** | **9.9 (2,602)** | **9.8 (2,602)** |
| **Raw Water Withdrawal, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.0 (0)** | **0.0 (0)** |
| **Quench/Wash** | **0.0 (0)** | **0.0 (0)** |
| **Condenser Makeup** | **5.7 (1,507)** | **5.7 (1,507)** |
| Gasifier Steam | 0.83 (219) | 0.83 (219) |
| Shift Steam | 4.6 (1,226) | 4.6 (1,226) |
| BFW Makeup | 0.23 (62) | 0.23 (62) |
| **Cooling Tower** | **12.7 (3,347)** | **15.1 (4,002)** |
| **Total** | **18.4 (4,853)** | **20.9 (5,509)** |
| **Process Water Discharge, $m^3$/min (gpm)** | | |
| **SWS Blowdown** | **0.07 (18)** | **0.07 (18)** |
| **Cooling Tower Blowdown** | **4.0 (1,052)** | **4.5 (1,200)** |
| **Total** | **4.1 (1,071)** | **4.6 (1,218)** |
| **Raw Water Consumption, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.0 (0)** | **0.0 (0)** |
| **Quench/Wash** | **0.0 (0)** | **0.0 (0)** |
| **SWS Blowdown** | **-0.07 (-18)** | **-0.07 (-18)** |
| **Condenser Makeup** | **5.7 (1,507)** | **5.7 (1,507)** |
| **Cooling Tower** | **8.7 (2,295)** | **10.6 (2,803)** |
| **Total** | **14.3 (3,783)** | **16.2 (4,291)** |

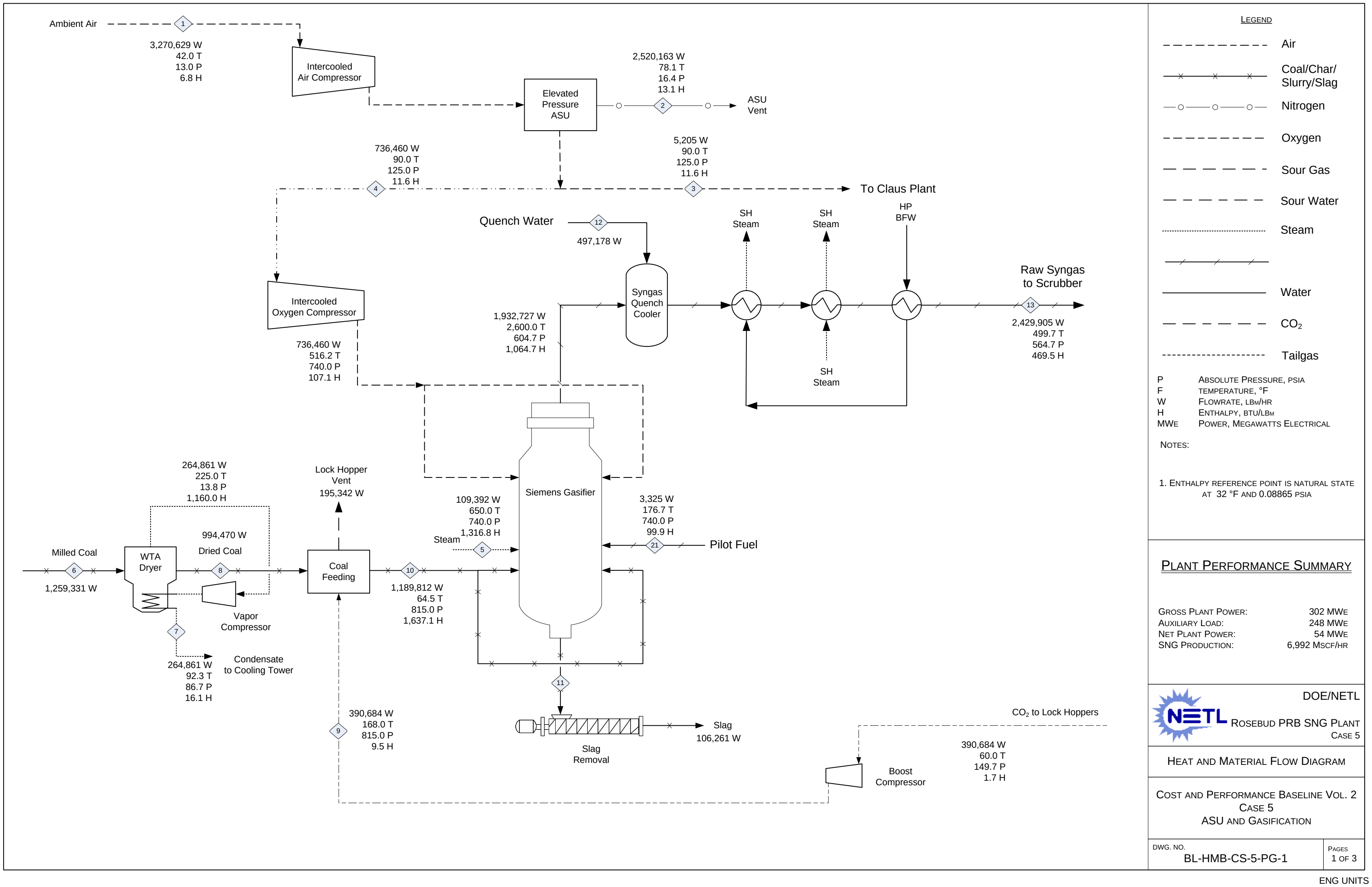
## Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 5-11 through Exhibit 5-16:

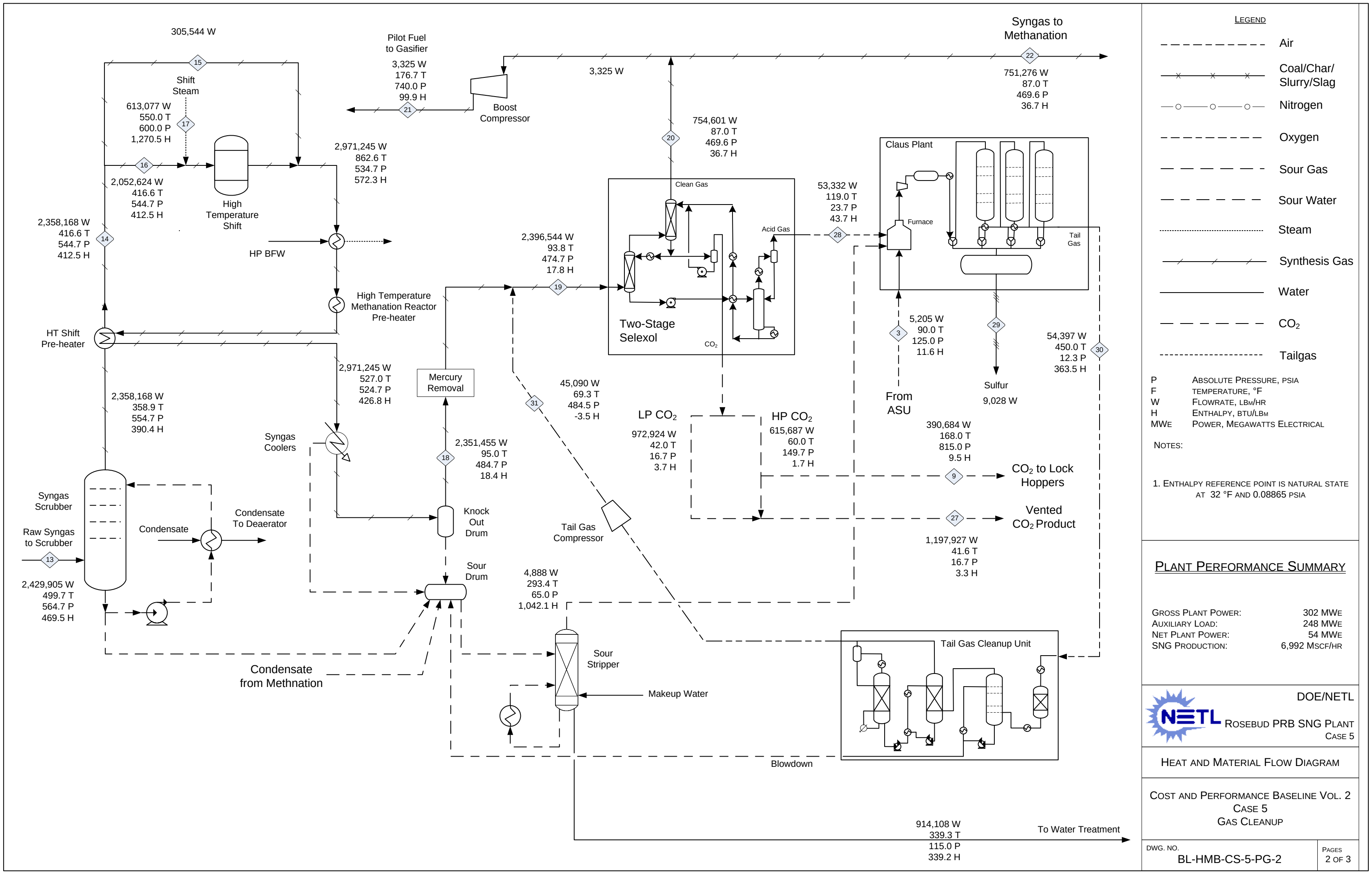
- Coal gasification and ASU
- Syngas cleanup (including sulfur recovery and tail gas recycle)
- Methanation, SNG purification, SNG compression, and power cycle

An overall plant energy balance is provided in tabular form in Exhibit 5-17 for the two cases.

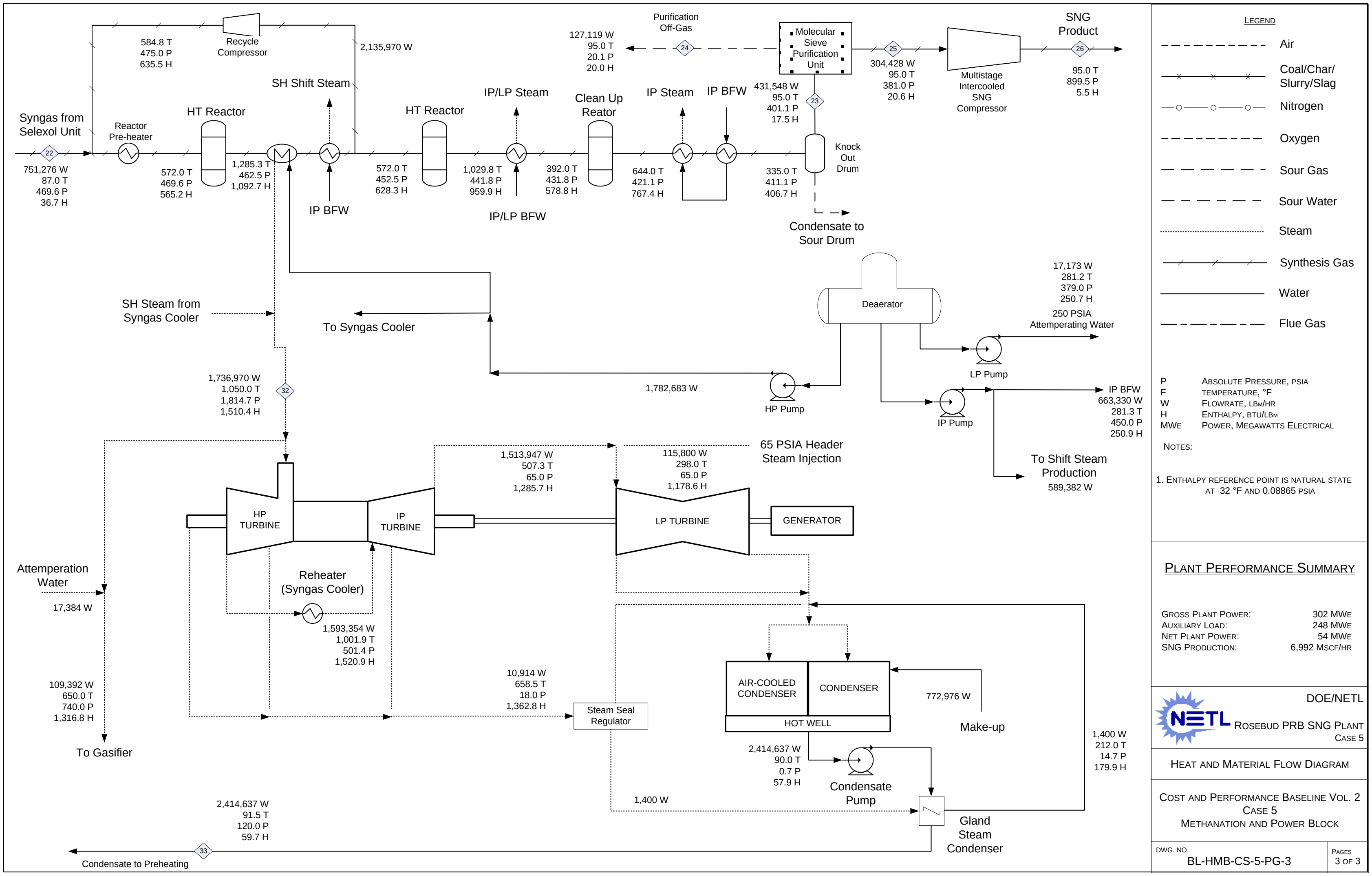
# Exhibit 5-11 Case 5 Gasification and ASU Heat and Mass Balance



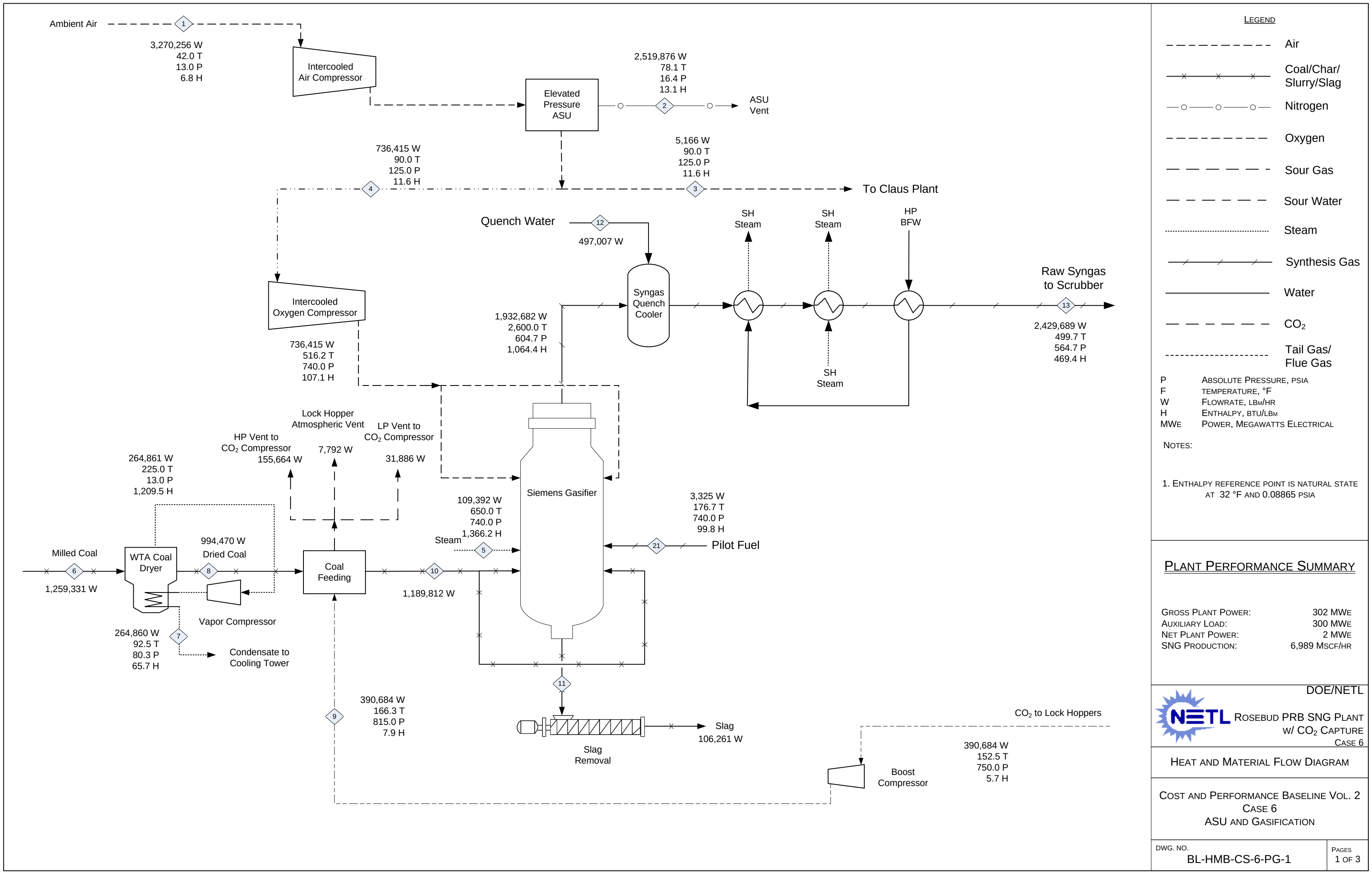
## Exhibit 5-12 Case 5 Gas Cleanup Heat and Mass Balance



# Exhibit 5-13 Case 5 Methanation and Power Block Heat and Mass Balance

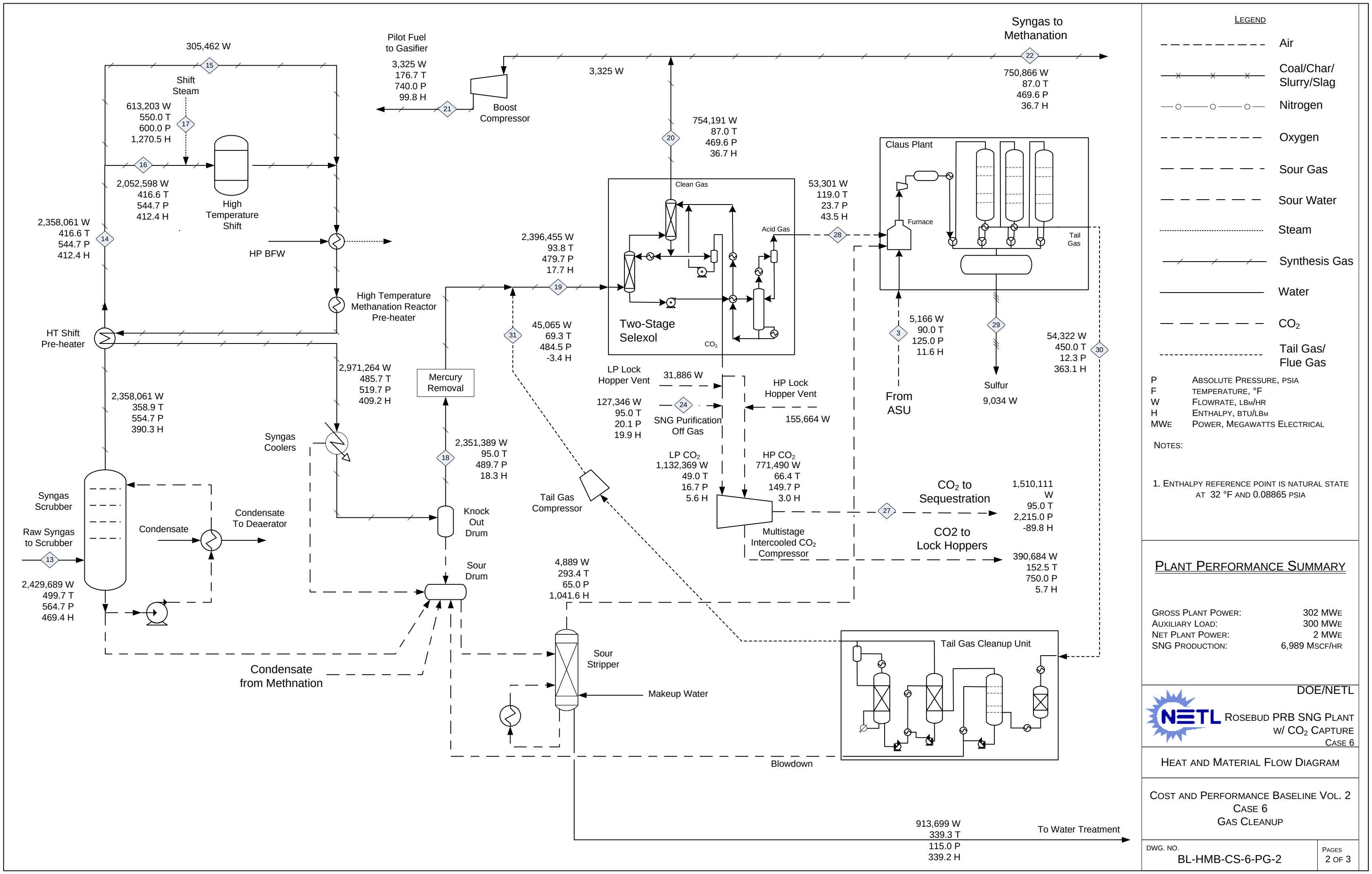


# Exhibit 5-14 Case 6 Gasification and ASU Heat and Mass Balance




ENG UNITS

# Exhibit 5-15 Case 6 Gas Cleanup Heat and Mass Balance



# Exhibit 5-16 Case 6 Methanation and Power Block Heat and Mass Balance

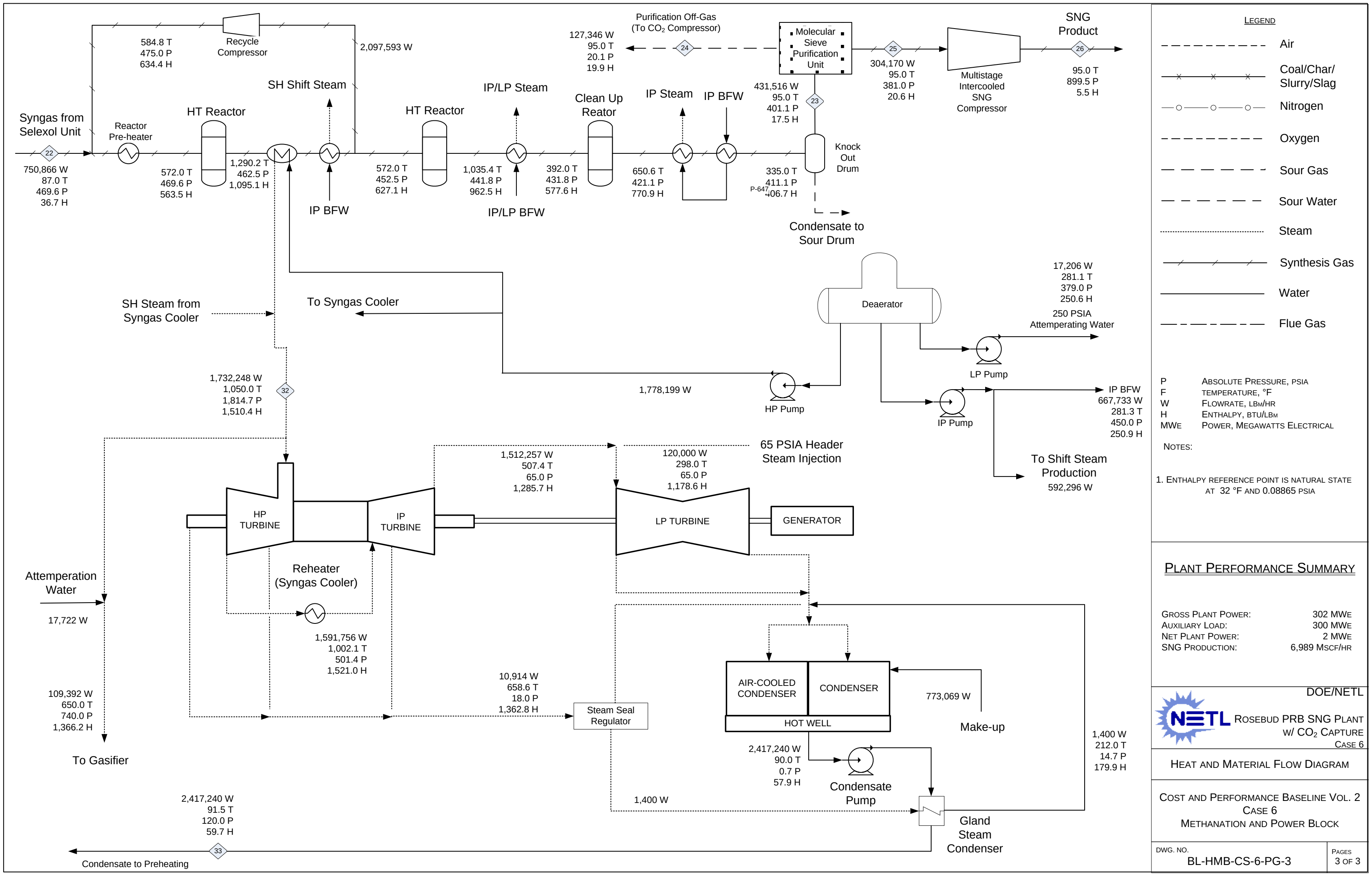


**Exhibit 5-17 Cases 5 and 6 Energy Balance**

| | HHV | | Sensible + Latent | | Power | | Total | |
|---|---|---|---|---|---|---|---|---|
| | Case 5 | Case 6 | Case 5 | Case 6 | Case 5 | Case 6 | Case 5 | Case 6 |
| **Heat In, GJ/hr (MMBtu/hr)** | | | | | | | | |
| Coal | 11,379 (10,785) | 11,379 (10,785) | 5.8 (5.5) | 5.8 (5.5) | 0 (0) | 0 (0) | 11,385 (10,790) | 11,385 (10,790) |
| ASU Air | 0 (0) | 0 (0) | 23.3 (22.1) | 23.3 (22.1) | 0 (0) | 0 (0) | 23 (22) | 23 (22) |
| Raw Water Makeup | 0 (0) | 0 (0) | 25.6 (24.2) | 29.0 (27.5) | 0 (0) | 0 (0) | 26 (24) | 29 (27) |
| Auxiliary Power | 0 (0) | 0 (0) | 0.0 (0.0) | 0.0 (0.0) | 891 (845) | 1,081 (1,024) | 891 (845) | 1,081 (1,024) |
| **Totals** | **11,379 (10,785)** | **11,379 (10,785)** | **54.7 (51.9)** | **58.2 (55.1)** | **891 (845)** | **1,081 (1,024)** | **12,325 (11,682)** | **12,518 (11,864)** |
| **Heat Out, GJ/hr (MMBtu/hr)** | | | | | | | | |
| ASU Intercoolers | 0 (0) | 0 (0) | 325 (308) | 325 (308) | 0 (0) | 0 (0) | 325 (308) | 325 (308) |
| ASU Vent | 0 (0) | 0 (0) | 34.9 (33.1) | 34.9 (33.1) | 0 (0) | 0 (0) | 35 (33) | 35 (33) |
| Slag | 47 (44) | 47 (44) | 81.4 (77.1) | 81.4 (77.1) | 0 (0) | 0 (0) | 128 (122) | 128 (122) |
| Sulfur | 38 (36) | 38 (36) | 0.5 (0.4) | 0.5 (0.4) | 0 (0) | 0 (0) | 38 (36) | 38 (36) |
| $CO_2$ Intercoolers | 0 (0) | 0 (0) | 0.0 (0.0) | 343.7 (325.8) | 0 (0) | 0 (0) | 0 (0) | 344 (326) |
| $CO_2$ | 0 (0) | 0 (0) | 2.0 (1.9) | -143.1 (-135.6) | 0 (0) | 0 (0) | 2 (2) | -143 (-136) |
| Cooling Tower Blowdown | 0 (0) | 0 (0) | 22.2 (21.0) | 25.3 (24.0) | 0 (0) | 0 (0) | 22 (21) | 25 (24) |
| SNG | 7,163 (6,789) | 7,158 (6,784) | 1.8 (1.7) | 1.8 (1.7) | 0 (0) | 0 (0) | 7,165 (6,791) | 7,160 (6,786) |
| SNG Purification Off-Gas | 0 (0) | 0 (0) | 3 (3) | 0 (0) | 0 (0) | 0 (0) | 3 (3) | 0 (0) |
| Condenser | 0 (0) | 0 (0) | 1,618 (1,533) | 1,620 (1,536) | 0 (0) | 0 (0) | 1,618 (1,533) | 1,620 (1,536) |
| Non-Condenser Cooling Tower Loads* | 0 (0) | 0 (0) | 902 (855) | 1,224 (1,160) | 0 (0) | 0 (0) | 902 (855) | 1,224 (1,160) |
| Process Losses* | 0 (0) | 0 (0) | 1,000 (947) | 674 (639) | 0 (0) | 0 (0) | 1,000 (947) | 674 (639) |
| Power | 0 (0) | 0 (0) | 0.0 (0.0) | 0.0 (0.0) | 1,087 (1,030) | 1,087 (1,030) | 1,087 (1,030) | 1,087 (1,030) |
| **Totals** | **7,248 (6,870)** | **7,243 (6,865)** | **3,989 (3,781)** | **4,188 (3,969)** | **1,087 (1,030)** | **1,087 (1,030)** | **12,325 (11,682)** | **12,518 (11,864)** |

* Includes other energy losses not explicitly accounted for in the model

### 5.1.4 Cases 5 and 6 Equipment Lists

Major equipment items for the Siemens gasifier SNG plant without and with carbon sequestration using PRB coal are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 5.1.5. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

**ACCOUNT 1 COAL HANDLING**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Bottom Trestle Dumper and Receiving Hoppers | N/A | 2 | 0 | 181 tonne (200 ton) | 181 tonne (200 ton) |
| 2 | Feeder | Belt | 2 | 0 | 572 tonne/hr (630 tph) | 572 tonne/hr (630 tph) |
| 3 | Conveyor No. 1 | Belt | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 4 | Transfer Tower No. 1 | Enclosed | 1 | 0 | N/A | N/A |
| 5 | Conveyor No. 2 | Belt | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 6 | As-Received Coal Sampling System | Two-stage | 1 | 0 | N/A | N/A |
| 7 | Stacker/Reclaimer | Traveling, linear | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 8 | Reclaim Hopper | N/A | 2 | 1 | 118 tonne (130 ton) | 118 tonne (130 ton) |
| 9 | Feeder | Vibratory | 2 | 1 | 472 tonne/hr (520 tph) | 472 tonne/hr (520 tph) |
| 10 | Conveyor No. 3 | Belt w/ tripper | 1 | 0 | 943 tonne/hr (1,040 tph) | 943 tonne/hr (1,040 tph) |
| 11 | Crusher Tower | N/A | 1 | 0 | N/A | N/A |
| 12 | Coal Surge Bin w/ Vent Filter | Dual outlet | 2 | 0 | 472 tonne (520 ton) | 472 tonne (520 ton) |
| 13 | Crusher | Impactor reduction | 2 | 0 | 8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0) | 8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0) |
| 14 | As-Fired Coal Sampling System | Swing hammer | 1 | 1 | N/A | N/A |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 15 | Conveyor No. 4 | Belt w/tripper | 1 | 0 | 943 tonne/hr (1,040 tph) | 943 tonne/hr (1,040 tph) |
| 16 | Transfer Tower No. 2 | Enclosed | 1 | 0 | N/A | N/A |
| 17 | Conveyor No. 5 | Belt w/ tripper | 1 | 0 | 943 tonne/hr (1,040 tph) | 943 tonne/hr (1,040 tph) |
| 18 | Coal Silo w/ Vent Filter and Slide Gates | Field erected | 3 | 0 | 2,087 tonne (2,300 ton) | 2,087 tonne (2,300 ton) |

**ACCOUNT 2 COAL PREPARATION AND FEED**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Feeder | Vibratory | 6 | 0 | 109 tonne/hr (120 tph) | 109 tonne/hr (120 tph) |
| 2 | Conveyor No. 6 | Belt w/tripper | 3 | 0 | 209 tonne/hr (230 tph) | 209 tonne/hr (230 tph) |
| 3 | Roller Mill Feed Hopper | Dual Outlet | 3 | 0 | 417 tonne (460 ton) | 417 tonne (460 ton) |
| 4 | Weigh Feeder | Belt | 6 | 0 | 109 tonne/hr (120 tph) | 109 tonne/hr (120 tph) |
| 5 | Pulverizer | Rotary | 6 | 0 | 109 tonne/hr (120 tph) | 109 tonne/hr (120 tph) |
| 6 | Coal Dryer Feed Hopper | Vertical Hopper | 3 | 0 | 417 tonne (460 ton) | 417 tonne (460 ton) |
| 7 | Coal Preheater | Water Heated Horizontal Rotary Kiln | 3 | 0 | Coal feed: 209 tonne/hr (230 tph) Heat duty: 23.3 GJ/hr (22.1 MMBtu/hr) | Coal feed: 209 tonne/hr (230 tph) Heat duty: 23.3 GJ/hr (22.1 MMBtu/hr) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 8 | Coal Dryer | Fluidized Bed with Internal Coils | 3 | 0 | Coal feed: 209 tonne/hr (230 tph) Heat duty: 117 GJ/hr (111 MMBtu/hr) Bed diameter: 14.9 m (49 ft) | Coal feed: 209 tonne/hr (230 tph) Heat duty: 117 GJ/hr (111 MMBtu/hr) Bed diameter: 14.9 m (49 ft) |
| 9 | Steam Compressor | Reciprocating, Multi-Stage | 3 | 0 | 878 $m^3$/min (30,990 scfm) Suction - 0.10 MPa (13.8 psia) Discharge - 0.77 MPa (112 psia) | 878 $m^3$/min (30,990 scfm) Suction - 0.09 MPa (13 psia) Discharge - 0.73 MPa (105 psia) |
| 10 | Dryer Exhaust Filter | Hot Baghouse | 3 | 0 | Steam - 44,044 kg/hr (97,100 lb/hr) Temperature - 107°C (225°F) | Steam - 44,044 kg/hr (97,100 lb/hr) Temperature - 107°C (225°F) |
| 11 | Dry Coal Cooler | Water Cooled Horizontal Rotary Kiln | 3 | 0 | 165 tonne/hr (182 tph) Heat duty - 10 GJ/hr (9 MMBtu/hr) | 165 tonne/hr (182 tph) Heat duty - 10 GJ/hr 9 MMBtu/hr) |

**ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Demineralized Water Storage Tank | Vertical, cylindrical, outdoor | 3 | 0 | 605,666 liters (160,000 gal) | 605,666 liters (160,000 gal) |
| 2 | Condensate Pumps | Vertical canned | 2 | 1 | 10,107 lpm @ 91 m $H_2O$ (2,670 gpm @ 300 ft $H_2O$) | 10,107 lpm @ 91 m $H_2O$ (2,670 gpm @ 300 ft $H_2O$) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 3 | Deaerator | Horizontal spray type | 2 | 0 | 769,293 kg/hr (1,696,000 lb/hr) | 770,200 kg/hr (1,698,000 lb/hr) |
| 4 | Intermediate Pressure Feedwater Pump | Horizontal centrifugal, single stage | 2 | 1 | 5,413 lpm @ 341 m $H_2O$ (1,430 gpm @ 1,120 ft $H_2O$) | 5,451 lpm @ 341 m $H_2O$ (1,440 gpm @ 1,120 ft $H_2O$) |
| 5 | High Pressure Feedwater Pump | Barrel type, multi-stage, centrifugal | 2 | 1 | HP water: 7,722 lpm @ 1,676 m $H_2O$ (2,040 gpm @ 5,500 ft $H_2O$) | HP water: 7,684 lpm @ 1,676 m $H_2O$ (2,030 gpm @ 5,500 ft $H_2O$) |
| 6 | Auxiliary Boiler | Shop fabricated, water tube | 1 | 0 | 18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F) | 18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F) |
| 7 | Service Air Compressors | Flooded Screw | 3 | 1 | 28 $m^3$/min @ 0.7 MPa (1,000 scfm @ 100 psig) | 28 $m^3$/min @ 0.7 MPa (1,000 scfm @ 100 psig) |
| 8 | Instrument Air Dryers | Duplex, regenerative | 3 | 1 | 28 $m^3$/min (1,000 scfm) | 28 $m^3$/min (1,000 scfm) |
| 9 | Closed Cycle Cooling Heat Exchangers | Plate and frame | 2 | 0 | 675 GJ/hr (639 MMBtu/hr) each | 852 GJ/hr (808 MMBtu/hr) each |
| 10 | Closed Cycle Cooling Water Pumps | Horizontal centrifugal | 2 | 1 | 241,888 lpm @ 21 m $H_2O$ (63,900 gpm @ 70 ft $H_2O$) | 305,483 lpm @ 21 m $H_2O$ (80,700 gpm @ 70 ft $H_2O$) |
| 11 | Engine-Driven Fire Pump | Vertical turbine, diesel engine | 1 | 1 | 3,785 lpm @ 107 m $H_2O$ (1,000 gpm @ 350 ft $H_2O$) | 3,785 lpm @ 107 m $H_2O$ (1,000 gpm @ 350 ft $H_2O$) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 12 | Fire Service Booster Pump | Two-stage horizontal centrifugal | 1 | 1 | 2,650 lpm @ 76 m $H_2O$ (700 gpm @ 250 ft $H_2O$) | 2,650 lpm @ 76 m $H_2O$ (700 gpm @ 250 ft $H_2O$) |
| 13 | Raw Water Pumps | Stainless steel, single suction | 2 | 1 | 5,110 lpm @ 18 m $H_2O$ (1,350 gpm @ 60 ft $H_2O$) | 5,792 lpm @ 18 m $H_2O$ (1,530 gpm @ 60 ft $H_2O$) |
| 14 | Ground Water Pumps | Stainless steel, single suction | 3 | 1 | 3,407 lpm @ 268 m $H_2O$ (900 gpm @ 880 ft $H_2O$) | 3,861 lpm @ 268 m $H_2O$ (1,020 gpm @ 880 ft $H_2O$) |
| 15 | Filtered Water Pumps | Stainless steel, single suction | 2 | 1 | 5,905 lpm @ 49 m $H_2O$ (1,560 gpm @ 160 ft $H_2O$) | 5,905 lpm @ 49 m $H_2O$ (1,560 gpm @ 160 ft $H_2O$) |
| 16 | Filtered Water Tank | Vertical, cylindrical | 2 | 0 | 2,831,488 liter (748,000 gal) | 2,831,488 liter (748,000 gal) |
| 17 | Makeup Water Demineralizer | Anion, cation, and mixed bed | 4 | 0 | 1,401 lpm (370 gpm) | 1,401 lpm (370 gpm) |
| 18 | Liquid Waste Treatment System | | 1 | 0 | 10 years, 24-hour storm | 10 years, 24-hour storm |

**ACCOUNT 4 GASIFIER, ASU, AND ACCESSORIES INCLUDING LOW TEMPERATURE HEAT RECOVERY**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Gasifier | Pressurized dry-feed, entrained bed | 6 | 0 | 2,540 tonne/day, 4.2 MPa (2,800 tpd, 615 psia) | 2,540 tonne/day, 4.2 MPa (2,800 tpd, 615 psia) |
| 2 | Synthesis Gas Cooler | Convective spiral-wound tube boiler | 6 | 0 | 201,849 kg/hr (445,000 lb/hr) | 201,849 kg/hr (445,000 lb/hr) |
| 3 | Syngas Scrubber Including Sour Water Stripper | Vertical upflow | 6 | 0 | 201,849 kg/hr (445,000 lb/hr) | 201,849 kg/hr (445,000 lb/hr) |
| 4 | Raw Gas Coolers | Shell and tube with condensate drain | 12 | 0 | 423,655 kg/hr (934,000 lb/hr) | 423,655 kg/hr (934,000 lb/hr) |
| 5 | Raw Gas Knockout Drum | Vertical with mist eliminator | 3 | 0 | 390,997 kg/hr, 35°C, 3.4 MPa (862,000 lb/hr, 95°F, 490 psia) | 390,997 kg/hr, 35°C, 3.4 MPa (862,000 lb/hr, 95°F, 495 psia) |
| 6 | Flare Stack | Self-supporting, carbon steel, stainless steel top, pilot ignition | 3 | 0 | 404,151 kg/hr (891,000 lb/hr) syngas | 404,151 kg/hr (891,000 lb/hr) syngas |
| 7 | ASU Main Air Compressor | Centrifugal, multi-stage | 3 | 0 | 7,419 $m^3$/min @ 1.3 MPa (262,000 scfm @ 190 psia) | 7,419 $m^3$/min @ 1.3 MPa (262,000 scfm @ 190 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 8 | Cold Box | Vendor design | 3 | 0 | 2,994 tonne/day (3,300 tpd) of 99% purity oxygen | 2,994 tonne/day (3,300 tpd) of 99% purity oxygen |
| 9 | Oxygen Compressor | Centrifugal, multi-stage | 3 | 0 | 1,501 $m^3$/min (53,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia) | 1,501 $m^3$/min (53,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia) |
| 10 | $CO_2$ Solid Feed System Compressor | Centrifugal, multi-stage | 3 | 0 | 595 $m^3$/min (21,000 scfm) Suction - 1.0 MPa (150 psia) Discharge - 5.7 MPa (820 psia) | 595 $m^3$/min (21,000 scfm) Suction - 5.2 MPa (750 psia) Discharge - 5.7 MPa (820 psia) |

**ACCOUNT 5A SYNGAS CLEANUP**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Mercury Adsorber | Sulfated carbon bed | 3 | 0 | 390,997 kg/hr (862,000 lb/hr) 35°C (95°F) 3.3 MPa (485 psia) | 390,997 kg/hr (862,000 lb/hr) 35°C (95°F) 3.4 MPa (490 psia) |
| 2 | ZnO Guard Bed | Fixed bed | | | 444 $Nm^3$/min (15,684 acfm) | 444 $Nm^3$/min (15,672 acfm) |
| 3 | Sulfur Plant | Claus type | 2 | 0 | 54 tonne/day (60 tpd) | 54 tonne/day (60 tpd) |
| 4 | Water Gas Shift Reactor | Fixed bed, catalytic | 3 | 0 | 443,160 kg/hr (977,000 lb/hr) 232°C (450°F) 3.7 MPa (540 psia) | 443,160 kg/hr (977,000 lb/hr) 232°C (450°F) 3.7 MPa (540 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 5 | Shift Reactor Heat Recovery Exchanger | Shell and Tube | 3 each | 0 | Exchanger 1: 62 GJ/hr (58 MMBtu/hr) Exchanger 2: 82 GJ/hr (78 MMBtu/hr) Exchanger 3: 17 GJ/hr (16 MMBtu/hr) | Exchanger 1: 62 GJ/hr (58 MMBtu/hr) Exchanger 2: 82 GJ/hr (78 MMBtu/hr) Exchanger 3: 17 GJ/hr (16 MMBtu/hr) |
| 6 | Acid Gas Removal Plant | Two-stage Selexol | 3 | 0 | 398,708 kg/hr (879,000 lb/hr) 34°C (94°F) 3.3 MPa (475 psia) | 398,708 kg/hr (879,000 lb/hr) 34°C (94°F) 3.3 MPa (480 psia) |
| 7 | Hydrogenati on Reactor | Fixed bed, catalytic | 2 | 0 | 13,571 kg/hr (29,919 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia) | 13,555 kg/hr (29,883 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia) |
| 8 | Tail Gas Recycle Compressor | Centrifugal | 2 | 0 | 11,284 kg/hr (24,878 lb/hr) | 11,279 kg/hr (24,865 lb/hr) |

**ACCOUNT 5B $CO_2$ COMPRESSION**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | $CO_2$ Compressor | Integrally geared, multi-stage centrifugal | 4 | 0 | N/A | 1,699 m3/min @ 15.3 MPa (60,000 scfm @ 2,215 psia) |

## ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Combustion Turbine | N/A | N/A | N/A | N/A | N/A |
| 2 | Combustion Turbine Generator | N/A | N/A | N/A | N/A | N/A |

## ACCOUNT 7 DUCTING & STACK

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Stack | CS plate, type 409SS liner | 1 | 0 | 76 m (250 ft) high x 2.1 m (7 ft) diameter | 76 m (250 ft) high x 0.6 m (2 ft) diameter |

## ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Steam Turbine | Commercially available | 1 | 0 | 318 MW 12.4 MPa/ /566°C (1800 psig/ 1050°F) | 318 MW 12.4 MPa/ 566°C (1800 psig/ 1050°F) |
| 2 | Steam Turbine Generator | Hydrogen cooled, static excitation | 1 | 0 | 350 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase | 350 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 3 | Surface Condenser | Single pass, divided waterbox including vacuum pumps | 1 | 0 | 886 GJ/hr (840 MMBtu/hr), Condensing temperature 32°C (90°F), I nlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F) | 886 GJ/hr (840 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F) |
| 4 | Air-cooled Condenser | --- | 1 | 0 | 886 GJ/hr (840 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F) | 886 GJ/hr (840 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F) |

**ACCOUNT 9 COOLING WATER SYSTEM**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Circulating Water Pumps | Vertical, wet pit | 2 | 1 | 454,249 lpm @ 30 m (120,000 gpm @ 100 ft) | 518,601 lpm @ 30 m (137,000 gpm @ 100 ft) |
| 2 | Cooling Tower | Evaporative, mechanical draft, multi-cell | 1 | 0 | 3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 2,543 GJ/hr (2,410 MMBtu/hr) heat duty | 3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 2,901 GJ/hr (2,750 MMBtu/hr) heat duty |

## ACCOUNT 10 SLAG RECOVERY AND HANDLING

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Slag Quench Tank | Water bath | 6 | 0 | 170,344 liters (45,000 gal) | 170,344 liters (45,000 gal) |
| 2 | Slag Crusher | Roll | 6 | 0 | 9 tonne/hr (10 tph) | 9 tonne/hr (10 tph) |
| 3 | Slag Depressurizer | Proprietary | 6 | 0 | 9 tonne/hr (10 tph) | 9 tonne/hr (10 tph) |
| 4 | Slag Receiving Tank | Horizontal, weir | 6 | 0 | 102,206 liters (27,000 gal) | 102,206 liters (27,000 gal) |
| 5 | Black Water Overflow Tank | Shop fabricated | 6 | 0 | 45,425 liters (12,000 gal) | 45,425 liters (12,000 gal) |
| 6 | Slag Conveyor | Drag chain | 6 | 0 | 9 tonne/hr (10 tph) | 9 tonne/hr (10 tph) |
| 7 | Slag Separation Screen | Vibrating | 6 | 0 | 9 tonne/hr (10 tph) | 9 tonne/hr (10 tph) |
| 8 | Coarse Slag Conveyor | Belt/bucket | 6 | 0 | 9 tonne/hr (10 tph) | 9 tonne/hr (10 tph) |
| 9 | Fine Ash Settling Tank | Vertical, gravity | 6 | 0 | 143,846 liters (38,000 gal) | 143,846 liters (38,000 gal) |
| 10 | Fine Ash Recycle Pumps | Horizontal centrifugal | 6 | 3 | 38 lpm @ 14 m $H_2O$ (10 gpm @ 46 ft $H_2O$) | 38 lpm @ 14 m $H_2O$ (10 gpm @ 46 ft $H_2O$) |
| 11 | Grey Water Storage Tank | Field erected | 6 | 0 | 45,425 liters (12,000 gal) | 45,425 liters (12,000 gal) |
| 12 | Grey Water Pumps | Centrifugal | 6 | 3 | 151 lpm @ 433 m $H_2O$ (40 gpm @ 1,420 ft $H_2O$) | 151 lpm @ 433 m $H_2O$ (40 gpm @ 1,420 ft $H_2O$) |
| 13 | Slag Storage Bin | Vertical, field erected | 6 | 0 | 635 tonne (700 tons) | 635 tonne (700 tons) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 14 | Unloading Equipment | Telescoping chute | 2 | 0 | 109 tonne/hr (120 tph) | 109 tonne/hr (120 tph) |

**ACCOUNT 11 SNG PRODUCTION/METHANATION SYSTEM**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Methanation Reactor Preheater | Shell and Tube | 3 | 0 | 95 GJ/hr (90 MMBtu/hr) | 95 GJ/hr (90 MMBtu/hr) |
| 2 | Methanation Reactor 1 | Fixed Bed, catalytic | 3 | 0 | 480,354 kg/hr (1,059,000 lb/hr)<br>699°C (1,290°F)<br>3.2 MPa (470 psia) | 474,004 kg/hr (1,045,000 lb/hr)<br>699°C (1,290°F)<br>3.2 MPa (470 psia) |
| 3 | Methanation Reactor Superheater 1 | Shell and Tube | 3 | 0 | 433 GJ/hr (410 MMBtu/hr) | 431 GJ/hr (408 MMBtu/hr) |
| 4 | Methanation Reactor Intercooler 2 | Shell and Tube | 3 | 0 | 86 GJ/hr (82 MMBtu/hr) | 85 GJ/hr (81 MMBtu/hr) |
| 5 | Methanation Reactor 2 | Fixed Bed, catalytic | 3 | 0 | 124,738 kg/hr (275,000 lb/hr)<br>554°C (1,030°F)<br>3.1 MPa ( psia) | 124,738 kg/hr (275,000 lb/hr)<br>560°C (1,040°F)<br>3.1 MPa ( psia) |
| 6 | Methanation Reactor Intercooler 3 | Shell and Tube | 3 | 0 | 29 GJ/hr (28 MMBtu/hr) | 29 GJ/hr (28 MMBtu/hr) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 7 | Methanation Reactor 3 | Fixed Bed, catalytic | 3 | 0 | 124,738 kg/hr (275,000 lb/hr) 338°C (640°F) 3.0 MPa (440 psia) | 124,738 kg/hr (275,000 lb/hr) 343°C (650°F) 3.0 MPa (440 psia) |
| 8 | Methanation Reactor Intercooler 4 | Shell and Tube | 3 | 0 | 39 GJ/hr (37 MMBtu/hr) | 40 GJ/hr (38 MMBtu/hr) |
| 9 | Methanation Reactor Intercooler 5 | Shell and Tube | 3 | 0 | 66 GJ/hr (62 MMBtu/hr) | 66 GJ/hr (62 MMBtu/hr) |
| 10 | SNG Purification Condenser 1 | Shell and Tube | 3 | 0 | 54 GJ/hr (51 MMBtu/hr) | 54 GJ/hr (51 MMBtu/hr) |
| 11 | SNG Purification Condenser 2 | Shell and Tube | 3 | 0 | 33 GJ/hr (31 MMBtu/hr) | 33 GJ/hr (31 MMBtu/hr) |
| 12 | Methanation Recycle Compressor | Centrifugal | 3 | 1 | 8,634 $m^3$/min (304,900 scfm) Suction - 3.1 MPa (453 psia) Discharge - 3.3 MPa 475 psia) | 8,506 $m^3$/min (300,400 scfm) Suction - 3.1 MPa (453 psia) Discharge - 3.3 MPa (475 psia) |
| 13 | Molecular Sieve Purification Reactor | Fixed bed | 6 | 0 | 35,834 kg/hr (79,000 lb/hr) 35°C (95°F) 2.8 MPa (401 psia) | 35,834 kg/hr (79,000 lb/hr) 35°C (95°F) 2.8 MPa (401 psia) |
| 14 | SNG Product Compressor | Centrifugal, Multi-staged | 3 | 1 | 1,209 $m^3$/min (42,700 scfm) 2.6 MPa (381 psia) 6.2 MPa (900 psia) | 1,209 $m^3$/min (42,700 scfm) 2.6 MPa (381 psia) 6.2 MPa (900 psia) |

**ACCOUNT 12 ACCESSORY ELECTRIC PLANT**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | STG Step-up Transformer | Oil-filled | 1 | 0 | 24 kV/345 kV, 350 MVA, 3-ph, 60 Hz | 345 kV/24 kV, 350 MVA, 3-ph, 60 Hz |
| 2 | Auxiliary Transformer | Oil-filled | 1 | 1 | 24 kV/4.16 kV, 273 MVA, 3-ph, 60 Hz | 24 kV/4.16 kV, 332 MVA, 3-ph, 60 Hz |
| 3 | Low Voltage Transformer | Dry ventilated | 1 | 1 | 4.16 kV/480 V, 41 MVA, 3-ph, 60 Hz | 4.16 kV/480 V, 50 MVA, 3-ph, 60 Hz |
| 4 | STG Isolated Phase Bus Duct and Tap Bus | Aluminum, self-cooled | 1 | 0 | 24 kV, 3-ph, 60 Hz | 24 kV, 3-ph, 60 Hz |
| 5 | Medium Voltage Switchgear | Metal clad | 1 | 1 | 4.16 kV, 3-ph, 60 Hz | 4.16 kV, 3-ph, 60 Hz |
| 6 | Low Voltage Switchgear | Metal enclosed | 1 | 1 | 480 V, 3-ph, 60 Hz | 480 V, 3-ph, 60 Hz |
| 7 | Emergency Diesel Generator | Sized for emergency shutdown | 1 | 0 | 750 kW, 480 V, 3-ph, 60 Hz | 750 kW, 480 V, 3-ph, 60 Hz |

## ACCOUNT 13 INSTRUMENTATION AND CONTROLS

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 5 Design Condition | Case 6 Design Condition |
|---|---|---|---|---|---|---|
| 1 | DCS - Main Control | Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W) | 1 | 0 | Operator stations/printers and engineering stations/printers | Operator stations/printers and engineering stations/printers |
| 2 | DCS - Processor | Microprocessor with redundant input/output | 1 | 0 | N/A | N/A |
| 3 | DCS - Data Highway | Fiber optic | 1 | 0 | Fully redundant, 25% spare | Fully redundant, 25% spare |

### 5.1.5 Cases 5 and 6 Cost Estimating

The cost estimating methodology was described previously in Section 2.8.

The TPC organized by cost account; owner's costs; TOC; and initial and annual O&M costs for the SNG plant without sequestration using Montana Rosebud PRB coal (Case 5) are shown in Exhibit 5-18 and Exhibit 5-19, respectively. The same data for the SNG plant with sequestration using Montana Rosebud PRB coal (Case 6) are shown in Exhibit 5-20 and Exhibit 5-21.

The estimated TOC of the SNG plant without carbon sequestration is 3.35 billion dollars and with carbon sequestration is 3.44 billion dollars. Project and process contingencies represent 12.0 and 3.7 percent for both cases. The FYCOP is $19.15/MMBtu for the non-sequestration case and $21.01/MMBtu for the sequestration case as shown in Exhibit ES-6.

# Exhibit 5-18 Case 5 Total Plant Cost Details

**Client:** USDOE/NETL — **Report Date:** 2010-Feb-18
**Project:** Cost & Performance Baseline for Fossil Energy Plants Vol.2

## TOTAL PLANT COST SUMMARY

**Case:** Case 5 - Siemens Quench SNG Production w/o CO2
**Plant Size:** 54.43 MW,net — **Estimate Type:** Conceptual — **Cost Base (Jun)** 2007 ($x1000)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 1 | COAL & SORBENT HANDLING | | | | | | | | | | |
| 1.1 | Coal Receive & Unload | $6,758 | $0 | $3,303 | $0 | $0 | $10,061 | $901 | $0 | $2,192 | $13,155 |
| 1.2 | Coal Stackout & Reclaim | $8,733 | $0 | $2,117 | $0 | $0 | $10,851 | $951 | $0 | $2,360 | $14,162 |
| 1.3 | Coal Conveyors & Yd Crush | $8,120 | $0 | $2,095 | $0 | $0 | $10,215 | $897 | $0 | $2,222 | $13,334 |
| 1.4 | Other Coal Handling | $2,124 | $0 | $485 | $0 | $0 | $2,609 | $228 | $0 | $567 | $3,405 |
| 1.5 | Sorbent Receive & Unload | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.6 | Sorbent Stackout & Reclaim | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.7 | Sorbent Conveyors | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.8 | Other Sorbent Handling | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.9 | Coal & Sorbent Hnd.Foundations | $0 | $4,782 | $11,959 | $0 | $0 | $16,741 | $1,605 | $0 | $3,669 | $22,015 |
| | **SUBTOTAL 1.** | **$25,736** | **$4,782** | **$19,958** | **$0** | **$0** | **$50,477** | **$4,582** | **$0** | **$11,012** | **$66,071** |
| 2 | COAL & SORBENT PREP & FEED | | | | | | | | | | |
| 2.1 | Coal Crushing & Drying | $110,813 | $0 | $16,147 | $0 | $0 | $126,960 | $11,530 | $0 | $27,698 | $166,188 |
| 2.2 | Prepared Coal Storage & Feed | $5,249 | $1,256 | $823 | $0 | $0 | $7,328 | $627 | $0 | $1,591 | $9,545 |
| 2.3 | Dry Coal Injection System | $172,735 | $2,005 | $16,042 | $0 | $0 | $190,781 | $16,432 | $0 | $41,443 | $248,656 |
| 2.4 | Misc.Coal Prep & Feed | $2,886 | $2,100 | $6,297 | $0 | $0 | $11,284 | $1,037 | $0 | $2,464 | $14,785 |
| 2.5 | Sorbent Prep Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.6 | Sorbent Storage & Feed | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.7 | Sorbent Injection System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.8 | Booster Air Supply System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.9 | Coal & Sorbent Feed Foundation | $0 | $11,219 | $9,211 | $0 | $0 | $20,429 | $1,892 | $0 | $4,464 | $26,786 |
| | **SUBTOTAL 2.** | **$291,682** | **$16,580** | **$48,520** | **$0** | **$0** | **$356,782** | **$31,518** | **$0** | **$77,660** | **$465,959** |
| 3 | FEEDWATER & MISC. BOP SYSTEMS | | | | | | | | | | |
| 3.1 | Feedwater System | $3,669 | $1,705 | $2,068 | $0 | $0 | $7,441 | $673 | $0 | $1,623 | $9,737 |
| 3.2 | Water Makeup & Pretreating | $631 | $66 | $353 | $0 | $0 | $1,050 | $100 | $0 | $345 | $1,494 |
| 3.3 | Other Feedwater Subsystems | $3,983 | $1,204 | $1,970 | $0 | $0 | $7,157 | $650 | $0 | $1,561 | $9,368 |
| 3.4 | Service Water Systems | $361 | $744 | $2,581 | $0 | $0 | $3,685 | $360 | $0 | $1,213 | $5,258 |
| 3.5 | Other Boiler Plant Systems | $1,938 | $751 | $1,861 | $0 | $0 | $4,549 | $432 | $0 | $996 | $5,977 |
| 3.6 | FO Supply Sys & Nat Gas | $102 | $210 | $188 | $0 | $0 | $500 | $48 | $0 | $110 | $658 |
| 3.7 | Waste Treatment Equipment | $882 | $0 | $538 | $0 | $0 | $1,420 | $138 | $0 | $468 | $2,026 |
| 3.8 | Misc. Power Plant Equipment | $1,679 | $225 | $862 | $0 | $0 | $2,765 | $267 | $0 | $910 | $3,942 |
| | **SUBTOTAL 3.** | **$13,244** | **$4,904** | **$10,420** | **$0** | **$0** | **$28,568** | **$2,667** | **$0** | **$7,225** | **$38,460** |
| 4 | GASIFIER & ACCESSORIES | | | | | | | | | | |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries | $324,959 | $0 | $150,456 | $0 | $0 | $475,415 | $42,226 | $71,312 | $88,343 | $677,295 |
| 4.2 | Syngas Cooling | w/4.1 | $0 | w/ 4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.3 | ASU/Oxidant Compression | $241,072 | $0 | w/equip. | $0 | $0 | $241,072 | $23,367 | $0 | $26,444 | $290,883 |
| 4.4 | LT Heat Recovery & FG Saturation | $50,885 | $0 | $18,854 | $0 | $0 | $69,739 | $6,680 | $0 | $15,284 | $91,704 |
| 4.5 | Misc. Gasification Equipment | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.6 | Flare Stack System | $0 | $2,286 | $930 | $0 | $0 | $3,216 | $308 | $0 | $705 | $4,229 |
| 4.7 | CO2 Solid Feed System Compressors | $11,668 | $2,800 | $4,200 | $0 | $0 | $18,668 | $1,867 | $0 | $4,107 | $24,642 |
| 4.8 | Major Component Rigging | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.9 | Gasification Foundations | $0 | $24,156 | $13,784 | $0 | $0 | $37,940 | $3,473 | $0 | $10,353 | $51,766 |
| | **SUBTOTAL 4.** | **$628,583** | **$29,242** | **$188,224** | **$0** | **$0** | **$846,050** | **$77,921** | **$71,312** | **$145,236** | **$1,140,519** |

# Exhibit 5-18 Case 5 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 5A | GAS CLEANUP & PIPING | | | | | | | | | | |
| 5A.1 | Double Stage Selexol | $129,739 | $0 | $110,087 | $0 | $0 | $239,827 | $23,194 | $47,965 | $62,197 | $373,183 |
| 5A.2 | Elemental Sulfur Plant | $8,452 | $1,685 | $10,905 | $0 | $0 | $21,041 | $2,044 | $0 | $4,617 | $27,702 |
| 5A.3 | Mercury Removal | $2,420 | $0 | $1,841 | $0 | $0 | $4,261 | $412 | $213 | $977 | $5,863 |
| 5A.4 | Shift Reactors | $10,589 | $0 | $4,262 | $0 | $0 | $14,851 | $1,424 | $0 | $3,255 | $19,529 |
| 5A.5 | Methanation | $27,900 | $11,160 | $16,740 | $0 | $0 | $55,800 | $5,580 | $5,580 | $13,392 | $80,352 |
| 5A.6 | SNG Purification & Compression | $19,150 | $7,660 | $11,490 | $0 | $0 | $38,300 | $3,830 | $0 | $8,426 | $50,556 |
| 5A.7 | Fuel Gas Piping | $0 | $1,711 | $1,198 | $0 | $0 | $2,910 | $270 | $0 | $636 | $3,815 |
| 5A.9 | Process Interconnects | $0 | $12,000 | $18,000 | $0 | $0 | $30,000 | $3,000 | $0 | $6,600 | $39,600 |
| 5A.10 | HGCU Foundations | $0 | $1,733 | $1,117 | $0 | $0 | $2,850 | $262 | $0 | $934 | $4,046 |
| 5A.11 | Zinc Oxide Guard Bed | $787 | $0 | $145 | $0 | $0 | $932 | $93 | $0 | $205 | $1,230 |
| | **SUBTOTAL 5A.** | **$199,036** | **$35,949** | **$175,786** | **$0** | **$0** | **$410,771** | **$40,108** | **$53,758** | **$101,239** | **$605,876** |
| 5B | CO2 COMPRESSION | | | | | | | | | | |
| 5B.1 | CO2 Removal System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 5B.2 | CO2 Compression & Drying | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 5B.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 6 | COMBUSTION TURBINE/ACCESSORIES | | | | | | | | | | |
| 6.1 | Combustion Turbine Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.2 | Syngas Expander | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.3 | Compressed Air Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.9 | Combustion Turbine Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 6.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 7 | HRSG, DUCTING & STACK | | | | | | | | | | |
| 7.1 | Heat Recovery Steam Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.2 | HRSG Accessories | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $334 | $238 | $0 | $0 | $572 | $50 | $0 | $124 | $746 |
| 7.4 | Stack | $1,465 | $0 | $550 | $0 | $0 | $2,015 | $193 | $0 | $221 | $2,429 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $293 | $282 | $0 | $0 | $575 | $54 | $0 | $189 | $818 |
| | **SUBTOTAL 7.** | **$1,465** | **$627** | **$1,070** | **$0** | **$0** | **$3,162** | **$297** | **$0** | **$534** | **$3,993** |
| 8 | STEAM TURBINE GENERATOR | | | | | | | | | | |
| 8.1 | Steam TG & Accessories | $30,976 | $0 | $5,362 | $0 | $0 | $36,338 | $3,487 | $0 | $3,982 | $43,807 |
| 8.2 | Turbine Plant Auxiliaries | $170 | $0 | $494 | $0 | $0 | $664 | $65 | $0 | $73 | $802 |
| 8.3a | Condenser & Auxiliaries | $3,219 | $0 | $1,583 | $0 | $0 | $4,802 | $461 | $0 | $526 | $5,790 |
| 8.3b | Air Cooled Condenser | $29,498 | $0 | $5,914 | $0 | $0 | $35,412 | $3,541 | $0 | $7,791 | $46,744 |
| 8.4 | Steam Piping | $5,733 | $0 | $4,033 | $0 | $0 | $9,765 | $839 | $0 | $2,651 | $13,255 |
| 8.9 | TG Foundations | $0 | $1,069 | $1,806 | $0 | $0 | $2,875 | $273 | $0 | $944 | $4,092 |
| | **SUBTOTAL 8.** | **$69,595** | **$1,069** | **$19,192** | **$0** | **$0** | **$89,856** | **$8,666** | **$0** | **$15,968** | **$114,489** |

## Exhibit 5-18 Case 5 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 9 | COOLING WATER SYSTEM | | | | | | | | | | |
| 9.1 | Cooling Towers | $7,073 | $0 | $1,287 | $0 | $0 | $8,360 | $796 | $0 | $1,373 | $10,530 |
| 9.2 | Circulating Water Pumps | $1,875 | $0 | $87 | $0 | $0 | $1,961 | $165 | $0 | $319 | $2,445 |
| 9.3 | Circ.Water System Auxiliaries | $154 | $0 | $22 | $0 | $0 | $176 | $17 | $0 | $29 | $221 |
| 9.4 | Circ.Water Piping | $0 | $6,417 | $1,664 | $0 | $0 | $8,081 | $730 | $0 | $1,762 | $10,573 |
| 9.5 | Make-up Water System | $348 | $0 | $498 | $0 | $0 | $846 | $81 | $0 | $185 | $1,113 |
| 9.6 | Component Cooling Water Sys | $757 | $906 | $645 | $0 | $0 | $2,308 | $216 | $0 | $505 | $3,029 |
| 9.9 | Circ.Water System Foundations | $0 | $2,356 | $4,005 | $0 | $0 | $6,360 | $603 | $0 | $2,089 | $9,052 |
| | **SUBTOTAL 9.** | **$10,207** | **$9,679** | **$8,206** | **$0** | **$0** | **$28,092** | **$2,608** | **$0** | **$6,263** | **$36,963** |
| 10 | ASH/SPENT SORBENT HANDLING SYS | | | | | | | | | | |
| 10.1 | Slag Dewatering & Cooling | $38,544 | $0 | $38,016 | $0 | $0 | $76,559 | $7,418 | $0 | $8,398 | $92,375 |
| 10.2 | Gasifier Ash Depressurization | w/10.1 | $0 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.3 | Cleanup Ash Depressurization | w/10.1 | $0 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.4 | High Temperature Ash Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.5 | Other Ash Rrecovery Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.6 | Ash Storage Silos | $862 | $0 | $938 | $0 | $0 | $1,800 | $175 | $0 | $296 | $2,271 |
| 10.7 | Ash Transport & Feed Equipment | $1,156 | $0 | $279 | $0 | $0 | $1,435 | $134 | $0 | $235 | $1,805 |
| 10.8 | Misc. Ash Handling Equipment | $1,786 | $2,188 | $654 | $0 | $0 | $4,628 | $440 | $0 | $760 | $5,828 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $76 | $96 | $0 | $0 | $172 | $16 | $0 | $56 | $245 |
| | **SUBTOTAL 10.** | **$42,348** | **$2,265** | **$39,982** | **$0** | **$0** | **$84,594** | **$8,183** | **$0** | **$9,746** | **$102,524** |
| 11 | ACCESSORY ELECTRIC PLANT | | | | | | | | | | |
| 11.1 | Generator Equipment | $428 | $0 | $424 | $0 | $0 | $852 | $81 | $0 | $93 | $1,026 |
| 11.2 | Station Service Equipment | $5,579 | $0 | $503 | $0 | $0 | $6,082 | $561 | $0 | $664 | $7,307 |
| 11.3 | Switchgear & Motor Control | $10,315 | $0 | $1,876 | $0 | $0 | $12,191 | $1,131 | $0 | $1,998 | $15,319 |
| 11.4 | Conduit & Cable Tray | $0 | $4,792 | $15,807 | $0 | $0 | $20,598 | $1,992 | $0 | $5,648 | $28,238 |
| 11.5 | Wire & Cable | $0 | $9,155 | $6,015 | $0 | $0 | $15,170 | $1,102 | $0 | $4,068 | $20,340 |
| 11.6 | Protective Equipment | $0 | $830 | $3,021 | $0 | $0 | $3,851 | $376 | $0 | $634 | $4,861 |
| 11.7 | Standby Equipment | $323 | $0 | $315 | $0 | $0 | $639 | $61 | $0 | $105 | $804 |
| 11.8 | Main Power Transformers | $9,154 | $0 | $177 | $0 | $0 | $9,331 | $708 | $0 | $1,506 | $11,545 |
| 11.9 | Electrical Foundations | $0 | $192 | $503 | $0 | $0 | $695 | $67 | $0 | $228 | $990 |
| | **SUBTOTAL 11.** | **$25,800** | **$14,968** | **$28,640** | **$0** | **$0** | **$69,408** | **$6,079** | **$0** | **$14,945** | **$90,432** |
| 12 | INSTRUMENTATION & CONTROL | | | | | | | | | | |
| 12.1 | IGCC Control Equipment | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.2 | Combustion Turbine Control | w/6.1 | $0 | w/6.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.3 | Steam Turbine Control | w/8.1 | $0 | w/8.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.4 | Other Major Component Control | $528 | $0 | $353 | $0 | $0 | $881 | $83 | $44 | $151 | $1,160 |
| 12.5 | Signal Processing Equipment | w/12.7 | $0 | w/12.7 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.6 | Control Boards,Panels & Racks | $351 | $0 | $225 | $0 | $0 | $576 | $54 | $29 | $132 | $791 |
| 12.7 | Computer & Accessories | $5,617 | $0 | $180 | $0 | $0 | $5,797 | $532 | $290 | $662 | $7,280 |
| 12.8 | Instrument Wiring & Tubing | $0 | $2,845 | $5,816 | $0 | $0 | $8,661 | $735 | $433 | $2,457 | $12,286 |
| 12.9 | Other I & C Equipment | $4,188 | $0 | $2,034 | $0 | $0 | $6,221 | $585 | $311 | $1,068 | $8,185 |
| | **SUBTOTAL 12.** | **$10,684** | **$2,845** | **$8,608** | **$0** | **$0** | **$22,137** | **$1,990** | **$1,107** | **$4,470** | **$29,703** |

# Exhibit 5-18 Case 5 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 13 | IMPROVEMENTS TO SITE | | | | | | | | | | |
| 13.1 | Site Preparation | $0 | $144 | $3,080 | $0 | $0 | $3,224 | $320 | $0 | $1,063 | $4,607 |
| 13.2 | Site Improvements | $0 | $2,563 | $3,406 | $0 | $0 | $5,969 | $589 | $0 | $1,967 | $8,525 |
| 13.3 | Site Facilities | $4,593 | $0 | $4,846 | $0 | $0 | $9,439 | $931 | $0 | $3,111 | $13,481 |
| | **SUBTOTAL 13.** | **$4,593** | **$2,707** | **$11,332** | **$0** | **$0** | **$18,632** | **$1,840** | **$0** | **$6,141** | **$26,613** |
| 14 | BUILDINGS & STRUCTURES | | | | | | | | | | |
| 14.1 | Combustion Turbine Area | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 14.2 | Steam Turbine Building | $0 | $2,580 | $3,675 | $0 | $0 | $6,255 | $576 | $0 | $1,025 | $7,855 |
| 14.3 | Administration Building | $0 | $870 | $630 | $0 | $0 | $1,500 | $134 | $0 | $245 | $1,879 |
| 14.4 | Circulation Water Pumphouse | $0 | $160 | $90 | $0 | $0 | $250 | $22 | $0 | $41 | $312 |
| 14.5 | Water Treatment Buildings | $0 | $527 | $515 | $0 | $0 | $1,042 | $94 | $0 | $170 | $1,307 |
| 14.6 | Machine Shop | $0 | $445 | $305 | $0 | $0 | $750 | $67 | $0 | $123 | $939 |
| 14.7 | Warehouse | $0 | $726 | $464 | $0 | $0 | $1,190 | $105 | $0 | $194 | $1,489 |
| 14.8 | Other Buildings & Structures | $0 | $521 | $406 | $0 | $0 | $927 | $83 | $0 | $202 | $1,212 |
| 14.9 | Waste Treating Building & Str. | $0 | $1,165 | $2,226 | $0 | $0 | $3,391 | $316 | $0 | $741 | $4,448 |
| | **SUBTOTAL 14.** | **$0** | **$6,994** | **$8,311** | **$0** | **$0** | **$15,304** | **$1,396** | **$0** | **$2,741** | **$19,441** |
| | | | | | | | | | | | |
| | **TOTAL COST** | **$1,322,974** | **$132,611** | **$568,248** | **$0** | **$0** | **$2,023,833** | **$187,854** | **$126,177** | **$403,179** | **$2,741,044** |
| | | | | | | | | | | | |
| | **Owner's Costs** | | | | | | | | | | |
| | **Preproduction Costs** | | | | | | | | | | |
| | 6 Months All Labor | | | | | | | | | | $21,718 |
| | 1 Month Maintenance Materials | | | | | | | | | | $3,840 |
| | 1 Month Non-fuel Consumables | | | | | | | | | | $575 |
| | 1 Month Waste Disposal | | | | | | | | | | $634 |
| | 25% of 1 Months Fuel Cost at 100% CF | | | | | | | | | | $1,749 |
| | 2% of TPC | | | | | | | | | | $54,821 |
| | **Total** | | | | | | | | | | **$83,337** |
| | **Inventory Capital** | | | | | | | | | | |
| | 60 day supply of fuel and consumables at 100% CF | | | | | | | | | | $14,910 |
| | 0.5% of TPC (spare parts) | | | | | | | | | | $13,705 |
| | **Total** | | | | | | | | | | **$28,615** |
| | | | | | | | | | | | |
| | **Initial Cost for Catalyst and Chemicals** | | | | | | | | | | $15,381 |
| | **Land** | | | | | | | | | | $900 |
| | **Other Owner's Costs** | | | | | | | | | | $411,157 |
| | **Financing Costs** | | | | | | | | | | $74,008 |
| | **Total Overnight Costs (TOC)** | | | | | | | | | | **$3,354,442** |
| | TASC Multiplier | | | | | | | | | | 1.201 |
| | **Total As-Spent Cost (TASC)** | | | | | | | | | | **$4,028,684** |

# Exhibit 5-19 Case 5 Initial and Annual O&M Costs

| | | | | | | |
|---|---|---|---|---|---|---|
| INITIAL & ANNUAL O&M EXPENSES | | | | | Cost Base (Jun): | 2007 |
| Case 5 - Siemens Quench SNG Production w/o CO2 | | | | | MWe-net: | 54 |
| SNG (MMbtu/hr): | 6789 | | | | Capacity Factor (%): | 90 |
| OPERATING & MAINTENANCE LABOR | | | | | | |
| Operating Labor | | | | | | |
| Operating Labor Rate(base): | 34.65 | $/hour | | | | |
| Operating Labor Burden: | 30.00 | % of base | | | | |
| Labor O-H Charge Rate: | 25.00 | % of labor | | | | |
| | | | Total | | | |
| Operating Labor Requirements(O.J.)per Shift: | 1 unit/mod. | | Plant | | | |
| Skilled Operator | 2.0 | | 2.0 | | | |
| Operator | 12.0 | | 12.0 | | | |
| Foreman | 1.0 | | 1.0 | | | |
| Lab Tech's, etc. | 3.0 | | 3.0 | | | |
| TOTAL-O.J.'s | 18.0 | | 18.0 | | | |
| | | | | | Annual Cost | Annual Unit Cost |
| | | | | | $ | $/MMBtu |
| Annual Operating Labor Cost | | | | | $7,102,696 | $0.133 |
| Maintenance Labor Cost | | | | | $27,646,499 | $0.516 |
| Administrative & Support Labor | | | | | $8,687,299 | $0.162 |
| Property Taxes and Insurance | | | | | $54,820,873 | $1.024 |
| **TOTAL FIXED OPERATING COSTS** | | | | | **$98,257,367** | **$1.836** |
| VARIABLE OPERATING COSTS | | | | | | |
| | | | | | | $/MMBtu |
| **Maintenance Material Cost** | | | | | **$41,469,749** | **$0.77474** |
| Consumables | Consumption | | Unit | Initial Fill | | |
| | Initial Fill | /Day | Cost | Cost | | |
| **Water(/1000 gallons)** | 0 | 3,494 | 1.08 | **$0** | **$1,241,581** | **$0.02320** |
| **Chemicals** | | | | | | |
| MU & WT Chem. (lb) | 0 | 20,817 | 0.17 | $0 | $1,183,517 | $0.02211 |
| Carbon (Mercury Removal) (lb) | 223,863 | 383 | 1.05 | $235,094 | $132,241 | $0.00247 |
| COS Catalyst (m3) | 0 | 0 | 2,397.36 | $0 | $0 | $0.00000 |
| Water Gas Shift Catalyst (ft3) | 5,564 | 4.76 | 498.83 | $2,775,453 | $780,596 | $0.01458 |
| ZnO Sorbent (ton) | 58 | 0.20 | 12,574.00 | $723,651 | $814,107 | $0.01521 |
| Methanation Catalyst (ft3) | 12,183 | 9.27 | 440.00 | $5,360,314 | $1,340,078 | $0.02504 |
| Selexol Solution (gal) | 469,206 | 149 | 13.40 | $6,286,534 | $656,108 | $0.01226 |
| SCR Catalyst (m3) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Aqueous Ammonia (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Claus Catalyst (ft3) | w/equip | 1.53 | 131.27 | $0 | $66,023 | $0.00123 |
| **Subtotal Chemicals** | | | | **$15,381,047** | **$4,972,671** | **$0.09290** |
| **Other** | | | | | | |
| Supplemental Fuel (MBtu) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for consumption) (MW | 0 | 0 | 61.60 | $0 | $0 | $0.00000 |
| Gases,N2 etc. (/100scf) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| L.P. Steam (/1000 pounds) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| **Subtotal Other** | | | | **$0** | **$0** | **$0.00000** |
| **Waste Disposal** | | | | | | |
| Spent Mercury Catalyst (lb.) | 0 | 383 | 0.42 | $0 | $52,520 | $0.00098 |
| Spent ZnO Sorbent (ton) | 0 | 0.20 | 16.23 | $0 | $1,051 | $0.00002 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Slag (ton) | 0 | 1,275 | 16.23 | $0 | $6,796,412 | $0.12697 |
| **Subtotal-Waste Disposal** | | | | **$0** | **$6,849,982** | **$0.12797** |
| **By-products & Emissions** | | | | | | |
| Sulfur (tons) | 0 | 108 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for sale) (MWh) | 0 | 1,306 | 58.00 | $0 | -$24,889,315 | -$0.46498 |
| **Subtotal By-Products** | | | | **$0** | **-$24,889,315** | **-$0.46498** |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$15,381,047** | **$29,644,668** | **$0.55382** |
| **Fuel (ton)** | 0 | 15,112 | 15.22 | **$0** | **$75,540,499** | **$1.41125** |

# Exhibit 5-20 Case 6 Total Plant Cost Details

| | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | Client: | USDOE/NETL | | | | | | | | Report Date: | 2010-Feb-18 |
| | Project: | Cost & Performance Baseline for Fossil Energy Plants Vol.2 | | | | | | | | | |
| | | TOTAL PLANT COST SUMMARY | | | | | | | | | |
| | Case: | Case 6 - Siemens Quench SNG Production w/ CO2 | | | | | | | | | |
| | Plant Size: | 1.81 | MW,net | Estimate Type: | | Conceptual | | | Cost Base (Jun) | 2007 | ($x1000) |
| Acct | | Equipment | Material | Labor | | Sales | Bare Erected | Eng'g CM | Contingencies | | Total Plant Cost |
| No. | Item/Description | Cost | Cost | Direct | Indirect | Tax | Cost $ | H.O.& Fee | Process | Project | $ |
| 1 | COAL & SORBENT HANDLING | | | | | | | | | | |
| 1.1 | Coal Receive & Unload | $6,758 | $0 | $3,303 | $0 | $0 | $10,061 | $901 | $0 | $2,192 | $13,155 |
| 1.2 | Coal Stackout & Reclaim | $8,733 | $0 | $2,117 | $0 | $0 | $10,851 | $951 | $0 | $2,360 | $14,162 |
| 1.3 | Coal Conveyors & Yd Crush | $8,120 | $0 | $2,095 | $0 | $0 | $10,215 | $897 | $0 | $2,222 | $13,334 |
| 1.4 | Other Coal Handling | $2,124 | $0 | $485 | $0 | $0 | $2,609 | $228 | $0 | $567 | $3,405 |
| 1.5 | Sorbent Receive & Unload | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.6 | Sorbent Stackout & Reclaim | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.7 | Sorbent Conveyors | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.8 | Other Sorbent Handling | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.9 | Coal & Sorbent Hnd.Foundations | $0 | $4,782 | $11,959 | $0 | $0 | $16,741 | $1,605 | $0 | $3,669 | $22,015 |
| | **SUBTOTAL 1.** | **$25,736** | **$4,782** | **$19,958** | **$0** | **$0** | **$50,477** | **$4,582** | **$0** | **$11,012** | **$66,071** |
| 2 | COAL & SORBENT PREP & FEED | | | | | | | | | | |
| 2.1 | Coal Crushing & Drying | $110,813 | $0 | $16,147 | $0 | $0 | $126,960 | $11,530 | $0 | $27,698 | $166,188 |
| 2.2 | Prepared Coal Storage & Feed | $5,249 | $1,256 | $823 | $0 | $0 | $7,328 | $627 | $0 | $1,591 | $9,545 |
| 2.3 | Dry Coal Injection System | $172,735 | $2,005 | $16,042 | $0 | $0 | $190,781 | $16,432 | $0 | $41,443 | $248,656 |
| 2.4 | Misc.Coal Prep & Feed | $2,886 | $2,100 | $6,297 | $0 | $0 | $11,284 | $1,037 | $0 | $2,464 | $14,785 |
| 2.5 | Sorbent Prep Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.6 | Sorbent Storage & Feed | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.7 | Sorbent Injection System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.8 | Booster Air Supply System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.9 | Coal & Sorbent Feed Foundation | $0 | $11,219 | $9,211 | $0 | $0 | $20,429 | $1,892 | $0 | $4,464 | $26,786 |
| | **SUBTOTAL 2.** | **$291,682** | **$16,580** | **$48,520** | **$0** | **$0** | **$356,782** | **$31,518** | **$0** | **$77,660** | **$465,959** |
| 3 | FEEDWATER & MISC. BOP SYSTEMS | | | | | | | | | | |
| 3.1 | Feedwater System | $3,662 | $1,702 | $2,064 | $0 | $0 | $7,428 | $671 | $0 | $1,620 | $9,719 |
| 3.2 | Water Makeup & Pretreating | $690 | $72 | $386 | $0 | $0 | $1,148 | $109 | $0 | $377 | $1,635 |
| 3.3 | Other Feedwater Subsystems | $3,976 | $1,202 | $1,966 | $0 | $0 | $7,144 | $649 | $0 | $1,559 | $9,351 |
| 3.4 | Service Water Systems | $395 | $814 | $2,824 | $0 | $0 | $4,032 | $393 | $0 | $1,328 | $5,753 |
| 3.5 | Other Boiler Plant Systems | $2,120 | $822 | $2,036 | $0 | $0 | $4,978 | $472 | $0 | $1,090 | $6,540 |
| 3.6 | FO Supply Sys & Nat Gas | $102 | $210 | $188 | $0 | $0 | $500 | $48 | $0 | $110 | $658 |
| 3.7 | Waste Treatment Equipment | $965 | $0 | $589 | $0 | $0 | $1,554 | $151 | $0 | $512 | $2,217 |
| 3.8 | Misc. Power Plant Equipment | $1,679 | $225 | $862 | $0 | $0 | $2,765 | $267 | $0 | $910 | $3,942 |
| | **SUBTOTAL 3.** | **$13,589** | **$5,046** | **$10,915** | **$0** | **$0** | **$29,550** | **$2,761** | **$0** | **$7,504** | **$39,816** |
| 4 | GASIFIER & ACCESSORIES | | | | | | | | | | |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries | $324,934 | $0 | $150,445 | $0 | $0 | $475,379 | $42,223 | $71,307 | $88,336 | $677,245 |
| 4.2 | Syngas Cooling | w/4.1 | $0 | w/ 4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.3 | ASU/Oxidant Compression | $241,072 | $0 | w/equip. | $0 | $0 | $241,072 | $23,367 | $0 | $26,444 | $290,883 |
| 4.4 | LT Heat Recovery & FG Saturation | $50,884 | $0 | $18,854 | $0 | $0 | $69,738 | $6,680 | $0 | $15,284 | $91,702 |
| 4.5 | Misc. Gasification Equipment | w/4.1 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.6 | Other Gasification Equipment | $0 | $2,285 | $930 | $0 | $0 | $3,215 | $308 | $0 | $705 | $4,228 |
| 4.7 | CO2 Solid Feed System Compressors | $7,427 | $1,782 | $2,674 | $0 | $0 | $11,883 | $1,188 | $0 | $2,614 | $15,685 |
| 4.8 | Major Component Rigging | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.9 | Gasification Foundations | $0 | $24,156 | $13,784 | $0 | $0 | $37,940 | $3,473 | $0 | $10,353 | $51,766 |
| | **SUBTOTAL 4.** | **$624,317** | **$28,224** | **$186,685** | **$0** | **$0** | **$839,226** | **$77,240** | **$71,307** | **$143,736** | **$1,131,508** |

# Exhibit 5-20 Case 6 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 5A | GAS CLEANUP & PIPING | | | | | | | | | | |
| 5A.1 | Double Stage Selexol | $129,739 | $0 | $110,087 | $0 | $0 | $239,826 | $23,194 | $47,965 | $62,197 | $373,182 |
| 5A.2 | Elemental Sulfur Plant | $8,455 | $1,685 | $10,909 | $0 | $0 | $21,049 | $2,045 | $0 | $4,619 | $27,713 |
| 5A.3 | Mercury Removal | $2,401 | $0 | $1,827 | $0 | $0 | $4,229 | $408 | $211 | $970 | $5,818 |
| 5A.4 | Shift Reactors | $10,434 | $0 | $4,200 | $0 | $0 | $14,633 | $1,403 | $0 | $3,207 | $19,243 |
| 5A.5 | Methanation | $27,400 | $10,960 | $16,440 | $0 | $0 | $54,800 | $5,480 | $5,480 | $13,152 | $78,912 |
| 5A.6 | SNG Purification & Compression | $19,150 | $7,660 | $11,490 | $0 | $0 | $38,300 | $3,830 | $0 | $8,426 | $50,556 |
| 5A.7 | Fuel Gas Piping | $0 | $1,710 | $1,198 | $0 | $0 | $2,908 | $270 | $0 | $635 | $3,813 |
| 5A.9 | Process Interconnects | $0 | $12,000 | $18,000 | $0 | $0 | $30,000 | $3,000 | $0 | $6,600 | $39,600 |
| 5A.10 | HGCU Foundations | $0 | $1,732 | $1,117 | $0 | $0 | $2,848 | $262 | $0 | $933 | $4,043 |
| 5A.11 | Zinc Oxide Guard Bed | $786 | $0 | $145 | $0 | $0 | $931 | $93 | $0 | $205 | $1,229 |
| | **SUBTOTAL 5A.** | **$198,366** | **$35,747** | **$175,412** | **$0** | **$0** | **$409,525** | **$39,984** | **$53,657** | **$100,944** | **$604,110** |
| 5B | CO2 COMPRESSION | | | | | | | | | | |
| 5B.1 | CO2 Removal System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 5B.2 | CO2 Compression & Drying | $36,117 | $0 | $21,269 | $0 | $0 | $57,386 | $5,525 | $0 | $12,582 | $75,493 |
| | **SUBTOTAL 5B.** | **$36,117** | **$0** | **$21,269** | **$0** | **$0** | **$57,386** | **$5,525** | **$0** | **$12,582** | **$75,493** |
| 6 | COMBUSTION TURBINE/ACCESSORIES | | | | | | | | | | |
| 6.1 | Combustion Turbine Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.2 | Syngas Expander | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.3 | Compressed Air Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.9 | Combustion Turbine Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 6.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 7 | HRSG, DUCTING & STACK | | | | | | | | | | |
| 7.1 | Heat Recovery Steam Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.2 | HRSG Accessories | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.4 | Stack | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.9 | HRSG, Duct & Stack Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 7.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 8 | STEAM TURBINE GENERATOR | | | | | | | | | | |
| 8.1 | Steam TG & Accessories | $30,976 | $0 | $5,362 | $0 | $0 | $36,338 | $3,487 | $0 | $3,982 | $43,807 |
| 8.2 | Turbine Plant Auxiliaries | $169 | $0 | $494 | $0 | $0 | $663 | $65 | $0 | $73 | $801 |
| 8.3a | Condenser & Auxiliaries | $3,219 | $0 | $1,583 | $0 | $0 | $4,802 | $461 | $0 | $526 | $5,790 |
| 8.3b | Air Cooled Condenser | $29,498 | $0 | $5,914 | $0 | $0 | $35,412 | $3,541 | $0 | $7,791 | $46,744 |
| 8.4 | Steam Piping | $5,722 | $0 | $4,025 | $0 | $0 | $9,747 | $837 | $0 | $2,646 | $13,230 |
| 8.9 | TG Foundations | $0 | $1,069 | $1,806 | $0 | $0 | $2,875 | $273 | $0 | $944 | $4,092 |
| | **SUBTOTAL 8.** | **$69,584** | **$1,069** | **$19,184** | **$0** | **$0** | **$89,837** | **$8,664** | **$0** | **$15,963** | **$114,464** |

# Exhibit 5-20 Case 6 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 9 | COOLING WATER SYSTEM | | | | | | | | | | |
| 9.1 | Cooling Towers | $7,758 | $0 | $1,411 | $0 | $0 | $9,169 | $873 | $0 | $1,506 | $11,549 |
| 9.2 | Circulating Water Pumps | $2,057 | $0 | $87 | $0 | $0 | $2,143 | $180 | $0 | $349 | $2,672 |
| 9.3 | Circ.Water System Auxiliaries | $166 | $0 | $24 | $0 | $0 | $190 | $18 | $0 | $31 | $240 |
| 9.4 | Circ.Water Piping | $0 | $6,948 | $1,801 | $0 | $0 | $8,749 | $791 | $0 | $1,908 | $11,448 |
| 9.5 | Make-up Water System | $376 | $0 | $537 | $0 | $0 | $913 | $88 | $0 | $200 | $1,200 |
| 9.6 | Component Cooling Water Sys | $820 | $981 | $698 | $0 | $0 | $2,499 | $234 | $0 | $547 | $3,280 |
| 9.9 | Circ.Water System Foundations | $0 | $2,550 | $4,335 | $0 | $0 | $6,884 | $653 | $0 | $2,261 | $9,798 |
| | **SUBTOTAL 9.** | **$11,177** | **$10,479** | **$8,893** | **$0** | **$0** | **$30,549** | **$2,836** | **$0** | **$6,802** | **$40,187** |
| 10 | ASH/SPENT SORBENT HANDLING SYS | | | | | | | | | | |
| 10.1 | Slag Dewatering & Cooling | $38,544 | $0 | $38,016 | $0 | $0 | $76,559 | $7,418 | $0 | $8,398 | $92,375 |
| 10.2 | Gasifier Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.3 | Cleanup Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.4 | High Temperature Ash Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.5 | Other Ash Recovery Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.6 | Ash Storage Silos | $862 | $0 | $938 | $0 | $0 | $1,800 | $175 | $0 | $296 | $2,271 |
| 10.7 | Ash Transport & Feed Equipment | $1,156 | $0 | $279 | $0 | $0 | $1,435 | $134 | $0 | $235 | $1,805 |
| 10.8 | Misc. Ash Handling Equipment | $1,786 | $2,188 | $654 | $0 | $0 | $4,628 | $440 | $0 | $760 | $5,828 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $76 | $96 | $0 | $0 | $172 | $16 | $0 | $56 | $245 |
| | **SUBTOTAL 10.** | **$42,348** | **$2,265** | **$39,982** | **$0** | **$0** | **$84,594** | **$8,183** | **$0** | **$9,746** | **$102,524** |
| 11 | ACCESSORY ELECTRIC PLANT | | | | | | | | | | |
| 11.1 | Generator Equipment | $428 | $0 | $424 | $0 | $0 | $852 | $81 | $0 | $93 | $1,026 |
| 11.2 | Station Service Equipment | $6,060 | $0 | $546 | $0 | $0 | $6,606 | $609 | $0 | $721 | $7,936 |
| 11.3 | Switchgear & Motor Control | $11,203 | $0 | $2,037 | $0 | $0 | $13,240 | $1,228 | $0 | $2,170 | $16,638 |
| 11.4 | Conduit & Cable Tray | $0 | $5,204 | $17,168 | $0 | $0 | $22,372 | $2,164 | $0 | $6,134 | $30,670 |
| 11.5 | Wire & Cable | $0 | $9,943 | $6,533 | $0 | $0 | $16,476 | $1,197 | $0 | $4,418 | $22,092 |
| 11.6 | Protective Equipment | $0 | $830 | $3,021 | $0 | $0 | $3,851 | $376 | $0 | $634 | $4,861 |
| 11.7 | Standby Equipment | $323 | $0 | $315 | $0 | $0 | $639 | $61 | $0 | $105 | $804 |
| 11.8 | Main Power Transformers | $9,866 | $0 | $177 | $0 | $0 | $10,043 | $762 | $0 | $1,621 | $12,425 |
| 11.9 | Electrical Foundations | $0 | $192 | $503 | $0 | $0 | $695 | $67 | $0 | $228 | $990 |
| | **SUBTOTAL 11.** | **$27,880** | **$16,169** | **$30,724** | **$0** | **$0** | **$74,773** | **$6,544** | **$0** | **$16,125** | **$97,443** |
| 12 | INSTRUMENTATION & CONTROL | | | | | | | | | | |
| 12.1 | IGCC Control Equipment | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.2 | Combustion Turbine Control | w/6.1 | $0 | w/6.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.3 | Steam Turbine Control | w/8.1 | $0 | w/8.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.4 | Other Major Component Control | $525 | $0 | $351 | $0 | $0 | $876 | $83 | $44 | $150 | $1,152 |
| 12.5 | Signal Processing Equipment | w/12.7 | $0 | w/12.7 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.6 | Control Boards,Panels & Racks | $349 | $0 | $223 | $0 | $0 | $572 | $54 | $29 | $131 | $786 |
| 12.7 | Computer & Accessories | $5,579 | $0 | $179 | $0 | $0 | $5,758 | $529 | $288 | $657 | $7,232 |
| 12.8 | Instrument Wiring & Tubing | $0 | $2,826 | $5,778 | $0 | $0 | $8,604 | $730 | $430 | $2,441 | $12,205 |
| 12.9 | Other I & C Equipment | $4,160 | $0 | $2,020 | $0 | $0 | $6,180 | $581 | $309 | $1,061 | $8,131 |
| | **SUBTOTAL 12.** | **$10,613** | **$2,826** | **$8,550** | **$0** | **$0** | **$21,989** | **$1,977** | **$1,099** | **$4,440** | **$29,505** |

## Exhibit 5-20 Case 6 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 13 | IMPROVEMENTS TO SITE | | | | | | | | | | |
| 13.1 | Site Preparation | $0 | $144 | $3,080 | $0 | $0 | $3,224 | $320 | $0 | $1,063 | $4,607 |
| 13.2 | Site Improvements | $0 | $2,563 | $3,406 | $0 | $0 | $5,969 | $589 | $0 | $1,967 | $8,525 |
| 13.3 | Site Facilities | $4,593 | $0 | $4,846 | $0 | $0 | $9,439 | $931 | $0 | $3,111 | $13,481 |
| | **SUBTOTAL 13.** | **$4,593** | **$2,707** | **$11,332** | **$0** | **$0** | **$18,632** | **$1,840** | **$0** | **$6,141** | **$26,613** |
| 14 | BUILDINGS & STRUCTURES | | | | | | | | | | |
| 14.1 | Combustion Turbine Area | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 14.2 | Steam Turbine Building | $0 | $2,580 | $3,675 | $0 | $0 | $6,255 | $576 | $0 | $1,025 | $7,855 |
| 14.3 | Administration Building | $0 | $870 | $630 | $0 | $0 | $1,500 | $134 | $0 | $245 | $1,879 |
| 14.4 | Circulation Water Pumphouse | $0 | $160 | $90 | $0 | $0 | $250 | $22 | $0 | $41 | $312 |
| 14.5 | Water Treatment Buildings | $0 | $577 | $563 | $0 | $0 | $1,140 | $103 | $0 | $186 | $1,430 |
| 14.6 | Machine Shop | $0 | $445 | $305 | $0 | $0 | $750 | $67 | $0 | $123 | $939 |
| 14.7 | Warehouse | $0 | $726 | $464 | $0 | $0 | $1,190 | $105 | $0 | $194 | $1,489 |
| 14.8 | Other Buildings & Structures | $0 | $521 | $406 | $0 | $0 | $927 | $83 | $0 | $202 | $1,212 |
| 14.9 | Waste Treating Building & Str. | $0 | $1,165 | $2,226 | $0 | $0 | $3,391 | $316 | $0 | $741 | $4,448 |
| | **SUBTOTAL 14.** | **$0** | **$7,043** | **$8,359** | **$0** | **$0** | **$15,403** | **$1,405** | **$0** | **$2,757** | **$19,564** |
| | | | | | | | | | | | |
| | **TOTAL COST** | **$1,356,002** | **$132,937** | **$589,783** | **$0** | **$0** | **$2,078,723** | **$193,059** | **$126,063** | **$415,413** | **$2,813,258** |
| | | | | | | | | | | | |
| | **Owner's Costs** | | | | | | | | | | |
| | **Preproduction Costs** | | | | | | | | | | |
| | 6 Months All Labor | | | | | | | | | | $21,702 |
| | 1 Month Maintenance Materials | | | | | | | | | | $3,836 |
| | 1 Month Non-fuel Consumables | | | | | | | | | | $630 |
| | 1 Month Waste Disposal | | | | | | | | | | $634 |
| | 25% of 1 Months Fuel Cost at 100% CF | | | | | | | | | | $1,749 |
| | 2% of TPC | | | | | | | | | | $56,265 |
| | **Total** | | | | | | | | | | **$84,817** |
| | **Inventory Capital** | | | | | | | | | | |
| | 60 day supply of fuel and consumables at 100% CF | | | | | | | | | | $14,989 |
| | 0.5% of TPC (spare parts) | | | | | | | | | | $14,066 |
| | **Total** | | | | | | | | | | **$29,055** |
| | | | | | | | | | | | |
| | **Initial Cost for Catalyst and Chemicals** | | | | | | | | | | $16,334 |
| | **Land** | | | | | | | | | | $900 |
| | **Other Owner's Costs** | | | | | | | | | | $421,989 |
| | **Financing Costs** | | | | | | | | | | $75,958 |
| | **Total Overnight Costs (TOC)** | | | | | | | | | | **$3,442,310** |
| | TASC Multiplier | | | | | | | | | | 1.201 |
| | **Total As-Spent Cost (TASC)** | | | | | | | | | | **$4,134,215** |

# Exhibit 5-21 Case 6 Initial and Annual O&M Costs

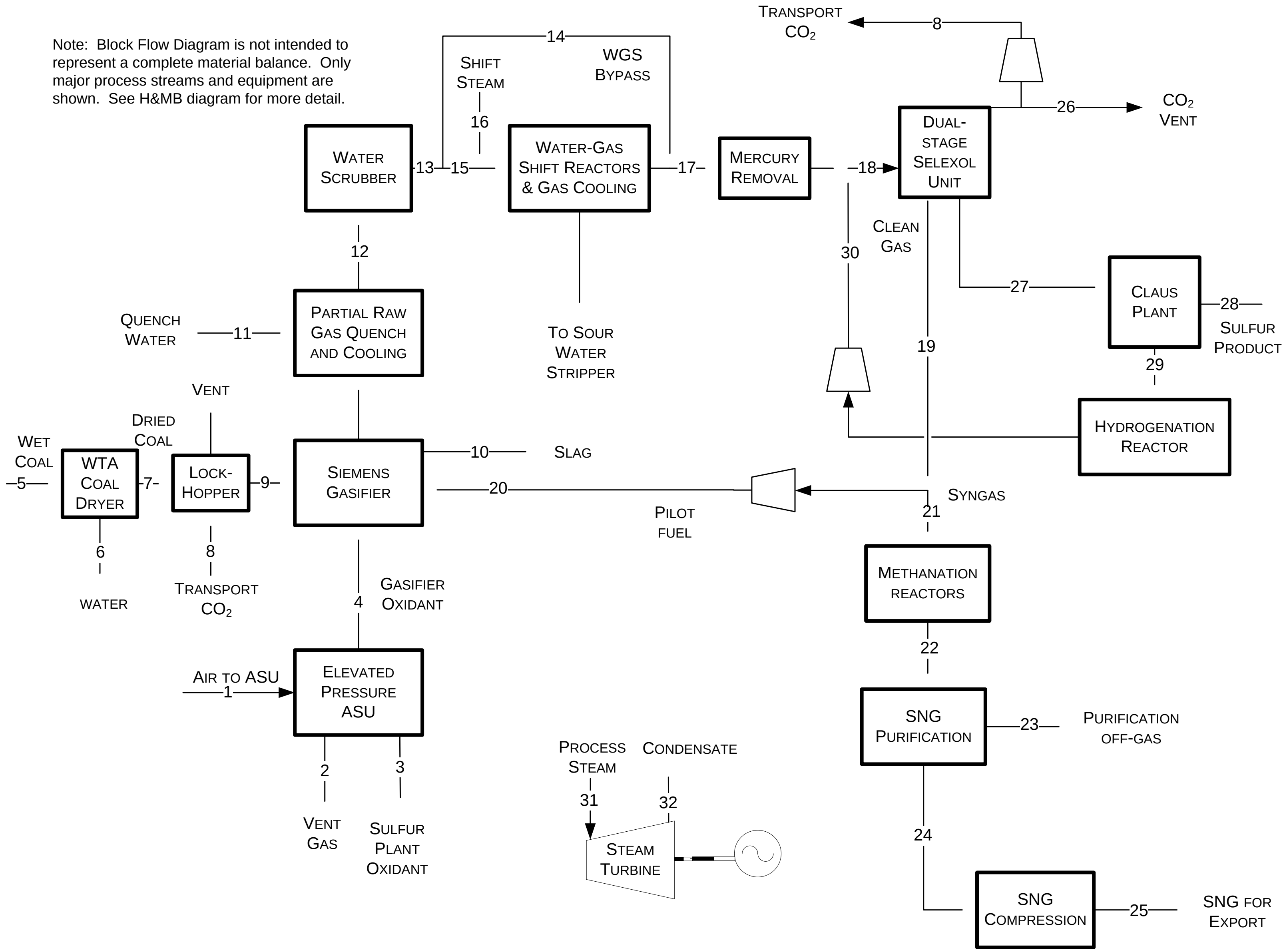
| | | | | | | |
|---|---|---|---|---|---|---|
| INITIAL & ANNUAL O&M EXPENSES | | | | | Cost Base (Jun): | 2007 |
| Case 6 - Siemens Quench SNG Production w/ CO2 | | | | | MWe-net: | 2 |
| SNG (MMbtu/hr): | 6784 | | | | Capacity Factor (%): | 90 |
| OPERATING & MAINTENANCE LABOR | | | | | | |
| Operating Labor | | | | | | |
| Operating Labor Rate(base): | 34.65 | $/hour | | | | |
| Operating Labor Burden: | 30.00 | % of base | | | | |
| Labor O-H Charge Rate: | 25.00 | % of labor | | | | |
| | | | Total | | | |
| Operating Labor Requirements(O.J.)per Shift: | 1 unit/mod. | | Plant | | | |
| Skilled Operator | 2.0 | | 2.0 | | | |
| Operator | 12.0 | | 12.0 | | | |
| Foreman | 1.0 | | 1.0 | | | |
| Lab Tech's, etc. | 3.0 | | 3.0 | | | |
| TOTAL-O.J.'s | 18.0 | | 18.0 | | | |
| | | | | | Annual Cost | Annual Unit Cost |
| | | | | | $ | $/MMBtu |
| Annual Operating Labor Cost | | | | | $7,102,696 | $0.133 |
| Maintenance Labor Cost | | | | | $27,620,931 | $0.516 |
| Administrative & Support Labor | | | | | $8,680,907 | $0.162 |
| Property Taxes and Insurance | | | | | $56,265,151 | $1.052 |
| **TOTAL FIXED OPERATING COSTS** | | | | | **$99,669,685** | **$1.863** |
| VARIABLE OPERATING COSTS | | | | | | |
| | | | | | | $/MMBtu |
| **Maintenance Material Cost** | | | | | **$41,431,397** | **$0.77459** |
| Consumables | Consumption | | Unit | Initial Fill | | |
| | Initial Fill | /Day | Cost | Cost | | |
| **Water(/1000 gallons)** | 0 | 3,966 | 1.08 | **$0** | **$1,409,411** | **$0.02635** |
| **Chemicals** | | | | | | |
| MU & WT Chem.(lb) | 0 | 23,631 | 0.17 | $0 | $1,343,498 | $0.02512 |
| Carbon (Mercury Removal) (lb) | 221,425 | 379 | 1.05 | $232,533 | $130,800 | $0.00245 |
| COS Catalyst (m3) | 0 | 0 | 2,397.36 | $0 | $0 | $0.00000 |
| Water Gas Shift Catalyst (ft3) | 7,553 | 6.47 | 498.83 | $3,767,548 | $1,059,623 | $0.01981 |
| ZnO Sorbent (ton) | 57 | 0.20 | 12,574.00 | $721,842 | $812,073 | $0.01518 |
| Methanation Catalyst (ft3) | 12,102 | 9.21 | 440.00 | $5,324,846 | $1,331,211 | $0.02489 |
| Selexol Solution (gal) | 469,270 | 149 | 13.40 | $6,287,392 | $656,198 | $0.01227 |
| SCR Catalyst (m3) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Aqueous Ammonia (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Claus Catalyst (ft3) | w/equip | 1.53 | 131.27 | $0 | $66,058 | $0.00124 |
| **Subtotal Chemicals** | | | | **$16,334,162** | **$5,399,462** | **$0.10095** |
| **Other** | | | | | | |
| Supplemental Fuel (MBtu) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for consumption) (M\ | 0 | 0 | 61.60 | $0 | $0 | $0.00000 |
| Gases,N2 etc.(/100scf) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| L.P. Steam(/1000 pounds) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| **Subtotal Other** | | | | **$0** | **$0** | **$0.00000** |
| **Waste Disposal** | | | | | | |
| Spent Mercury Catalyst (lb.) | 0 | 379 | 0.42 | $0 | $51,948 | $0.00097 |
| Spent ZnO Sorbent (ton) | 0 | 0.20 | 16.23 | $0 | $1,048 | $0.00002 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Slag (ton) | 0 | 1,275 | 16.23 | $0 | $6,796,412 | $0.12706 |
| **Subtotal-Waste Disposal** | | | | **$0** | **$6,849,407** | **$0.12805** |
| **By-products & Emissions** | | | | | | |
| Sulfur (tons) | 0 | 108 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for sale) (MWh) | 0 | 43 | 58.00 | $0 | -$827,662 | -$0.01547 |
| **Subtotal By-Products** | | | | **$0** | **-$827,662** | **-$0.01547** |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$16,334,162** | **$54,262,014** | **$1.01446** |
| **Fuel (ton)** | 0 | 15,112 | 15.22 | **$0** | **$75,540,499** | **$1.41228** |

## 5.2 SYNTHETIC NATURAL GAS PRODUCTION USING NORTH DAKOTA LIGNITE COAL

### 5.2.1 Process Description

In this section the overall Siemens gasification process for SNG production using NDL coal is described. The plant configurations are similar to Cases 5 and 6, with the exception of the coal type utilized. Since the carbon content of the lignite coal used is lower than that of the bituminous coal and subbituminous, the coal feed rate is significantly increased. The process BFDs for the lignite coal non-sequestration and sequestration cases are shown in Exhibit 5-22 and Exhibit 5-24. The associated stream tables are shown in Exhibit 5-23 for the non-sequestration case and Exhibit 5-25 for the sequestration case.

**Exhibit 5-22 Case 7 Block Flow Diagram, North Dakota Lignite Coal to SNG without Carbon Sequestration**



## Exhibit 5-23 Case 7 Stream Table, North Dakota Lignite Coal to SNG without Carbon Sequestration

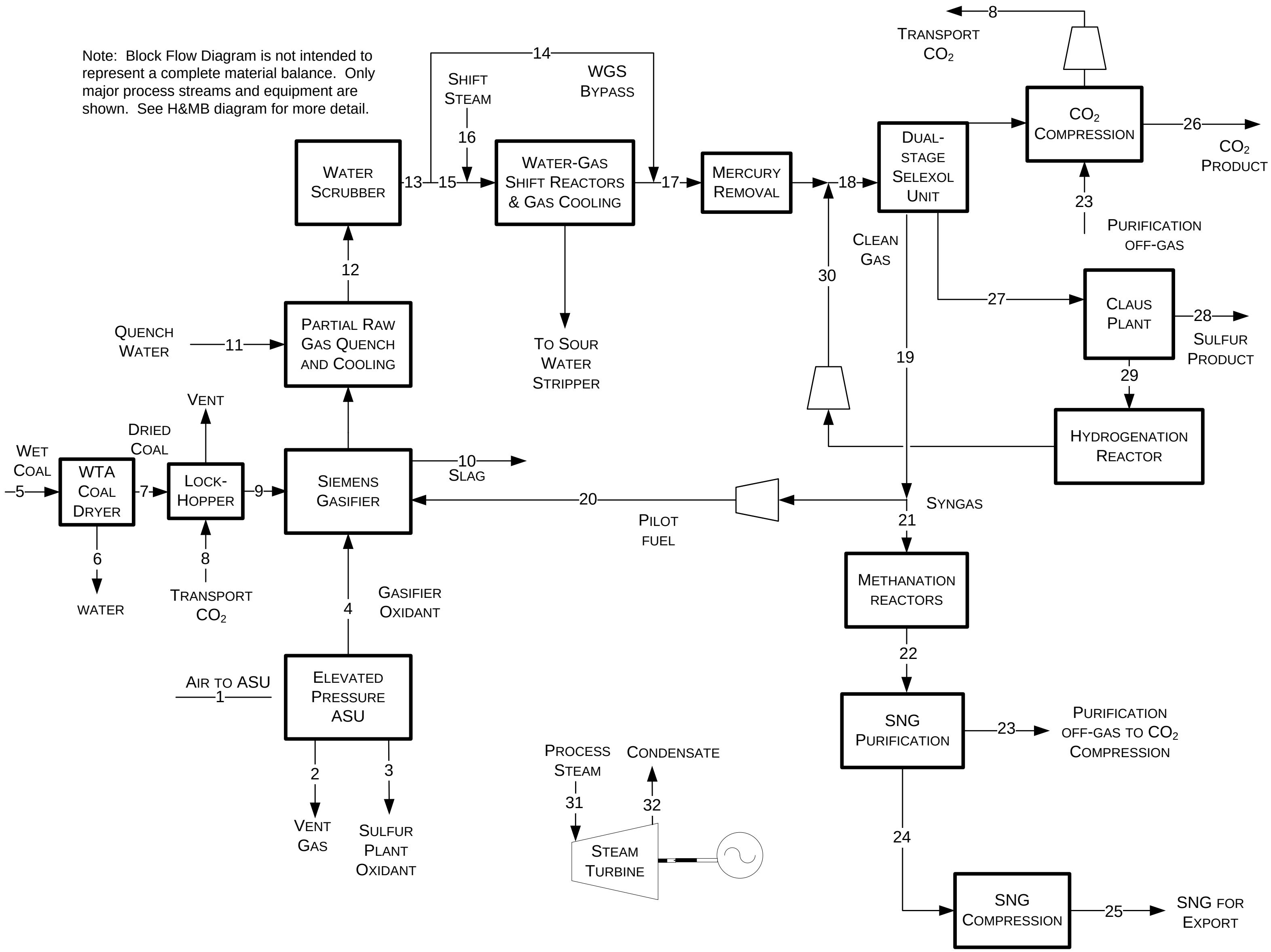
| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | |
| Ar | 0.0093 | 0.0091 | 0.0101 | 0.0101 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0021 | 0.0022 | 0.0022 | 0.0022 | 0.0000 | 0.0022 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0097 | 0.0013 | 0.0000 | 0.0000 | 0.4252 | 0.4405 | 0.4405 | 0.4405 | 0.0000 | 0.1549 |
| $CO_2$ | 0.0003 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.9785 | 0.1363 | 0.0000 | 0.0000 | 0.0718 | 0.0744 | 0.0744 | 0.0744 | 0.0000 | 0.3705 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0003 | 0.0003 | 0.0003 | 0.0003 | 0.0000 | 0.0001 |
| $H_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0103 | 0.5349 | 0.0000 | 0.0000 | 0.1607 | 0.1665 | 0.1665 | 0.1665 | 0.0000 | 0.4644 |
| $H_2O$ | 0.0064 | 0.0026 | 0.0000 | 0.0000 | 0.0000 | 1.0000 | 0.0000 | 0.0013 | 0.1899 | 0.0000 | 1.0000 | 0.3344 | 0.3103 | 0.3103 | 0.3103 | 1.0000 | 0.0018 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0024 | 0.0025 | 0.0025 | 0.0025 | 0.0000 | 0.0027 |
| $N_2$ | 0.7759 | 0.9803 | 0.0004 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0088 | 0.0000 | 0.0000 | 0.0032 | 0.0033 | 0.0033 | 0.0033 | 0.0000 | 0.0033 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.2081 | 0.0076 | 0.9895 | 0.9895 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1288 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 52,805 | 41,791 | 79 | 10,707 | 0 | 10,782 | 0 | 5,042 | 18,101 | 0 | 12,568 | 51,888 | 50,077 | 5,789 | 44,288 | 14,233 | 49,085 |
| V-L Flowrate (kg/hr) | 1,525,805 | 1,175,700 | 2,547 | 343,453 | 0 | 194,238 | 0 | 218,763 | 269,759 | 0 | 226,410 | 1,125,307 | 1,092,655 | 126,310 | 966,345 | 256,405 | 1,074,648 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 709,841 | 0 | 515,603 | 0 | 355,226 | 71,414 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 6 | 26 | 32 | 32 | 4 | 33 | 71 | 76 | 18 | 1,427 | 216 | 260 | 215 | 215 | 215 | 288 | 35 |
| Pressure (MPa, abs) | 0.09 | 0.11 | 0.86 | 0.86 | 0.10 | 0.35 | 0.09 | 5.62 | 5.62 | 4.24 | 8.27 | 3.89 | 3.756 | 3.76 | 3.76 | 4.14 | 3.34 |
| Enthalpy (kJ/kg) | 15.71 | 30.53 | 26.94 | 26.94 | --- | 147.56 | --- | 21.89 | 4,183.74 | --- | 1,017.87 | 1,092.44 | 964.077 | 964.08 | 964.08 | 3,070.04 | 40.47 |
| Density ($kg/m^3$) | 1.1 | 1.3 | 11.0 | 11.0 | --- | 986.2 | --- | 103.4 | 42.2 | --- | 782.0 | 19.3 | 20.5 | 20.5 | 20.5 | 18.2 | 29.2 |
| V-L Molecular Weight | 28.895 | 28.133 | 32.078 | 32.078 | --- | 18.015 | --- | 43.389 | 14.903 | --- | 18.015 | 21.687 | 21.820 | 21.820 | 21.820 | 18.015 | 21.894 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 116,415 | 92,133 | 175 | 23,605 | 0 | 23,770 | 0 | 11,115 | 39,906 | 0 | 27,707 | 114,394 | 110,400 | 12,762 | 97,638 | 31,378 | 108,214 |
| V-L Flowrate (lb/hr) | 3,363,825 | 2,591,975 | 5,614 | 757,185 | 0 | 428,222 | 0 | 482,290 | 594,716 | 0 | 499,149 | 2,480,878 | 2,408,892 | 278,465 | 2,130,426 | 565,276 | 2,369,194 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 1,564,932 | 0 | 1,136,710 | 0 | 783,138 | 157,442 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 42 | 78 | 90 | 90 | 40 | 91 | 160 | 168 | 65 | 2,600 | 420 | 500 | 418 | 418 | 418 | 550 | 95 |
| Pressure (psia) | 13.0 | 16.4 | 125.0 | 125.0 | 13.8 | 50.3 | 13.5 | 815.0 | 815.0 | 614.7 | 1,200.0 | 564.7 | 544.7 | 544.7 | 544.7 | 600.0 | 484.7 |
| Enthalpy (Btu/lb) | 6.8 | 13.1 | 11.6 | 11.6 | --- | 63.4 | --- | 9.4 | 1,798.7 | --- | 437.6 | 469.7 | 414.5 | 414.5 | 414.5 | 1,319.9 | 17.4 |
| Density ($lb/ft^3$) | 0.070 | 0.080 | 0.685 | 0.685 | --- | 61.569 | --- | 6.457 | 2.635 | --- | 48.817 | 1.203 | 1.282 | 1.282 | 1.282 | 1.135 | 1.822 |

A - Reference conditions are 32.02 F & 0.089 PSIA

## Exhibit 5-23 Case 7 Stream Table, North Dakota Lignite Coal to SNG without Carbon Sequestration (continued)

| | 18 | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | |
| Ar | 0.0022 | 0.0034 | 0.0034 | 0.0034 | 0.0118 | 0.0000 | 0.0138 | 0.0138 | 0.0000 | 0.0007 | 0.0000 | 0.0016 | 0.0024 | 0.0000 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.8145 | 0.0000 | 0.9559 | 0.9559 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.1532 | 0.2348 | 0.2348 | 0.2348 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0017 | 0.0543 | 0.0000 | 0.0861 | 0.0065 | 0.0000 | 0.0000 |
| $CO_2$ | 0.3756 | 0.0444 | 0.0444 | 0.0444 | 0.1478 | 0.9845 | 0.0026 | 0.0026 | 0.9925 | 0.5987 | 0.0000 | 0.4846 | 0.8126 | 0.0000 | 0.0000 |
| COS | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0019 | 0.0000 | 0.0003 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.4610 | 0.7121 | 0.7121 | 0.7121 | 0.0059 | 0.0000 | 0.0069 | 0.0069 | 0.0016 | 0.1041 | 0.0000 | 0.0507 | 0.1663 | 0.0000 | 0.0000 |
| $H_2O$ | 0.0018 | 0.0001 | 0.0001 | 0.0001 | 0.0023 | 0.0155 | 0.0000 | 0.0000 | 0.0041 | 0.0394 | 0.0000 | 0.3691 | 0.0009 | 1.0000 | 1.0000 |
| $H_2S$ | 0.0028 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.2003 | 0.0000 | 0.0014 | 0.0103 | 0.0000 | 0.0000 |
| $N_2$ | 0.0033 | 0.0051 | 0.0051 | 0.0051 | 0.0176 | 0.0000 | 0.0207 | 0.0207 | 0.0000 | 0.0005 | 0.0000 | 0.0007 | 0.0009 | 0.0000 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0054 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 49,657 | 31,944 | 183 | 31,762 | 9,214 | 1,363 | 7,851 | 7,851 | 11,969 | 701 | 0 | 820 | 572 | 40,826 | 56,932 |
| V-L Flowrate (kg/hr) | 1,095,675 | 327,418 | 1,874 | 325,544 | 189,733 | 59,420 | 130,313 | 130,313 | 524,399 | 25,095 | 0 | 25,423 | 21,027 | 735,487 | 1,025,654 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4,382 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | |
| Temperature (°C) | 34 | 31 | 80 | 31 | 35 | 35 | 35 | 35 | 5 | 48 | 175 | 232 | 21 | 566 | 33 |
| Pressure (MPa, abs) | 3.27 | 3.24 | 5.10 | 3.24 | 2.77 | 0.14 | 2.627 | 6.202 | 0.115 | 0.16 | 0.1 | 0.085 | 3.341 | 12.512 | 0.827 |
| Enthalpy (kJ/kg) | 39.13 | 83.91 | 228.57 | 83.91 | 39.74 | 45.04 | 47.668 | 12.706 | 7.673 | 98.02 | --- | 836.767 | -9.442 | 3,495.174 | 121.037 |
| Density (kg/$m^3$) | 28.9 | 13.0 | 17.4 | 13.0 | 23.6 | 2.4 | 17.9 | 44.7 | 2.2 | 2.2 | 5,285.7 | 0.6 | 58.2 | 34.9 | 995.1 |
| V-L Molecular Weight | 22.065 | 10.250 | 10.250 | 10.250 | 20.592 | 43.606 | 16.597 | 16.597 | 43.812 | 35.777 | --- | 30.993 | 36.771 | 18.015 | 18.015 |
| | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 109,475 | 70,425 | 403 | 70,022 | 20,314 | 3,004 | 17,309 | 17,309 | 26,388 | 1,546 | 0 | 1,808 | 1,261 | 90,005 | 125,515 |
| V-L Flowrate (lb/hr) | 2,415,550 | 721,834 | 4,131 | 717,703 | 418,290 | 130,998 | 287,291 | 287,291 | 1,156,103 | 55,324 | 0 | 56,047 | 46,356 | 1,621,472 | 2,261,180 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 9,660 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | |
| Temperature (°F) | 94 | 87 | 177 | 87 | 95 | 95 | 95 | 95 | 42 | 119 | 347 | 450 | 69 | 1,050 | 92 |
| Pressure (psia) | 474.7 | 469.6 | 740.0 | 469.6 | 401.1 | 20.1 | 381.0 | 899.5 | 16.7 | 23.7 | 17.3 | 12.3 | 484.5 | 1,814.7 | 120.0 |
| Enthalpy (Btu/lb) | 16.8 | 36.1 | 98.3 | 36.1 | 17.1 | 19.4 | 20.5 | 5.5 | 3.3 | 42.1 | --- | 359.7 | -4.1 | 1,502.7 | 52.0 |
| Density (lb/$ft^3$) | 1.803 | 0.811 | 1.088 | 0.811 | 1.473 | 0.148 | 1.115 | 2.789 | 0.137 | 0.137 | 330 | 0.039 | 3.636 | 2.177 | 62.122 |

### Exhibit 5-24 Case 8 Block Flow Diagram, North Dakota Lignite to SNG with Carbon Sequestration



## Exhibit 5-25 Case 8 Stream Table, North Dakota Lignite Coal to SNG with Carbon Sequestration

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | | | |
| Ar | 0.0093 | 0.0091 | 0.0101 | 0.0101 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0021 | 0.0022 | 0.0022 | 0.0022 | 0.0000 | 0.0022 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0037 | 0.0005 | 0.0000 | 0.0000 | 0.4252 | 0.4406 | 0.4406 | 0.4406 | 0.0000 | 0.1549 |
| $CO_2$ | 0.0003 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.9921 | 0.1371 | 0.0000 | 0.0000 | 0.0720 | 0.0746 | 0.0746 | 0.0746 | 0.0000 | 0.3708 |
| COS | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0003 | 0.0003 | 0.0003 | 0.0003 | 0.0000 | 0.0001 |
| $H_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0039 | 0.5346 | 0.0000 | 0.0000 | 0.1604 | 0.1663 | 0.1663 | 0.1663 | 0.0000 | 0.4642 |
| $H_2O$ | 0.0064 | 0.0026 | 0.0000 | 0.0000 | 0.0000 | 1.0000 | 0.0000 | 0.0001 | 0.1900 | 0.0000 | 1.0000 | 0.3344 | 0.3103 | 0.3103 | 0.3103 | 1.0000 | 0.0018 |
| $H_2S$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0024 | 0.0025 | 0.0025 | 0.0025 | 0.0000 | 0.0027 |
| $N_2$ | 0.7759 | 0.9803 | 0.0004 | 0.0004 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0088 | 0.0000 | 0.0000 | 0.0032 | 0.0033 | 0.0033 | 0.0033 | 0.0000 | 0.0033 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.2081 | 0.0076 | 0.9895 | 0.9895 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1289 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 52,781 | 41,772 | 75 | 10,706 | 0 | 10,782 | 0 | 4,996 | 18,078 | 0 | 12,563 | 51,861 | 50,047 | 5,784 | 44,263 | 14,236 | 49,057 |
| V-L Flowrate (kg/hr) | 1,525,100 | 1,175,157 | 2,411 | 343,429 | 0 | 194,238 | 0 | 218,763 | 269,759 | 0 | 226,325 | 1,125,197 | 1,092,505 | 126,252 | 966,253 | 256,473 | 1,074,531 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 709,841 | 0 | 515,603 | 0 | 355,226 | 71,414 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°C) | 6 | 26 | 32 | 32 | 4 | 33 | 71 | 75 | 18 | 1,427 | 216 | 260 | 215 | 215 | 215 | 288 | 35 |
| Pressure (MPa, abs) | 0.09 | 0.11 | 0.86 | 0.86 | 0.10 | 0.35 | 0.09 | 5.62 | 5.62 | 4.24 | 8.27 | 3.89 | 3.756 | 3.76 | 3.76 | 4.14 | 3.38 |
| Enthalpy (kJ/kg) | 15.71 | 30.53 | 26.94 | 26.94 | --- | 147.56 | --- | 18.43 | 4,187.50 | --- | 1,017.87 | 1,092.18 | 963.754 | 963.75 | 963.75 | 2,937.10 | 40.27 |
| Density ($kg/m^3$) | 1.1 | 1.3 | 11.0 | 11.0 | --- | 986.2 | --- | 105.6 | 42.3 | --- | 782.0 | 19.3 | 20.5 | 20.5 | 20.5 | 18.2 | 29.5 |
| V-L Molecular Weight | 28.895 | 28.133 | 32.078 | 32.078 | --- | 18.015 | --- | 43.785 | 14.922 | --- | 18.015 | 21.696 | 21.830 | 21.830 | 21.830 | 18.015 | 21.904 |
| | | | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 116,361 | 92,091 | 166 | 23,603 | 0 | 23,770 | 0 | 11,015 | 39,856 | 0 | 27,697 | 114,334 | 110,335 | 12,750 | 97,584 | 31,386 | 108,153 |
| V-L Flowrate (lb/hr) | 3,362,271 | 2,590,777 | 5,316 | 757,131 | 0 | 428,222 | 0 | 482,290 | 594,716 | 0 | 498,960 | 2,480,636 | 2,408,561 | 278,337 | 2,130,224 | 565,427 | 2,368,935 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 1,564,932 | 0 | 1,136,710 | 0 | 783,138 | 157,442 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | | | |
| Temperature (°F) | 42 | 78 | 90 | 90 | 40 | 91 | 160 | 166 | 64 | 2,600 | 420 | 500 | 418 | 418 | 418 | 550 | 95 |
| Pressure (psia) | 13.0 | 16.4 | 125.0 | 125.0 | 13.8 | 50.3 | 13.5 | 815.0 | 815.0 | 614.7 | 1,200.0 | 564.7 | 544.7 | 544.7 | 544.7 | 600.0 | 489.7 |
| Enthalpy (Btu/lb) | 6.8 | 13.1 | 11.6 | 11.6 | --- | 63.4 | --- | 7.9 | 1,800.3 | --- | 437.6 | 469.6 | 414.3 | 414.3 | 414.3 | 1,262.7 | 17.3 |
| Density ($lb/ft^3$) | 0.070 | 0.080 | 0.685 | 0.685 | --- | 61.569 | --- | 6.593 | 2.642 | --- | 48.817 | 1.204 | 1.282 | 1.282 | 1.282 | 1.135 | 1.842 |

A - Reference conditions are 32.02 F & 0.089 PSIA

## Exhibit 5-25 Case 8 Stream Table, North Dakota Lignite Coal to SNG with Carbon Sequestration (continued)

| | 18 | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| V-L Mole Fraction | | | | | | | | | | | | | | | |
| Ar | 0.0022 | 0.0034 | 0.0034 | 0.0034 | 0.0118 | 0.0000 | 0.0138 | 0.0138 | 0.0001 | 0.0007 | 0.0000 | 0.0016 | 0.0023 | 0.0000 | 0.0000 |
| $CH_4$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.8138 | 0.0000 | 0.9552 | 0.9552 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO | 0.1532 | 0.2348 | 0.2348 | 0.2348 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0037 | 0.0545 | 0.0000 | 0.0832 | 0.0072 | 0.0000 | 0.0000 |
| $CO_2$ | 0.3758 | 0.0444 | 0.0444 | 0.0444 | 0.1479 | 0.9845 | 0.0026 | 0.0026 | 0.9921 | 0.6020 | 0.0000 | 0.4928 | 0.8058 | 0.0000 | 0.0000 |
| COS | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0019 | 0.0000 | 0.0003 | 0.0000 | 0.0000 | 0.0000 |
| $H_2$ | 0.4609 | 0.7121 | 0.7121 | 0.7121 | 0.0065 | 0.0000 | 0.0077 | 0.0077 | 0.0039 | 0.1045 | 0.0000 | 0.0519 | 0.1788 | 0.0000 | 0.0000 |
| $H_2O$ | 0.0017 | 0.0001 | 0.0001 | 0.0001 | 0.0023 | 0.0155 | 0.0000 | 0.0000 | 0.0001 | 0.0393 | 0.0000 | 0.3670 | 0.0009 | 1.0000 | 1.0000 |
| $H_2S$ | 0.0028 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1964 | 0.0000 | 0.0015 | 0.0041 | 0.0000 | 0.0000 |
| $N_2$ | 0.0033 | 0.0051 | 0.0051 | 0.0051 | 0.0176 | 0.0000 | 0.0207 | 0.0207 | 0.0000 | 0.0005 | 0.0000 | 0.0007 | 0.0009 | 0.0000 | 0.0000 |
| $NH_3$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $O_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| $SO_2$ | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0011 | 0.0000 | 0.0000 | 0.0000 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 0.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| | | | | | | | | | | | | | | | |
| V-L Flowrate ($kg_{mol}$/hr) | 49,634 | 31,923 | 183 | 31,740 | 9,214 | 1,364 | 7,850 | 7,850 | 15,705 | 698 | 0 | 813 | 576 | 40,673 | 58,897 |
| V-L Flowrate (kg/hr) | 1,095,450 | 327,218 | 1,874 | 325,344 | 189,670 | 59,465 | 130,205 | 130,205 | 687,644 | 24,974 | 0 | 25,168 | 20,919 | 732,739 | 1,061,040 |
| Solids Flowrate (kg/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4,385 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | |
| Temperature (°C) | 34 | 31 | 80 | 31 | 35 | 35 | 35 | 35 | 35 | 48 | 175 | 232 | 21 | 566 | 33 |
| Pressure (MPa, abs) | 3.31 | 3.24 | 5.10 | 3.24 | 2.77 | 0.14 | 2.627 | 6.202 | 15.272 | 0.16 | 0.1 | 0.085 | 3.306 | 12.512 | 0.827 |
| Enthalpy (kJ/kg) | 38.97 | 83.90 | 228.55 | 83.90 | 39.77 | 45.02 | 47.726 | 12.794 | -209.181 | 97.84 | --- | 834.709 | -8.186 | 3,495.174 | 120.935 |
| Density ($kg/m^3$) | 29.2 | 13.0 | 17.4 | 13.0 | 23.6 | 2.4 | 17.8 | 44.6 | 780.4 | 2.2 | 5,286.1 | 0.6 | 56.4 | 34.9 | 995.1 |
| V-L Molecular Weight | 22.071 | 10.250 | 10.250 | 10.250 | 20.586 | 43.607 | 16.587 | 16.587 | 43.785 | 35.794 | --- | 30.956 | 36.297 | 18.015 | 18.015 |
| | | | | | | | | | | | | | | | |
| V-L Flowrate ($lb_{mol}$/hr) | 109,424 | 70,378 | 403 | 69,975 | 20,313 | 3,006 | 17,306 | 17,306 | 34,623 | 1,538 | 0 | 1,792 | 1,271 | 89,669 | 129,845 |
| V-L Flowrate (lb/hr) | 2,415,054 | 721,392 | 4,131 | 717,261 | 418,150 | 131,097 | 287,053 | 287,053 | 1,515,996 | 55,059 | 0 | 55,485 | 46,119 | 1,615,412 | 2,339,194 |
| Solids Flowrate (lb/hr) | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 9,667 | 0 | 0 | 0 | 0 |
| | | | | | | | | | | | | | | | |
| Temperature (°F) | 94 | 87 | 177 | 87 | 95 | 95 | 95 | 95 | 95 | 119 | 347 | 450 | 69 | 1,050 | 91 |
| Pressure (psia) | 479.5 | 469.6 | 740.0 | 469.6 | 401.1 | 20.1 | 381.0 | 899.5 | 2,215.0 | 23.7 | 17.3 | 12.3 | 479.5 | 1,814.7 | 120.0 |
| Enthalpy (Btu/lb) | 16.8 | 36.1 | 98.3 | 36.1 | 17.1 | 19.4 | 20.5 | 5.5 | -89.9 | 42.1 | --- | 358.9 | -3.5 | 1,502.7 | 52.0 |
| Density ($lb/ft^3$) | 1.822 | 0.811 | 1.089 | 0.811 | 1.473 | 0.148 | 1.114 | 2.787 | 48.716 | 0.137 | 330 | 0.039 | 3.524 | 2.177 | 62.123 |

### 5.2.2 Key System Assumptions

System assumptions for Cases 7 and 8, Siemens gasifier using NDL coal with and without carbon sequestration, are compiled in Exhibit 5-26.

**Exhibit 5-26 Cases 7 and 8 Plant Study Configuration Matrix**

| Case | Case 7 | Case 8 |
|---|---|---|
| Gasifier Pressure, MPa (psia) | 4.2 (615) | 4.2 (615) |
| $O_2$:Coal Ratio, kg $O_2$/kg dry coal | 0.657 | 0.657 |
| Carbon Conversion, % | 99.9 | 99.9 |
| Syngas HHV at Gasifier Outlet, $kJ/Nm^3$ (Btu/scf) | 9,350 (251) | 9,350 (251) |
| Steam Cycle, MPa/°C/°C (psig/°F/°F) | 12.4/566/536 (1800/1050/997) | 12.4/566/536 (1800/1050/997) |
| Condenser Pressure, mm Hg (in Hg) | 36 (1.4) | 36 (1.4) |
| Combustion Turbine | N/A | N/A |
| Gasifier Technology | Siemens | Siemens |
| Oxidant | 99 vol% Oxygen | 99 vol% Oxygen |
| Coal | ND Lignite | ND Lignite |
| Coal Feed Moisture Content, % | 36.08 | 36.08 |
| COS Hydrolysis | No | No |
| Water Gas Shift | Yes | Yes |
| $H_2S$ Separation | Selexol (1st Stage) | Selexol (1st Stage) |
| Sulfur Removal, % | 98.5 | 99.9 |
| $CO_2$ Separation | Selexol (2nd Stage) | Selexol (2nd Stage) |
| $CO_2$ Sequestration, % | N/A | 67.2 |
| Sulfur Recovery | Claus Plant with Tail Gas Treatment / Elemental Sulfur | Claus Plant with Tail Gas Treatment / Elemental Sulfur |
| Methanation System | Based on Haldor Topsoe TREMP™ Process | Based on Haldor Topsoe TREMP™ Process |
| Particulate Control | Scrubber and AGR Absorber | Scrubber and AGR Absorber |
| Mercury Control | Carbon Bed | Carbon Bed |
| NOx Control | N/A | N/A |

**Balance of Plant and Sparing Philosophy**

The balance of plant assumptions and sparing philosophy are common to all cases and are presented in Section 4.1.3.

### 5.2.3 Cases 7 and 8 Performance Results

The Siemens SNG plant without and with carbon sequestration and using NDL coal produces a net output of 52 Bscf/year at 90 percent capacity. The HHV conversion efficiency of both the non-sequestration case and the sequestration case is 61.5 percent.

Overall performance for the two plants is summarized in Exhibit 5-27, which includes auxiliary power requirements and production values. The ASU accounts for approximately 58 percent of the total auxiliary load in non-sequestration case and 48 percent in the sequestration case, distributed between the main air compressor, the oxygen compressor, and ASU auxiliaries. $CO_2$ compression accounts for about 19 percent of the auxiliary load in Case 8. The AGR process accounts for about 12 percent and 10 percent of the auxiliary load for the non-sequestration and sequestration cases, respectively. The WTA coal drying process accounts for 10 percent in the non-sequestration case and 8 percent in the sequestration case. All other individual auxiliary loads are less than 3 percent of the total.

## Exhibit 5-27 Cases 7 and 8 Plant Performance Summary

| **POWER SUMMARY (Gross Power at Generator Terminals, kWe)** | **Case 7** | **Case 8** |
|---|---|---|
| Steam Turbine Power | 305,800 | 310,500 |
| **TOTAL POWER, kWe** | **305,800** | **310,500** |
| **AUXILIARY LOAD SUMMARY, kWe** | | |
| Coal Handling | 990 | 990 |
| Coal Milling | 7,300 | 7,300 |
| Slag Handling | 1,860 | 1,860 |
| WTA Coal Dryer Compressor | 25,020 | 25,020 |
| WTA Coal Dryer Auxiliaries | 880 | 880 |
| Air Separation Unit Auxiliaries | 1,000 | 1,000 |
| Air Separation Unit Main Air Compressor | 127,710 | 127,650 |
| Oxygen Compressor | 21,850 | 21,850 |
| $CO_2$ Compressor | 0 | 58,880 |
| $CO_2$ Solid Feed System Compressor | 6,880 | 330 |
| SNG Compressors | 5,680 | 5,680 |
| Methanation Plant Recycle Compressor | 4,420 | 4,340 |
| Gasifier Pilot Fuel Compressor | 80 | 80 |
| Boiler Feedwater Pumps | 4,700 | 4,730 |
| Condensate Pump | 370 | 390 |
| Quench Water Pump | 570 | 570 |
| Circulating Water Pump | 4,550 | 5,250 |
| Ground Water Pumps | 400 | 460 |
| Cooling Tower Fans | 2,790 | 3,420 |
| Air Cooled Condenser Fans | 3,440 | 3,820 |
| Scrubber Pumps | 940 | 940 |
| Acid Gas Removal | 31,200 | 31,200 |
| Steam Turbine Auxiliaries | 100 | 100 |
| Claus Plant/TGTU Auxiliaries | 250 | 250 |
| Claus Plant TG Recycle Compressor | 2,130 | 2,150 |
| Miscellaneous Balance of Plant[2] | 3,000 | 3,000 |
| Transformer Losses | 1,630 | 1,790 |
| **TOTAL AUXILIARIES, kWe** | **259,740** | **313,930** |
| **NET POWER, kWe** | **46,060** | **-3,430** |
| SNG Production Rate, $MNm^3$/hr (Mscf/hr) | 185.9 (6,568) | 185.9 (6,568) |
| Capacity Factor | 90% | 90% |
| Net Exported Power Efficiency (HHV) | 1.5% | -0.1% |
| SNG Conversion Efficiency ($HHV_{product}/HHV_{coal}$), % | 61.5% | 61.5% |
| Coal to SNG Product Yield, $MNm^3_{SNG}/tonne_{coal}$ ($Mscf_{SNG}/ton_{coal}$) | 0.25 (8.4) | 0.25 (8.4) |
| Coal Feed Flow Rate, kg/hr (lb/hr) | 709,841 (1,564,932) | 709,841 (1,564,932) |
| Thermal Input,[1] kWth | 3,034,796 | 3,034,796 |
| Condenser Cooling Duty, GJ/hr (MMBtu/hr) | 1,635 (1,550) | 1,699 (1,610) |
| Raw Water Withdrawal, $m^3$/min (gpm) | 16.7 (4,421) | 19.4 (5,131) |
| Raw Water Consumption, $m^3$/min (gpm) | 12.7 (3,350) | 14.8 (3,901) |

1 - HHV of NDL coal is 15,391 kJ/kg (6,617 Btu/lb)

2 - Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

**Environmental Performance**

The environmental targets for emissions of Hg, NOx, $SO_2$, and PM were presented in Section 2.5. A summary of the plant air emissions for Cases 7 and 8 is presented in Exhibit 5-28.

**Exhibit 5-28 Cases 7 and 8 Air Emissions**

| | kg/GJ ($lb/10^6$ Btu) | | Tonne/year (ton/year) 90% capacity factor | |
|---|---|---|---|---|
| | **Case 7** | **Case 8** | **Case 7** | **Case 8** |
| **$SO_2$** | 0.010 (0.024) | 0.000 (0.000) | 886 (977) | 0 (0) |
| **$NO_X$** | Negligible | Negligible | Negligible | Negligible |
| **Particulates** | 0.003 (0.0071) | 0.003 (0.0071) | 263 (290) | 263 (290) |
| **Hg** | 2.41E-7 (5.60E-7) | 2.41E-7 (5.60E-7) | 0.021 (0.023) | 0.021 (0.023) |
| **$CO_2$** | 63.2 (147.0) | 0.4 (0.9) | 5,443,126 (6,000,019) | 34,303 (37,813) |

The low level of $SO_2$ emissions in Case 8 is achieved by capture of the sulfur in the gas by the two-stage Selexol AGR process and co-sequestration with the $CO_2$ product. Because the WTA coal drying process does not include combustion of SNG, there are no $SO_2$ emissions associated with drying. The clean syngas exiting the AGR process has a sulfur concentration of approximately 2 ppmv in both cases. The $H_2S$-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The tail gas is hydrogenated and recycled to the AGR system. The higher sulfur emissions in Case 7 are due to the sulfur contained in the $CO_2$-rich stream from the Selexol process, which is vented rather than sequestered.

$NO_X$ emissions are negligible because no CT is used for power generation.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Ninety five percent of mercury is captured from the syngas by an activated carbon bed. $CO_2$ emissions represent the uncontrolled (Case 7) and controlled (Case 8) discharge from the process.

The carbon balance for the two cases is shown in Exhibit 5-29. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not used in the carbon capture equation below, but it is not neglected in the balance since Aspen accounts for air components throughout. Carbon leaves the plant as unburned carbon in the slag, $CO_2$ in the stack gas, ASU vent gas, SNG product, and the $CO_2$ product (either vented or sequestered).

**Exhibit 5-29 Cases 7 and 8 Carbon Balance**

| Carbon In, kg/hr (lb/hr) | | | Carbon Out, kg/hr (lb/hr) | | |
|---|---|---|---|---|---|
| | **Case 7** | **Case 8** | | **Case 7** | **Case 8** |
| **Coal** | 280,768 (618,988) | 280,768 (618,988) | **Slag** | 1,404 (3,095) | 1,404 (3,095) |
| **Air ($CO_2$)** | 208 (458) | 208 (458) | **SNG Purification Off-Gas** | 16,113 (35,522) | 0 (0) |
| | | | **ASU Vent** | 208 (458) | 208 (458) |
| | | | **SNG** | 90,390 (199,276) | 90,309 (199,098) |
| | | | **$CO_2$ Vent** | 142,939 (315,126) | 187,863 (414,167) |
| | | | **$CO_2$ Feed Vent** | 29,923 (65,969) | 1,192 (2,628) |
| | | | **Stack** | N/A | 0 (0) |
| **Total** | 280,976 (619,447) | 280,976 (619,447) | **Total** | 280,976 (619,447) | 280,976 (619,447) |

The carbon capture efficiency is presented in two distinct ways. The first way defines capture as the amount of carbon in the $CO_2$ product for sequestration relative to the amount of carbon in the coal less carbon contained in the slag, and is represented by the following fraction:

(Carbon in Product for Sequestration)/[(Carbon in the Coal)-(Carbon in Slag)] or

Non sequestration (Case 7)

(414,167)/(618,988 – 3,095)*100 = 67.2% (Case 8)

The second way does not penalize the production facility for the carbon converted to SNG product, as the end use of the SNG is unknown. In a carbon constrained scenario, the SNG may or may not be used in a scenario where CCS is implemented. For this method, the burden of carbon mitigation falls on the end-user. This method is represented by the following fraction:

[(Carbon in Product for Sequestration) + (Carbon in SNG Product)]/[(Carbon in the Coal)-(Carbon in Slag)] or

(0 + 199,276)/(618,988 – 3,095)*100 = 32.4% (Case 7)

(414,167 + 199,098)/(618,988 – 3,095)*100 = 99.6% (Case 8)

Exhibit 5-30 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur in the SNG product, sulfur emitted in the stack gas, and sulfur co-sequestered with the $CO_2$ product. Sulfur in the slag is considered negligible. The total sulfur capture is represented by the following fraction:

(Sulfur byproduct + Sulfur in $CO_2$ product)/Sulfur in the coal or
(9,660 + 0)/9,803 = 98.5 percent (Case 7)
(9,667 + 130)/9,803 = 99.9 percent (Case 8)

**Exhibit 5-30 Cases 7 and 8 Sulfur Balance**

| Sulfur In, kg/hr (lb/hr) | | | Sulfur Out, kg/hr (lb/hr) | | |
|---|---|---|---|---|---|
| | **Case 7** | **Case 8** | | **Case 7** | **Case 8** |
| **Coal** | 4,447 (9,803) | 4,447 (9,803) | **Elemental Sulfur** | 4,382 (9,660) | 4,385 (9,667) |
| | | | **SNG** | 2 (5) | 2 (5) |
| | | | **$CO_2$ Vent Streams/Stack** | 63 (138) | 0.4 (1) |
| | | | **$CO_2$ Product** | N/A | 59 (130) |
| **Total** | 4,447 (9,803) | 4,447 (9,803) | **Total** | 4,447 (9,803) | 4,447 (9,803) |

Exhibit 5-31 shows the overall water balance for the plant. Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as syngas condensate, and is re-used as internal recycle. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is defined as the water removed from the ground or diverted from a surface-water source for use in the plant and was assumed to be provided 50 percent by a POTW and 50 percent from groundwater. Raw water withdrawal can be represented by the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, quench system makeup, and slag handling makeup. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source.

**Exhibit 5-31 Cases 7 and 8 Water Balance**

| | Case 7 | Case 8 |
|---|---|---|
| **Water Demand, $m^3$/min (gpm)** | | |
| **Slag Handling** | **1.5 (409)** | **1.5 (409)** |
| **Quench/Wash** | **3.8 (998)** | **3.8 (998)** |
| **Condenser Makeup** | **4.5 (1,189)** | **4.5 (1,191)** |
| Gasifier Steam | 0.0 (0) | 0.0 (0) |
| Shift Steam | 4.3 (1,131) | 4.3 (1,131) |
| BFW Makeup | 0.22 (59) | 0.23 (60) |
| **Cooling Tower** | **17.7 (4,683)** | **20.4 (5,392)** |
| **Total** | **27.6 (7,279)** | **30.2 (7,990)** |
| **Internal Recycle, $m^3$/min (gpm)** | | |
| **Slag Handling** | **1.5 (409)** | **1.5 (409)** |
| **Quench/Wash** | **3.8 (998)** | **3.8 (998)** |
| **Condenser Makeup** | **0.0 (0)** | **0.0 (0)** |
| **Cooling Tower** | **5.5 (1,451)** | **5.5 (1,452)** |
| Water from Coal Drying | 3.2 (856) | 3.2 (856) |
| BFW Blowdown | 0.22 (59) | 0.23 (60) |
| SWS Blowdown | 0.67 (177) | 0.67 (177) |
| SWS Excess | 1.4 (359) | 1.4 (359) |
| **Total** | **10.8 (2,858)** | **10.8 (2,859)** |
| **Raw Water Withdrawal, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.0 (0)** | **0.0 (0)** |
| **Quench/Wash** | **0.0 (0)** | **0.0 (0)** |
| **Condenser Makeup** | **4.5 (1,189)** | **4.5 (1,191)** |
| Gasifier Steam | 0.0 (0) | 0.0 (0) |
| Shift Steam | 4.3 (1,131) | 4.3 (1,131) |
| BFW Makeup | 0.22 (59) | 0.23 (60) |
| **Cooling Tower** | **12.2 (3,232)** | **14.9 (3,940)** |
| **Total** | **16.7 (4,421)** | **19.4 (5,131)** |
| **Process Water Discharge, $m^3$/min (gpm)** | | |
| **SWS Blowdown** | **0.07 (18)** | **0.07 (18)** |
| **Cooling Tower Blowdown** | **4.0 (1,053)** | **4.6 (1,213)** |
| **Total** | **4.1 (1,071)** | **4.7 (1,230)** |
| **Raw Water Consumption, $m^3$/min (gpm)** | | |
| **Slag Handling** | **0.0 (0)** | **0.0 (0)** |
| **Quench/Wash** | **0.0 (0)** | **0.0 (0)** |
| **SWS Blowdown** | **-0.07 (-18)** | **-0.07 (-18)** |
| **Condenser Makeup** | **4.5 (1,189)** | **4.5 (1,191)** |
| **Cooling Tower** | **8.2 (2,179)** | **10.3 (2,728)** |
| **Total** | **12.7 (3,350)** | **14.8 (3,901)** |

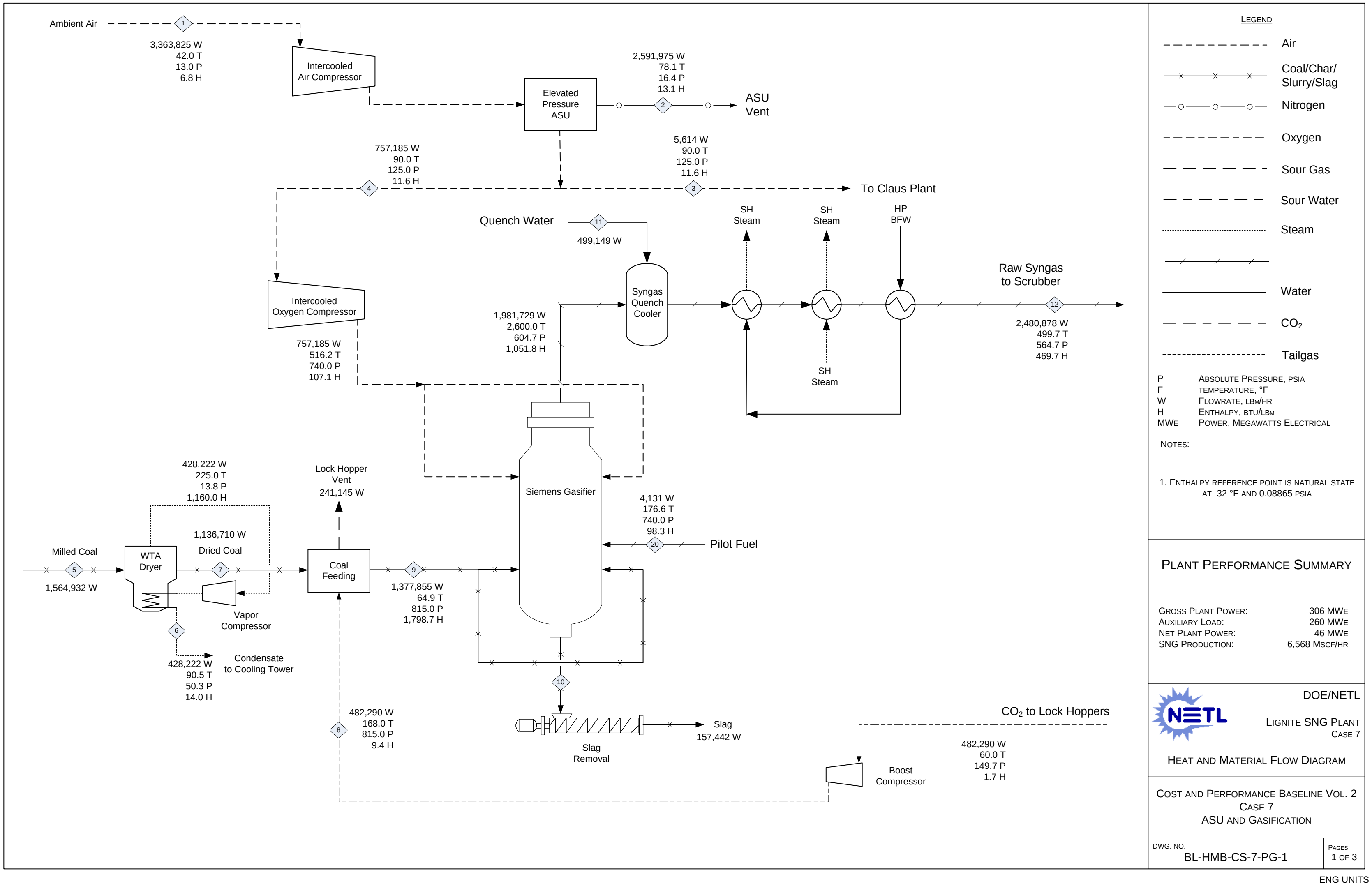
**Heat and Mass Balance Diagrams**

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 5-32 through Exhibit 5-37:

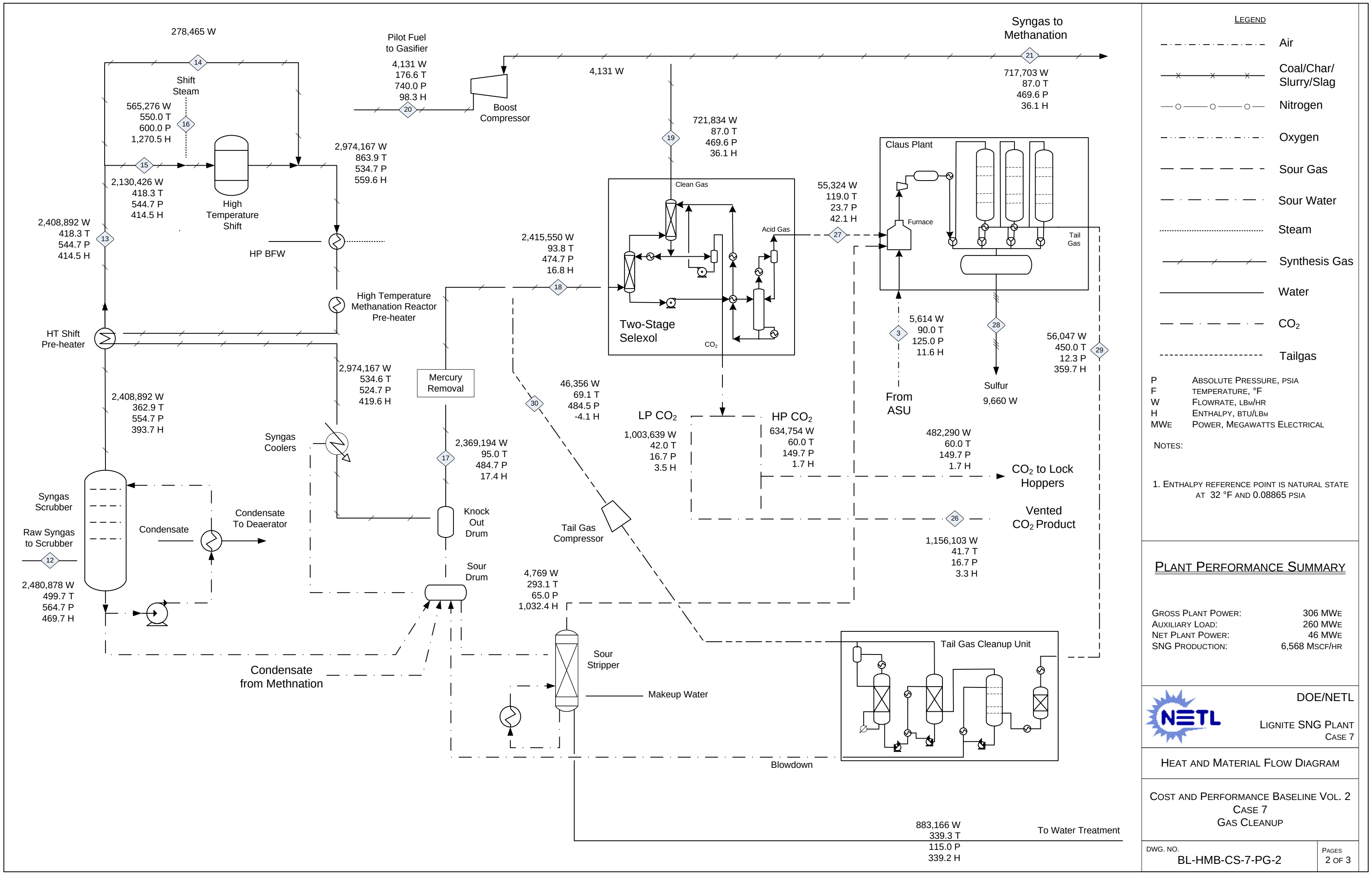
- Coal gasification and ASU
- ,Syngas cleanup (including sulfur recovery and tail gas recycle)
- Methanation, SNG purification, SNG compression and power cycle

An overall plant energy balance is provided in tabular form in Exhibit 5-38 for the two cases.

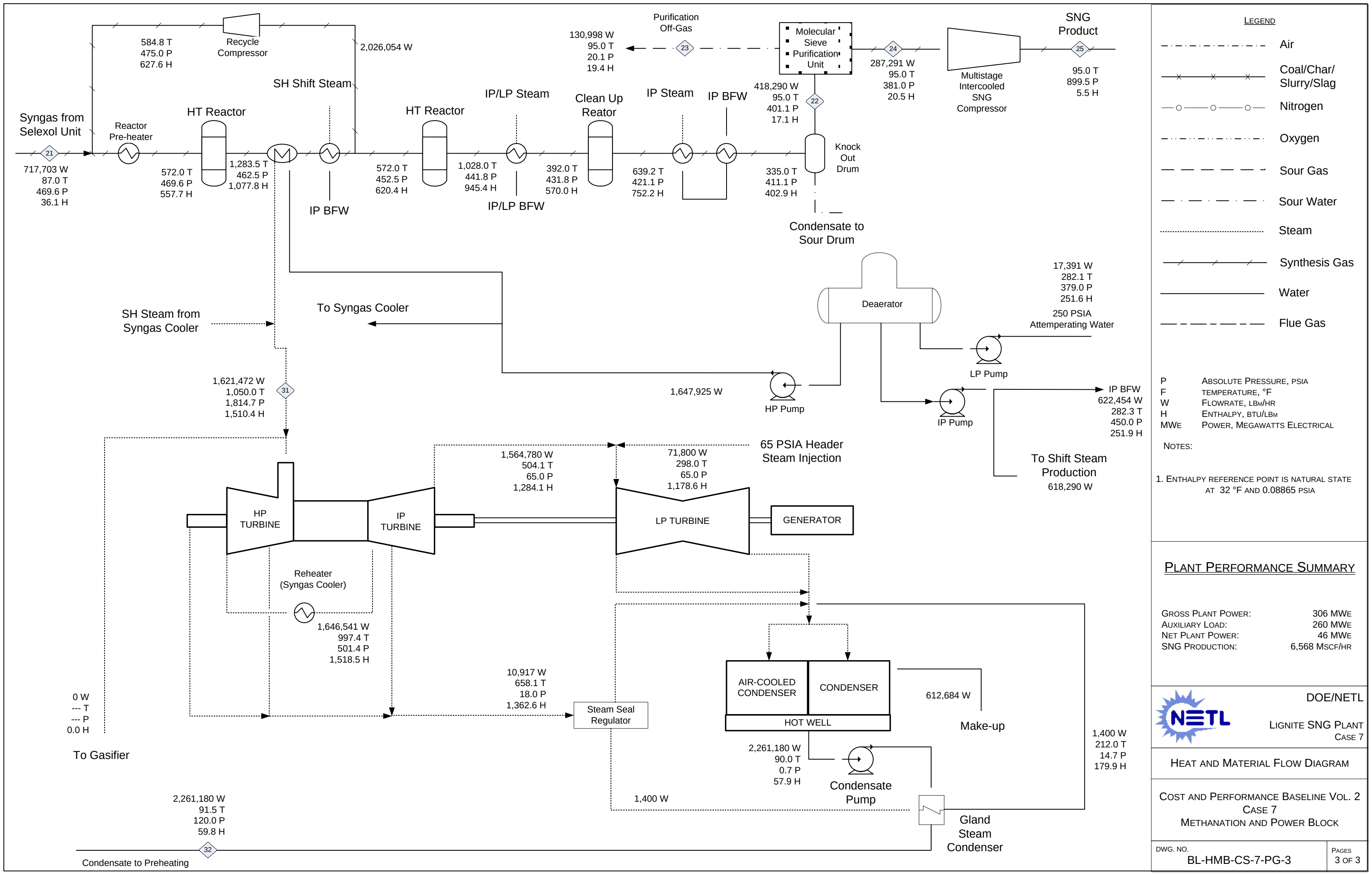
# Exhibit 5-32 Case 7 Gasification and ASU Heat and Mass Balance



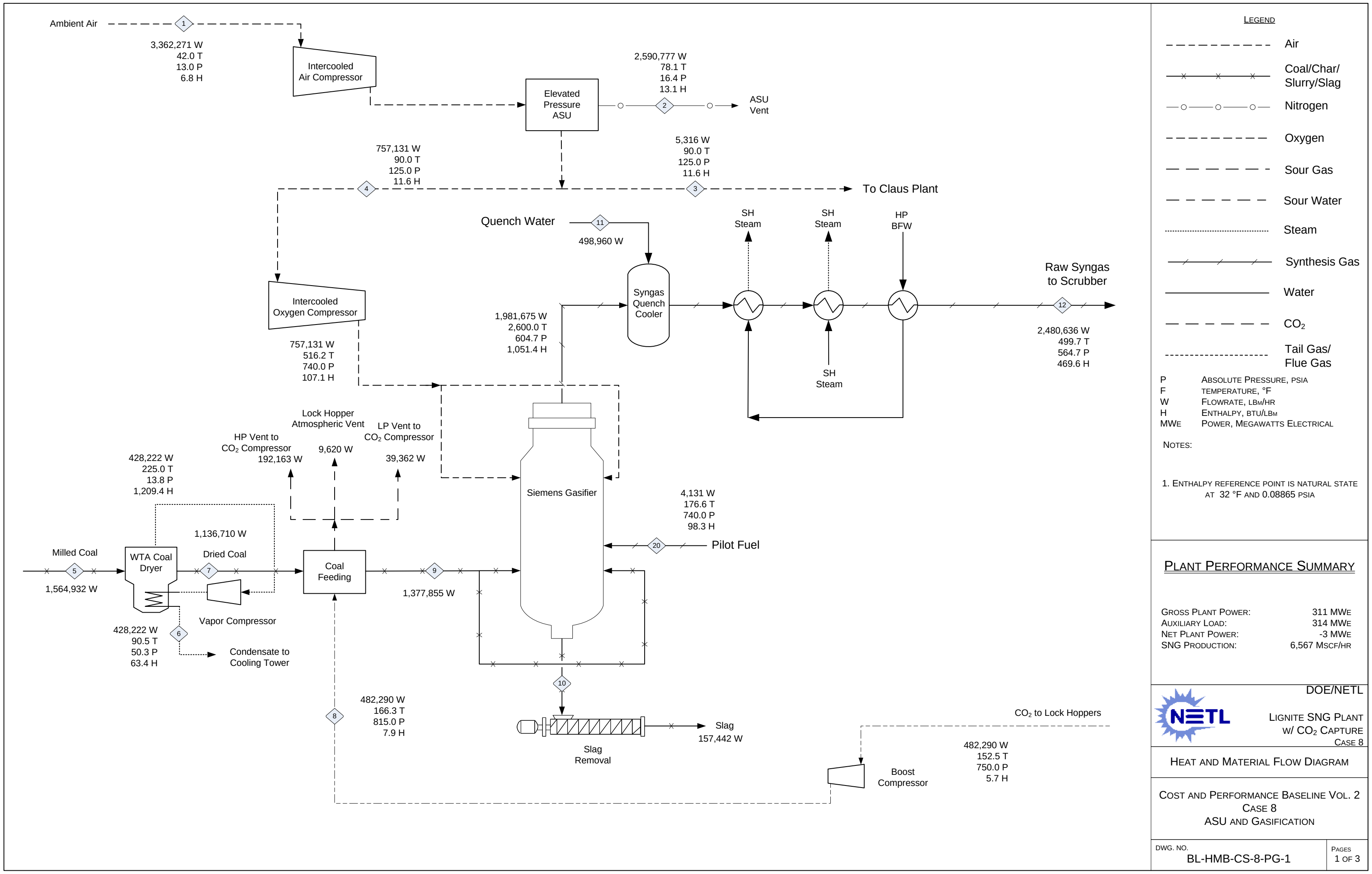
# Exhibit 5-33 Case 7 Gas Cleanup Heat and Mass Balance



# Exhibit 5-34 Case 7 Methanation and Power Block Heat and Mass Balance



## Exhibit 5-35 Case 8 Gasification and ASU Heat and Mass Balance

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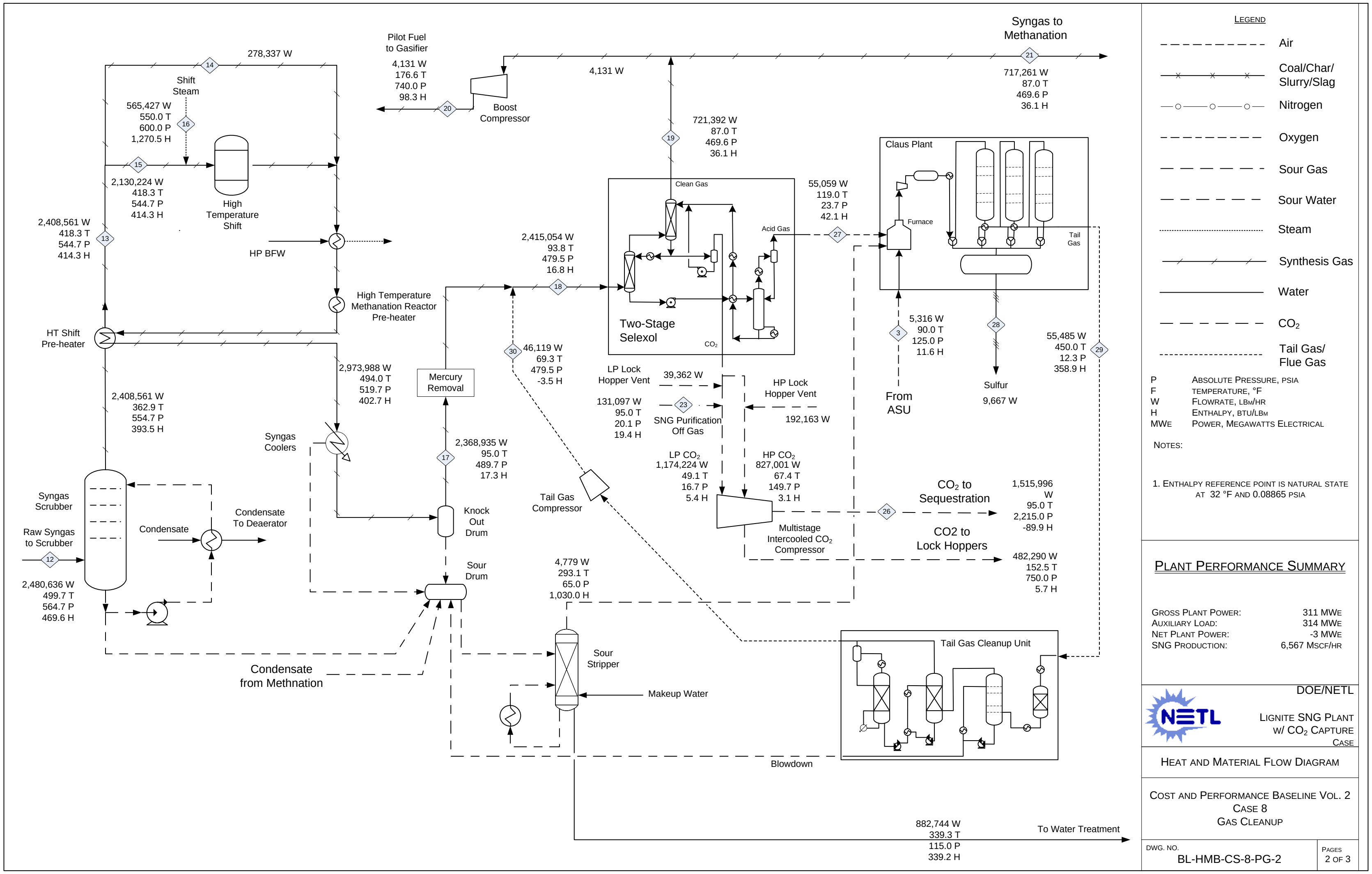
# Exhibit 5-36 Case 8 Gas Cleanup Heat and Mass Balance



## Exhibit 5-37 Case 8 Methanation and Power Block Heat and Mass Balance

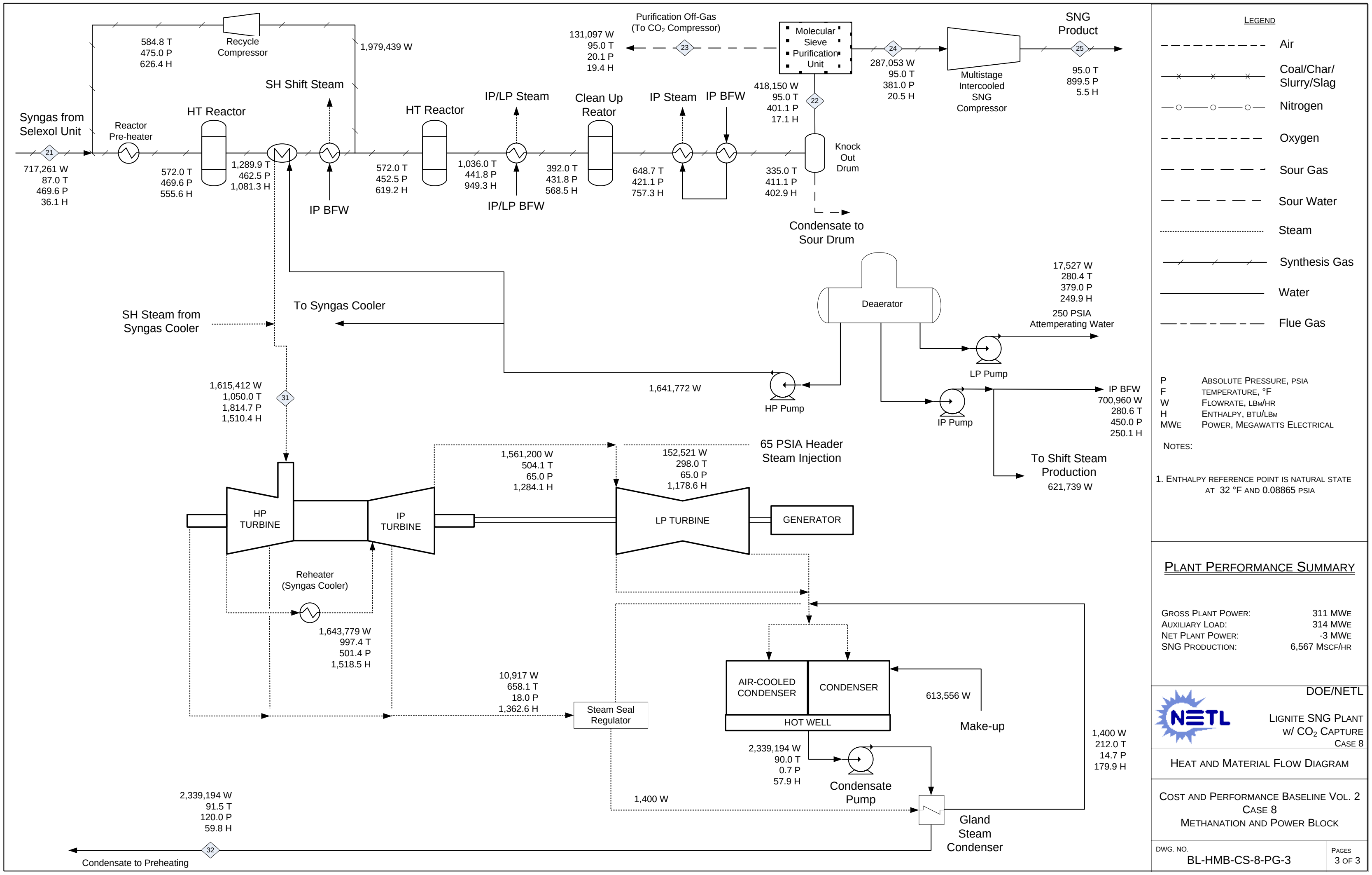


**Exhibit 5-38  Cases 7 and 8 Energy Balance**

| | HHV | | Sensible + Latent | | Power | | Total | |
|---|---|---|---|---|---|---|---|---|
| | **Case 7** | **Case 8** | **Case 7** | **Case 8** | **Case 7** | **Case 8** | **Case 7** | **Case 8** |
| **Heat In, GJ/hr (MMBtu/hr)** | | | | | | | | |
| Coal | 10,925 (10,355) | 10,925 (10,355) | 6.3 (6.0) | 7.9 (7.5) | 0 (0) | 0 (0) | 10,932 (10,361) | 10,933 (10,363) |
| ASU Air | 0 (0) | 0 (0) | 24.0 (22.7) | 24.0 (22.7) | 0 (0) | 0 (0) | 24 (23) | 24 (23) |
| Raw Water Makeup | 0 (0) | 0 (0) | 18.6 (17.6) | 27.0 (25.6) | 0 (0) | 0 (0) | 19 (18) | 27 (26) |
| Auxiliary Power | 0 (0) | 0 (0) | 0.0 (0.0) | 0.0 (0.0) | 935 (886) | 1,130 (1,071) | 935 (886) | 1,130 (1,071) |
| **Totals** | 10,925 (10,355) | 10,925 (10,355) | 48.9 (46.4) | 58.9 (55.8) | 935 (886) | 1,130 (1,071) | 11,909 (11,288) | 12,114 (11,482) |
| Heat Out, GJ/hr (MMBtu/hr) | | | | | | | | |
| ASU Intercoolers | 0 (0) | 0 (0) | 334 (317) | 334 (317) | 0 (0) | 0 (0) | 334 (317) | 334 (317) |
| ASU Vent | 0 (0) | 0 (0) | 35.9 (34.0) | 35.9 (34.0) | 0 (0) | 0 (0) | 36 (34) | 36 (34) |
| Slag | 46 (44) | 46 (44) | 120.5 (114.2) | 120.5 (114.2) | 0 (0) | 0 (0) | 167 (158) | 167 (158) |
| Sulfur | 41 (38) | 41 (39) | 0.5 (0.5) | 0.5 (0.5) | 0 (0) | 0 (0) | 41 (39) | 41 (39) |
| $CO_2$ Intercoolers | 0 (0) | 0 (0) | 0 (0) | 351.7 (333.4) | 0 (0) | 0 (0) | 0 (0) | 352 (333) |
| $CO_2$ | 0 (0) | 0 (0) | 2.4 (2.3) | -143.8 (-136.3) | 0 (0) | 0 (0) | 2 (2) | -144 (-136) |
| Cooling Tower Blowdown | 0 (0) | 0 (0) | 21.6 (20.5) | 25.6 (24.2) | 0 (0) | 0 (0) | 22 (20) | 26 (24) |
| SNG | 6,703 (6,353) | 6,699 (6,349) | 1.7 (1.6) | 1.7 (1.6) | 0 (0) | 0 (0) | 6,705 (6,355) | 6,700 (6,351) |
| SNG Purification Off-Gas | 0 (0) | 0 (0) | 3 (3) | 3 (3) | 0 (0) | 0 (0) | 3 (3) | 3 (3) |
| Condenser | 0 (0) | 0 (0) | 1,631 (1,546) | 1,703 (1,614) | 0 (0) | 0 (0) | 1,631 (1,546) | 1,703 (1,614) |
| Non-Condenser Cooling Loads | 0 (0) | 0 (0) | 879 (833) | 1,194 (1,131) | 0 (0) | 0 (0) | 879 (833) | 1,194 (1,131) |
| Process Losses* | 0 (0) | 0 (0) | 989 (938 | 586 (555) | 0 (0) | 0 (0) | 989 (938) | 586 (555) |
| Power | 0 (0) | 0 (0) | 0.0 (0.0) | 0.0 (0.0) | 1,101 (1,043) | 1,118 (1,059) | 1,101 (1,043) | 1,118 (1,059) |
| **Totals** | **6,790 (6,435)** | **6,785 (6,431)** | **4,019 (3,809)** | **4,211 (3,991)** | **1,101 (1,043)** | **1,118 (1,059)** | **11,909 (11,288)** | **12,114 (11,482)** |

* Includes other energy losses not explicitly accounted for in the model.

### 5.2.4 Cases 7 and 8 Equipment Lists

Major equipment items for the Siemens gasifier SNG plant without and with carbon sequestration using NDL coal are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 5.2.5. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

**ACCOUNT 1 COAL HANDLING**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Bottom Trestle Dumper and Receiving Hoppers | N/A | 2 | 0 | 181 tonne (200 ton) | 181 tonne (200 ton) |
| 2 | Feeder | Belt | 2 | 0 | 572 tonne/hr (630 tph) | 572 tonne/hr (630 tph) |
| 3 | Conveyor No. 1 | Belt | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 4 | Transfer Tower No. 1 | Enclosed | 1 | 0 | N/A | N/A |
| 5 | Conveyor No. 2 | Belt | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 6 | As-Received Coal Sampling System | Two-stage | 1 | 0 | N/A | N/A |
| 7 | Stacker/Reclaimer | Traveling, linear | 1 | 0 | 1,134 tonne/hr (1,250 tph) | 1,134 tonne/hr (1,250 tph) |
| 8 | Reclaim Hopper | N/A | 2 | 1 | 145 tonne (160 ton) | 145 tonne (160 ton) |
| 9 | Feeder | Vibratory | 2 | 1 | 590 tonne/hr (650 tph) | 590 tonne/hr (650 tph) |
| 10 | Conveyor No. 3 | Belt w/ tripper | 1 | 0 | 1,170 tonne/hr (1,290 tph) | 1,170 tonne/hr (1,290 tph) |
| 11 | Crusher Tower | N/A | 1 | 0 | N/A | N/A |
| 12 | Coal Surge Bin w/ Vent Filter | Dual outlet | 2 | 0 | 590 tonne (650 ton) | 590 tonne (650 ton) |
| 13 | Crusher | Impactor reduction | 2 | 0 | 8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0) | 8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 14 | As-Fired Coal Sampling System | Swing hammer | 1 | 1 | N/A | N/A |
| 15 | Conveyor No. 4 | Belt w/tripper | 1 | 0 | 1,170 tonne/hr (1,290 tph) | 1,170 tonne/hr (1,290 tph) |
| 16 | Transfer Tower No. 2 | Enclosed | 1 | 0 | N/A | N/A |
| 17 | Conveyor No. 5 | Belt w/ tripper | 1 | 0 | 1,170 tonne/hr (1,290 tph) | 1,170 tonne/hr (1,290 tph) |
| 18 | Coal Silo w/ Vent Filter and Slide Gates | Field erected | 3 | 0 | 2,631 tonne (2,900 ton) | 2,631 tonne (2,900 ton) |

### ACCOUNT 2 COAL PREPARATION AND FEED

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Feeder | Vibratory | 6 | 0 | 127 tonne/hr (140 tph) | 127 tonne/hr (140 tph) |
| 2 | Conveyor No. 6 | Belt w/tripper | 3 | 0 | 263 tonne/hr (290 tph) | 263 tonne/hr (290 tph) |
| 3 | Roller Mill Feed Hopper | Dual Outlet | 3 | 0 | 517 tonne (570 ton) | 517 tonne (570 ton) |
| 4 | Weigh Feeder | Belt | 6 | 0 | 127 tonne/hr (140 tph) | 127 tonne/hr (140 tph) |
| 5 | Pulverizer | Rotary | 6 | 0 | 127 tonne/hr (140 tph) | 127 tonne/hr (140 tph) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 6 | Coal Dryer Feed Hopper | Vertical Hopper | 3 | 0 | 517 tonne (570 ton) | 517 tonne (570 ton) |
| 7 | Coal Preheater | Water Heated Horizontal Rotary Kiln | 3 | 0 | Coal feed: 263 tonne/hr (290 tph)<br>Heat duty: 32.1 GJ/hr (30.5 MMBtu/hr) | Coal feed: 263 tonne/hr (290 tph)<br>Heat duty: 32.1 GJ/hr (30.5 MMBtu/hr) |
| 8 | Coal Dryer | Fluidized Bed with Internal Coils | 3 | 0 | Coal feed: 263 tonne/hr (290 tph)<br>Heat duty: 186 GJ/hr (177 MMBtu/hr)<br>Bed diameter: 16.5 m (54 ft) | Coal feed: 263 tonne/hr (290 tph)<br>Heat duty: 186 GJ/hr (177 MMBtu/hr)<br>Bed diameter: 16.5 m (54 ft) |
| 9 | Steam Compressor | Reciprocating, Multi-Stage | 3 | 0 | 1,419 $m^3$/min (50,110 scfm)<br>Suction - 0.10 MPa (13.8 psia)<br>Discharge - 0.52 MPa (75.3 psia) | 1,419 $m^3$/min (50,110 scfm)<br>Suction - 0.10 MPa (13.8 psia)<br>Discharge - 0.52 MPa (75.3 psia) |
| 10 | Dryer Exhaust Filter | Hot Baghouse | 3 | 0 | Steam - 71,214 kg/hr (157,000 lb/hr)<br>Temperature - 107°C (225°F) | Steam - 71,214 kg/hr (157,000 lb/hr)<br>Temperature - 107°C (225°F) |
| 11 | Dry Coal Cooler | Water Cooled Horizontal Rotary Kiln | 3 | 0 | 189 tonne/hr (208 tph)<br>Heat duty - 12 GJ/hr (11 MMBtu/hr) | 189 tonne/hr (208 tph)<br>Heat duty - 12 GJ/hr (11 MMBtu/hr) |

## ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Demineralized Water Storage Tank | Vertical, cylindrical, outdoor | 3 | 0 | 575,383 liters (152,000 gal) | 590,524 liters (156,000 gal) |
| 2 | Condensate Pumps | Vertical canned | 2 | 1 | 9,464 lpm @ 91 m $H_2O$ (2,500 gpm @ 300 ft $H_2O$) | 9,804 lpm @ 91 m $H_2O$ (2,590 gpm @ 300 ft $H_2O$) |
| 3 | Deaerator | Horizontal spray type | 2 | 0 | 732,098 kg/hr (1,614,000 lb/hr) | 751,603 kg/hr (1,657,000 lb/hr) |
| 4 | Intermediate Pressure Feedwater Pump | Horizontal centrifugal, single stage | 2 | 1 | 5,375 lpm @ 341 m $H_2O$ (1,420 gpm @ 1120 ft $H_2O$) | 5,716 lpm @ 341 m $H_2O$ (1,510 gpm @ 1120 ft $H_2O$) |
| 5 | High Pressure Feedwater Pump | Barrel type, multi-stage, centrifugal | 2 | 1 | HP water: 7,117 lpm @ 1,676 m $H_2O$ (1,880 gpm @ 5,500 ft $H_2O$) | HP water: 7,117 lpm @ 1,676 m $H_2O$ (1,880 gpm @ 5,500 ft $H_2O$) |
| 6 | Auxiliary Boiler | Shop fabricated, water tube | 1 | 0 | 18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F) | 18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F) |
| 7 | Service Air Compressors | Flooded Screw | 3 | 1 | 28 $m^3$/min @ 0.7 MPa (1,000 scfm @ 100 psig) | 28 $m^3$/min @ 0.7 MPa (1,000 scfm @ 100 psig) |
| 8 | Instrument Air Dryers | Duplex, regenerative | 3 | 1 | 28 $m^3$/min (1,000 scfm) | 28 $m^3$/min (1,000 scfm) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 9 | Closed Cycle Cooling Heat Exchangers | Plate and frame | 2 | 0 | 667 GJ/hr (632 MMBtu/hr) each | 840 GJ/hr (796 MMBtu/hr) each |
| 10 | Closed Cycle Cooling Water Pumps | Horizontal centrifugal | 2 | 1 | 239,238 lpm @ 21 m $H_2O$ (63,200 gpm @ 70 ft $H_2O$) | 301,319 lpm @ 21 m $H_2O$ (79,600 gpm @ 70 ft $H_2O$) |
| 11 | Engine-Driven Fire Pump | Vertical turbine, diesel engine | 1 | 1 | 3,785 lpm @ 107 m $H_2O$ (1,000 gpm @ 350 ft $H_2O$) | 3,785 lpm @ 107 m $H_2O$ (1,000 gpm @ 350 ft $H_2O$) |
| 12 | Fire Service Booster Pump | Two-stage horizontal centrifugal | 1 | 1 | 2,650 lpm @ 76 m $H_2O$ (700 gpm @ 250 ft $H_2O$) | 2,650 lpm @ 76 m $H_2O$ (700 gpm @ 250 ft $H_2O$) |
| 13 | Raw Water Pumps | Stainless steel, single suction | 2 | 1 | 4,656 lpm @ 18 m $H_2O$ (1,230 gpm @ 60 ft $H_2O$) | 5,413 lpm @ 18 m $H_2O$ (1,430 gpm @ 60 ft $H_2O$) |
| 14 | Ground Water Pumps | Stainless steel, single suction | 3 | 1 | 3,104 lpm @ 268 m $H_2O$ (820 gpm @ 880 ft $H_2O$) | 3,596 lpm @ 268 m $H_2O$ (950 gpm @ 880 ft $H_2O$) |
| 15 | Filtered Water Pumps | Stainless steel, single suction | 2 | 1 | 5,527 lpm @ 49 m $H_2O$ (1,460 gpm @ 160 ft $H_2O$) | 5,527 lpm @ 49 m $H_2O$ (1,460 gpm @ 160 ft $H_2O$) |
| 16 | Filtered Water Tank | Vertical, cylindrical | 2 | 0 | 2,649,788 liter (700,000 gal) | 2,653,574 liter (701,000 gal) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 17 | Makeup Water Demineralizer | Anion, cation, and mixed bed | 4 | 0 | 1,287 lpm (340 gpm) | 1,287 lpm (340 gpm) |
| 18 | Liquid Waste Treatment System | | 1 | 0 | 10 years, 24-hour storm | 10 years, 24-hour storm |

**ACCOUNT 4 GASIFIER, ASU, AND ACCESSORIES INCLUDING LOW TEMPERATURE HEAT RECOVERY**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Gasifier | Pressurized dry-feed, entrained bed | 6 | 0 | 3,084 tonne/day, 4.2 MPa (3,400 tpd, 615 psia) | 3,084 tonne/day, 4.2 MPa (3,400 tpd, 615 psia) |
| 2 | Synthesis Gas Cooler | Convective spiral-wound tube boiler | 6 | 0 | 206,385 kg/hr (455,000 lb/hr) | 206,385 kg/hr (455,000 lb/hr) |
| 3 | Syngas Scrubber Including Sour Water Stripper | Vertical upflow | 6 | 0 | 206,385 kg/hr (455,000 lb/hr) | 206,385 kg/hr (455,000 lb/hr) |
| 4 | Raw Gas Coolers | Shell and tube with condensate drain | 12 | 0 | 424,109 kg/hr (935,000 lb/hr) | 424,109 kg/hr (935,000 lb/hr) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 5 | Raw Gas Knockout Drum | Vertical with mist eliminator | 3 | 0 | 394,172 kg/hr, 35°C, 3.4 MPa (869,000 lb/hr, 95°F, 490 psia) | 394,172 kg/hr, 35°C, 3.4 MPa (869,000 lb/hr, 95°F, 495 psia) |
| 6 | Saturation Water Economizers | Shell and tube | 3 | 0 | 494,869 kg/hr (1,091,000 lb/hr) | 494,416 kg/hr (1,090,000 lb/hr) |
| 7 | Flare Stack | Self-supporting, carbon steel, stainless steel top, pilot ignition | 3 | 0 | 412,769 kg/hr (910,000 lb/hr) syngas | 412,769 kg/hr (910,000 lb/hr) syngas |
| 8 | ASU Main Air Compressor | Centrifugal, multi-stage | 3 | 0 | 7,646 $m^3$/min @ 1.3 MPa (270,000 scfm @ 190 psia) | 7,646 $m^3$/min @ 1.3 MPa (270,000 scfm @ 190 psia) |
| 9 | Cold Box | Vendor design | 3 | 0 | 3,084 tonne/day (3,400 tpd) of 99% purity oxygen | 3,084 tonne/day (3,400 tpd) of 99% purity oxygen |
| 10 | Oxygen Compressor | Centrifugal, multi-stage | 3 | 0 | 1,557 $m^3$/min (55,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia) | 1,557 $m^3$/min (55,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 11 | Coal Feed System $CO_2$ Compressor | Centrifugal, multi-stage | 3 | 0 | 736 $m^3$/min (26,000 scfm)<br>Suction - 1.0 MPa (150 psia)<br>Discharge - 5.7 MPa (820 psia) | 736 $m^3$/min (26,000 scfm)<br>Suction - 5.2 MPa (750 psia)<br>Discharge - 5.7 MPa (820 psia) |

**ACCOUNT 5A SYNGAS CLEANUP**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Mercury Adsorber | Sulfated carbon bed | 3 | 0 | 394,172 kg/hr (869,000 lb/hr)<br>35°C (95°F)<br>3.3 MPa (485 psia) | 394,172 kg/hr (869,000 lb/hr)<br>35°C (95°F)<br>3.4 MPa (490 psia) |
| 2 | ZnO Guard Bed | Fixed bed | 1 | 0 | 418 $Nm^3$/min (14,750 acfm) | 417 $Nm^3$/min (14,740 acfm) |
| 3 | Sulfur Plant | Claus type | 2 | 0 | 58 tonne/day (64 tpd) | 58 tonne/day (64 tpd) |
| 4 | Water Gas Shift Reactor | Fixed bed, catalytic | 3 | 0 | 448,149 kg/hr (988,000 lb/hr)<br>232°C (450°F)<br>3.7 MPa (540 psia) | 448,149 kg/hr (988,000 lb/hr)<br>232°C (450°F)<br>3.7 MPa (540 psia) |
| 5 | Shift Reactor Heat Recovery Exchanger | Shell and Tube | 3 each | 0 | Exchanger 1: 61 GJ/hr (58 MMBtu/hr)<br>Exchanger 2: 77 GJ/hr (73 MMBtu/hr)<br>Exchanger 3: 17 GJ/hr (16 MMBtu/hr) | Exchanger 1: 61 GJ/hr (58 MMBtu/hr)<br>Exchanger 2: 77 GJ/hr (73 MMBtu/hr)<br>Exchanger 3: 17 GJ/hr (16 MMBtu/hr) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 6 | Acid Gas Removal Plant | Two-stage Selexol | 3 | 0 | 401,883 kg/hr (886,000 lb/hr) 34°C (94°F) 3.3 MPa (475 psia) | 401,883 kg/hr (886,000 lb/hr) 34°C (94°F) 3.3 MPa (480 psia) |
| 7 | Hydrogenation Reactor | Fixed bed, catalytic | 2 | 0 | 13,982 kg/hr (30,826 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia) | 13,842 kg/hr (30,517 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia) |
| 8 | Tail Gas Recycle Compressor | Centrifugal | 2 | 0 | 11,601 kg/hr @ 3.3 MPa (25,575 lb/hr @ 485 psia) | 11,542 kg/hr @ 3.3 MPa (25,445 lb/hr @ 485 psia) |

**ACCOUNT 5B $CO_2$ COMPRESSION**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | $CO_2$ Compressor | Integrally geared, multi-stage centrifugal | 4 | 0 | N/A | 1,705 $m^3$/min @ 15.3 MPa (60,200 scfm @ 2,215 psia) |

## ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Combustion Turbine | N/A | N/A | N/A | N/A | N/A |
| 2 | Combustion Turbine Generator | N/A | N/A | N/A | N/A | N/A |

## ACCOUNT 7 DUCTING & STACK

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Stack | CS plate, type 409SS liner | 1 | 0 | 76 m (250 ft) high x 2.2 m (7 ft) diameter | 76 m (250 ft) high x 0.7 m (2 ft) diameter |

## ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Steam Turbine | Commercially available | 1 | 0 | 322 MW 12.4 MPa/ 566°C (1800 psig/ 1050°F) | 327 MW 12.4 MPa/ 566°C (1800 psig/ 1050°F) |
| 2 | Steam Turbine Generator | Hydrogen cooled, static excitation | 1 | 0 | 360 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase | 360 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 3 | Surface Condenser | Single pass, divided waterbox including vacuum pumps | 1 | 0 | 897 GJ/hr (850 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 8°C (47°F), Water temperature rise 11°C (20°F) | 939 GJ/hr (890 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F) |
| 4 | Air-cooled Condenser | --- | 1 | 0 | 897 GJ/hr (850 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 4°C (40°F) | 939 GJ/hr (890 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F) |

**ACCOUNT 9 COOLING WATER SYSTEM**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Circulating Water Pumps | Vertical, wet pit | 2 | 1 | 458,035 lpm @ 30 m (121,000 gpm @ 100 ft) | 526,172 lpm @ 30 m (139,000 gpm @ 100 ft) |
| 2 | Cooling Tower | Evaporative, mechanical draft, multi-cell | 1 | 0 | 2°C (36°F) wet bulb / 8°C (47°F) CWT / 19°C (67°F) HWT / 2,543 GJ/hr (2,410 MMBtu/hr) heat duty | 3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 2,933 GJ/hr (2,780 MMBtu/hr) heat duty |

**ACCOUNT 10 SLAG RECOVERY AND HANDLING**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Slag Quench Tank | Water bath | 6 | 0 | 249,837 liters (66,000 gal) | 249,837 liters (66,000 gal) |
| 2 | Slag Crusher | Roll | 6 | 0 | 13 tonne/hr (14 tph) | 13 tonne/hr (14 tph) |
| 3 | Slag Depressurizer | Proprietary | 6 | 0 | 13 tonne/hr (14 tph) | 13 tonne/hr (14 tph) |
| 4 | Slag Receiving Tank | Horizontal, weir | 6 | 0 | 151,416 liters (40,000 gal) | 151,416 liters (40,000 gal) |
| 5 | Black Water Overflow Tank | Shop fabricated | 6 | 0 | 68,137 liters (18,000 gal) | 68,137 liters (18,000 gal) |
| 6 | Slag Conveyor | Drag chain | 6 | 0 | 13 tonne/hr (14 tph) | 13 tonne/hr (14 tph) |
| 7 | Slag Separation Screen | Vibrating | 6 | 0 | 13 tonne/hr (14 tph) | 13 tonne/hr (14 tph) |
| 8 | Coarse Slag Conveyor | Belt/bucket | 6 | 0 | 13 tonne/hr (14 tph) | 13 tonne/hr (14 tph) |
| 9 | Fine Ash Settling Tank | Vertical, gravity | 6 | 0 | 211,983 liters (56,000 gal) | 211,983 liters (56,000 gal) |
| 10 | Fine Ash Recycle Pumps | Horizontal centrifugal | 6 | 3 | 38 lpm @ 14 m $H_2O$ (10 gpm @ 46 ft $H_2O$) | 38 lpm @ 14 m $H_2O$ (10 gpm @ 46 ft $H_2O$) |
| 11 | Grey Water Storage Tank | Field erected | 6 | 0 | 68,137 liters (18,000 gal) | 68,137 liters (18,000 gal) |
| 12 | Grey Water Pumps | Centrifugal | 6 | 3 | 227 lpm @ 433 m $H_2O$ (60 gpm @ 1,420 ft $H_2O$) | 227 lpm @ 433 m $H_2O$ (60 gpm @ 1,420 ft $H_2O$) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 13 | Slag Storage Bin | Vertical, field erected | 6 | 0 | 907 tonne (1,000 tons) | 907 tonne (1,000 tons) |
| 14 | Unloading Equipment | Telescoping chute | 2 | 0 | 163 tonne/hr (180 tph) | 163 tonne/hr (180 tph) |

**ACCOUNT 11 SNG PRODUCTION/METHANATION SYSTEM**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | Methanation Reactor Preheater | Shell and Tube | 3 | 0 | 90 GJ/hr (85 MMBtu/hr) | 90 GJ/hr (85 MMBtu/hr) |
| 2 | Methanation Reactor 1 | Fixed Bed, catalytic | 3 | 0 | 456,314 kg/hr (1,006,000 lb/hr) 693°C (1,280°F) 3.2 MPa (470 psia) | 448,603 kg/hr (989,000 lb/hr) 699°C (1,290°F) 3.2 MPa (470 psia) |
| 3 | Methanation Reactor Intercooler 1 | Shell and Tube | 3 | 0 | 381 GJ/hr (362 MMBtu/hr) | 380 GJ/hr (360 MMBtu/hr) |
| 4 | Methanation Reactor Intercooler 2 | Shell and Tube | 3 | 0 | 104 GJ/hr (99 MMBtu/hr) | 102 GJ/hr (97 MMBtu/hr) |
| 5 | Methanation Reactor 2 | Fixed Bed, catalytic | 3 | 0 | 119,295 kg/hr (263,000 lb/hr) 554°C (1030°F) 3.1 MPa (450 psia) | 119,295 kg/hr (263,000 lb/hr) 560°C (1040°F) 3.1 MPa (450 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 6 | Methanation Reactor Intercooler 3 | Shell and Tube | 3 | 0 | 27 GJ/hr (26 MMBtu/hr) | 27 GJ/hr (26 MMBtu/hr) |
| 7 | Methanation Reactor 3 | Fixed Bed, catalytic | 3 | 0 | 119,295 kg/hr (263,000 lb/hr) 338°C (640°F) 3.0 MPa (440 psia) | 119,295 kg/hr (263,000 lb/hr) 343°C (650°F) 3.0 MPa (440 psia) |
| 8 | Methanation Reactor Intercooler 4 | Shell and Tube | 3 | 0 | 36 GJ/hr (34 MMBtu/hr) | 38 GJ/hr (36 MMBtu/hr) |
| 9 | Methanation Reactor Intercooler 5 | Shell and Tube | 3 | 0 | 61 GJ/hr (58 MMBtu/hr) | 61 GJ/hr (58 MMBtu/hr) |
| 10 | SNG Purification Condenser 1 | Shell and Tube | 3 | 0 | 51 GJ/hr (49 MMBtu/hr) | 51 GJ/hr (49 MMBtu/hr) |
| 11 | SNG Purification Condenser 2 | Shell and Tube | 3 | 0 | 31 GJ/hr (30 MMBtu/hr) | 31 GJ/hr (30 MMBtu/hr) |
| 12 | Methanation Recycle Compressor | Centrifugal | 3 | 1 | 8,084 $m^3$/min (285,500 scfm) Suction - 3.1 MPa (453 psia) Discharge - 3.3 MPa (475 psia) | 7,926 $m^3$/min (279,900 scfm) Suction - 3.1 MPa (453 psia) Discharge - 3.3 MPa (475 psia) |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 13 | Molecular Sieve Purification Reactor | Fixed bed | 6 | 0 | 34,927 kg/hr (77,000 lb/hr)<br>35°C (95°F)<br>2.8 MPa (401 psia) | 34,927 kg/hr (77,000 lb/hr)<br>35°C (95°F)<br>2.8 MPa (401 psia) |
| 14 | SNG Product Compressor | Centrifugal, Multi-staged | 3 | 1 | 1,136 $m^3$/min<br>(40,100 scfm)<br>2.6 MPa (381 psia)<br>6.2 MPa (900 psia) | 1,136 $m^3$/min<br>(40,100 scfm)<br>2.6 MPa (381 psia)<br>6.2 MPa (900 psia) |

**ACCOUNT 12 ACCESSORY ELECTRIC PLANT**

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | STG Step-up Transformer | Oil-filled | 1 | 0 | 24 kV/345 kV, 360 MVA, 3-ph, 60 Hz | 345 kV/24 kV, 360 MVA, 3-ph, 60 Hz |
| 2 | Auxiliary Transformer | Oil-filled | 1 | 1 | 24 kV/4.16 kV, 287 MVA, 3-ph, 60 Hz | 24 kV/4.16 kV, 347 MVA, 3-ph, 60 Hz |
| 3 | Low Voltage Transformer | Dry ventilated | 1 | 1 | 4.16 kV/480 V, 43 MVA, 3-ph, 60 Hz | 4.16 kV/480 V, 52 MVA, 3-ph, 60 Hz |
| 4 | STG Isolated Phase Bus Duct and Tap Bus | Aluminum, self-cooled | 1 | 0 | 24 kV, 3-ph, 60 Hz | 24 kV, 3-ph, 60 Hz |
| 5 | Medium Voltage Switchgear | Metal clad | 1 | 1 | 4.16 kV, 3-ph, 60 Hz | 4.16 kV, 3-ph, 60 Hz |

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 6 | Low Voltage Switchgear | Metal enclosed | 1 | 1 | 480 V, 3-ph, 60 Hz | 480 V, 3-ph, 60 Hz |
| 7 | Emergency Diesel Generator | Sized for emergency shutdown | 1 | 0 | 750 kW, 480 V, 3-ph, 60 Hz | 750 kW, 480 V, 3-ph, 60 Hz |

## ACCOUNT 13 INSTRUMENTATION AND CONTROLS

| Equipment No. | Description | Type | Operating Qty. | Spares | Case 7 Design Condition | Case 8 Design Condition |
|---|---|---|---|---|---|---|
| 1 | DCS - Main Control | Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W) | 1 | 0 | Operator stations/printers and engineering stations/printers | Operator stations/printers and engineering stations/printers |
| 2 | DCS - Processor | Microprocessor with redundant input/output | 1 | 0 | N/A | N/A |
| 3 | DCS - Data Highway | Fiber optic | 1 | 0 | Fully redundant, 25% spare | Fully redundant, 25% spare |

### 5.2.5 Cases 7 and 8 Cost Estimating

The cost estimating methodology was described previously in Section 2.8.

The TPC organized by cost account; owner's costs; TOC; and initial and annual O&M costs for the SNG plant without sequestration using NDL coal (Case 7) are shown in Exhibit 5-39 and Exhibit 5-40, respectively. The same data for the SNG plant with sequestration using NDL coal (Case 8) are shown in Exhibit 5-41 and Exhibit 5-42.

The estimated TOC of the SNG plant without carbon sequestration using NDL coal is 3.50 billion dollars and with carbon sequestration is 3.60 billion dollars. Project and process contingencies represent 12.0 and 3.6 percent for both cases. The FYCOP is $21.27/MMBtu for the non-sequestration case and $23.24/MMBtu for the sequestration case as shown in Exhibit ES-6.

# Exhibit 5-39 Case 7 Total Plant Cost Details

| | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | **Client:** | USDOE/NETL | | | | | | | | **Report Date:** | 2010-Feb-18 |
| | **Project:** | Cost & Performance Baseline for Fossil Energy Plants Vol.2 | | | | | | | | | |
| | | **TOTAL PLANT COST SUMMARY** | | | | | | | | | |
| | **Case:** | Case 7 - Siemens Quench SNG Production w/o CO2 | | | | | | | | | |
| | **Plant Size:** | 46.06 | MW,net | **Estimate Type:** | | Conceptual | | | **Cost Base (Jun)** | 2007 | ($x1000) |

| Acct | | Equipment | Material | Labor | | Sales | Bare Erected | Eng'g CM | Contingencies | | Total Plant Cost |
|---|---|---|---|---|---|---|---|---|---|---|---|
| No. | Item/Description | Cost | Cost | Direct | Indirect | Tax | Cost $ | H.O.& Fee | Process | Project | $ |
| 1 | COAL & SORBENT HANDLING | | | | | | | | | | |
| 1.1 | Coal Receive & Unload | $7,733 | $0 | $3,779 | $0 | $0 | $11,512 | $1,031 | $0 | $2,509 | $15,052 |
| 1.2 | Coal Stackout & Reclaim | $9,993 | $0 | $2,423 | $0 | $0 | $12,415 | $1,088 | $0 | $2,701 | $16,204 |
| 1.3 | Coal Conveyors & Yd Crush | $9,291 | $0 | $2,397 | $0 | $0 | $11,688 | $1,026 | $0 | $2,543 | $15,256 |
| 1.4 | Other Coal Handling | $2,431 | $0 | $555 | $0 | $0 | $2,985 | $261 | $0 | $649 | $3,896 |
| 1.5 | Sorbent Receive & Unload | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.6 | Sorbent Stackout & Reclaim | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.7 | Sorbent Conveyors | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.8 | Other Sorbent Handling | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.9 | Coal & Sorbent Hnd.Foundations | $0 | $5,472 | $13,683 | $0 | $0 | $19,155 | $1,836 | $0 | $4,198 | $25,189 |
| | **SUBTOTAL 1.** | **$29,447** | **$5,472** | **$22,836** | **$0** | **$0** | **$57,755** | **$5,243** | **$0** | **$12,600** | **$75,598** |
| 2 | COAL & SORBENT PREP & FEED | | | | | | | | | | |
| 2.1 | Coal Crushing & Drying | $127,899 | $0 | $18,636 | $0 | $0 | $146,535 | $13,308 | $0 | $31,969 | $191,811 |
| 2.2 | Prepared Coal Storage & Feed | $6,058 | $1,450 | $950 | $0 | $0 | $8,457 | $723 | $0 | $1,836 | $11,017 |
| 2.3 | Dry Coal Injection System | $199,367 | $2,314 | $18,515 | $0 | $0 | $220,197 | $18,965 | $0 | $47,832 | $286,994 |
| 2.4 | Misc.Coal Prep & Feed | $3,331 | $2,424 | $7,268 | $0 | $0 | $13,024 | $1,197 | $0 | $2,844 | $17,065 |
| 2.5 | Sorbent Prep Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.6 | Sorbent Storage & Feed | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.7 | Sorbent Injection System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.8 | Booster Air Supply System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.9 | Coal & Sorbent Feed Foundation | $0 | $12,948 | $10,631 | $0 | $0 | $23,579 | $2,184 | $0 | $5,153 | $30,916 |
| | **SUBTOTAL 2.** | **$336,655** | **$19,136** | **$56,001** | **$0** | **$0** | **$411,792** | **$36,377** | **$0** | **$89,634** | **$537,802** |
| 3 | FEEDWATER & MISC. BOP SYSTEMS | | | | | | | | | | |
| 3.1 | Feedwater System | $3,463 | $1,610 | $1,952 | $0 | $0 | $7,025 | $635 | $0 | $1,532 | $9,192 |
| 3.2 | Water Makeup & Pretreating | $591 | $62 | $330 | $0 | $0 | $982 | $94 | $0 | $323 | $1,399 |
| 3.3 | Other Feedwater Subsystems | $3,760 | $1,137 | $1,860 | $0 | $0 | $6,757 | $613 | $0 | $1,474 | $8,845 |
| 3.4 | Service Water Systems | $338 | $696 | $2,415 | $0 | $0 | $3,449 | $337 | $0 | $1,136 | $4,921 |
| 3.5 | Other Boiler Plant Systems | $1,814 | $703 | $1,742 | $0 | $0 | $4,258 | $404 | $0 | $932 | $5,594 |
| 3.6 | FO Supply Sys & Nat Gas | $102 | $210 | $188 | $0 | $0 | $500 | $48 | $0 | $110 | $658 |
| 3.7 | Waste Treatment Equipment | $826 | $0 | $504 | $0 | $0 | $1,329 | $129 | $0 | $438 | $1,896 |
| 3.8 | Misc. Power Plant Equipment | $1,639 | $219 | $841 | $0 | $0 | $2,700 | $261 | $0 | $888 | $3,848 |
| | **SUBTOTAL 3.** | **$12,532** | **$4,636** | **$9,832** | **$0** | **$0** | **$27,001** | **$2,521** | **$0** | **$6,832** | **$36,353** |
| 4 | GASIFIER & ACCESSORIES | | | | | | | | | | |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries | $325,384 | $0 | $150,649 | $0 | $0 | $476,033 | $42,281 | $71,405 | $88,458 | $678,176 |
| 4.2 | Syngas Cooling | w/4.1 | w/4.1 | w/ 4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.3 | ASU/Oxidant Compression | $246,271 | $0 | w/equip. | $0 | $0 | $246,271 | $23,871 | $0 | $27,014 | $297,156 |
| 4.4 | LT Heat Recovery & FG Saturation | $51,285 | $0 | $19,002 | $0 | $0 | $70,288 | $6,733 | $0 | $15,404 | $92,425 |
| 4.5 | Misc. Gasification Equipment | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.6 | Flare Stack System | $0 | $2,261 | $920 | $0 | $0 | $3,182 | $305 | $0 | $697 | $4,184 |
| 4.7 | CO2 Solid Feed System Compressors | $13,665 | $3,280 | $4,919 | $0 | $0 | $21,863 | $2,186 | $0 | $4,810 | $28,860 |
| 4.8 | Major Component Rigging | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.9 | Gasification Foundations | $0 | $26,928 | $15,365 | $0 | $0 | $42,293 | $3,872 | $0 | $11,541 | $57,706 |
| | **SUBTOTAL 4.** | **$636,605** | **$32,469** | **$190,856** | **$0** | **$0** | **$859,930** | **$79,248** | **$71,405** | **$147,925** | **$1,158,507** |

## Exhibit 5-39 Case 7 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 5A | GAS CLEANUP & PIPING | | | | | | | | | | |
| 5A.1 | Double Stage Selexol | $130,462 | $0 | $110,700 | $0 | $0 | $241,162 | $23,323 | $48,232 | $62,543 | $375,260 |
| 5A.2 | Elemental Sulfur Plant | $8,844 | $1,763 | $11,410 | $0 | $0 | $22,017 | $2,139 | $0 | $4,831 | $28,986 |
| 5A.3 | Mercury Removal | $2,366 | $0 | $1,800 | $0 | $0 | $4,166 | $402 | $208 | $955 | $5,732 |
| 5A.4 | Shift Reactors | $10,266 | $0 | $4,132 | $0 | $0 | $14,398 | $1,380 | $0 | $3,156 | $18,934 |
| 5A.5 | Methanation | $26,500 | $10,600 | $15,900 | $0 | $0 | $53,000 | $5,300 | $5,300 | $12,720 | $76,320 |
| 5A.6 | SNG Purification & Compression | $18,600 | $7,440 | $11,160 | $0 | $0 | $37,200 | $3,720 | $0 | $8,184 | $49,104 |
| 5A.7 | Fuel Gas Piping | $0 | $1,631 | $1,142 | $0 | $0 | $2,773 | $257 | $0 | $606 | $3,636 |
| 5A.9 | Process Interconnects | $0 | $12,000 | $18,000 | $0 | $0 | $30,000 | $3,000 | $0 | $6,600 | $39,600 |
| 5A.10 | HGCU Foundations | $0 | $1,651 | $1,065 | $0 | $0 | $2,716 | $250 | $0 | $890 | $3,856 |
| 5A.11 | Zinc Oxide Guard Bed | $758 | $0 | $140 | $0 | $0 | $898 | $90 | $0 | $198 | $1,185 |
| | **SUBTOTAL 5A.** | **$197,795** | **$35,085** | **$175,449** | **$0** | **$0** | **$408,330** | **$39,861** | **$53,741** | **$100,683** | **$602,614** |
| 5B | CO2 COMPRESSION | | | | | | | | | | |
| 5B.1 | CO2 Removal System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 5B.2 | CO2 Compression & Drying | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 5B.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 6 | COMBUSTION TURBINE/ACCESSORIES | | | | | | | | | | |
| 6.1 | Combustion Turbine Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.2 | Syngas Expander | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.3 | Compressed Air Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.9 | Combustion Turbine Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 6.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 7 | HRSG, DUCTING & STACK | | | | | | | | | | |
| 7.1 | Heat Recovery Steam Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.2 | HRSG Accessories | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $335 | $239 | $0 | $0 | $573 | $50 | $0 | $125 | $748 |
| 7.4 | Stack | $1,465 | $0 | $550 | $0 | $0 | $2,015 | $193 | $0 | $221 | $2,429 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $293 | $282 | $0 | $0 | $575 | $54 | $0 | $189 | $818 |
| | **SUBTOTAL 7.** | **$1,465** | **$628** | **$1,071** | **$0** | **$0** | **$3,164** | **$297** | **$0** | **$534** | **$3,995** |
| 8 | STEAM TURBINE GENERATOR | | | | | | | | | | |
| 8.1 | Steam TG & Accessories | $31,248 | $0 | $5,416 | $0 | $0 | $36,664 | $3,518 | $0 | $4,018 | $44,200 |
| 8.2 | Turbine Plant Auxiliaries | $169 | $0 | $498 | $0 | $0 | $668 | $66 | $0 | $73 | $807 |
| 8.3a | Condenser & Auxiliaries | $3,246 | $0 | $1,521 | $0 | $0 | $4,767 | $458 | $0 | $522 | $5,747 |
| 8.3b | Air Cooled Condenser | $29,744 | $0 | $5,963 | $0 | $0 | $35,707 | $3,571 | $0 | $7,855 | $47,133 |
| 8.4 | Steam Piping | $5,463 | $0 | $3,843 | $0 | $0 | $9,306 | $799 | $0 | $2,526 | $12,632 |
| 8.9 | TG Foundations | $0 | $1,078 | $1,823 | $0 | $0 | $2,901 | $275 | $0 | $953 | $4,129 |
| | **SUBTOTAL 8.** | **$69,869** | **$1,078** | **$19,065** | **$0** | **$0** | **$90,013** | **$8,687** | **$0** | **$15,949** | **$114,648** |

## Exhibit 5-39 Case 7 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 9 | COOLING WATER SYSTEM | | | | | | | | | | |
| 9.1 | Cooling Towers | $7,073 | $0 | $1,287 | $0 | $0 | $8,360 | $796 | $0 | $1,373 | $10,530 |
| 9.2 | Circulating Water Pumps | $1,885 | $0 | $87 | $0 | $0 | $1,973 | $166 | $0 | $321 | $2,460 |
| 9.3 | Circ.Water System Auxiliaries | $155 | $0 | $22 | $0 | $0 | $177 | $17 | $0 | $29 | $222 |
| 9.4 | Circ.Water Piping | $0 | $6,449 | $1,672 | $0 | $0 | $8,121 | $734 | $0 | $1,771 | $10,626 |
| 9.5 | Make-up Water System | $329 | $0 | $471 | $0 | $0 | $800 | $77 | $0 | $175 | $1,052 |
| 9.6 | Component Cooling Water Sys | $761 | $911 | $648 | $0 | $0 | $2,320 | $217 | $0 | $507 | $3,044 |
| 9.9 | Circ.Water System Foundations | $0 | $2,356 | $4,005 | $0 | $0 | $6,360 | $603 | $0 | $2,089 | $9,052 |
| | **SUBTOTAL 9.** | **$10,204** | **$9,715** | **$8,191** | **$0** | **$0** | **$28,111** | **$2,610** | **$0** | **$6,266** | **$36,987** |
| 10 | ASH/SPENT SORBENT HANDLING SYS | | | | | | | | | | |
| 10.1 | Slag Dewatering & Cooling | $49,571 | $0 | $48,892 | $0 | $0 | $98,463 | $9,540 | $0 | $10,800 | $118,804 |
| 10.2 | Gasifier Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.3 | Cleanup Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.4 | High Temperature Ash Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.5 | Other Ash Rrecovery Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.6 | Ash Storage Silos | $1,072 | $0 | $1,167 | $0 | $0 | $2,239 | $217 | $0 | $368 | $2,825 |
| 10.7 | Ash Transport & Feed Equipment | $1,438 | $0 | $347 | $0 | $0 | $1,785 | $167 | $0 | $293 | $2,245 |
| 10.8 | Misc. Ash Handling Equipment | $2,221 | $2,722 | $813 | $0 | $0 | $5,756 | $548 | $0 | $946 | $7,250 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $95 | $119 | $0 | $0 | $214 | $20 | $0 | $70 | $304 |
| | **SUBTOTAL 10.** | **$54,303** | **$2,817** | **$51,338** | **$0** | **$0** | **$108,458** | **$10,492** | **$0** | **$12,477** | **$131,427** |
| 11 | ACCESSORY ELECTRIC PLANT | | | | | | | | | | |
| 11.1 | Generator Equipment | $431 | $0 | $427 | $0 | $0 | $858 | $82 | $0 | $94 | $1,034 |
| 11.2 | Station Service Equipment | $5,673 | $0 | $511 | $0 | $0 | $6,184 | $570 | $0 | $675 | $7,429 |
| 11.3 | Switchgear & Motor Control | $10,487 | $0 | $1,907 | $0 | $0 | $12,394 | $1,150 | $0 | $2,032 | $15,576 |
| 11.4 | Conduit & Cable Tray | $0 | $4,872 | $16,071 | $0 | $0 | $20,943 | $2,026 | $0 | $5,742 | $28,711 |
| 11.5 | Wire & Cable | $0 | $9,308 | $6,116 | $0 | $0 | $15,424 | $1,120 | $0 | $4,136 | $20,681 |
| 11.6 | Protective Equipment | $0 | $826 | $3,006 | $0 | $0 | $3,832 | $374 | $0 | $631 | $4,837 |
| 11.7 | Standby Equipment | $317 | $0 | $309 | $0 | $0 | $626 | $60 | $0 | $103 | $789 |
| 11.8 | Main Power Transformers | $9,424 | $0 | $172 | $0 | $0 | $9,596 | $728 | $0 | $1,549 | $11,872 |
| 11.9 | Electrical Foundations | $0 | $186 | $489 | $0 | $0 | $675 | $65 | $0 | $222 | $962 |
| | **SUBTOTAL 11.** | **$26,332** | **$15,192** | **$29,008** | **$0** | **$0** | **$70,533** | **$6,174** | **$0** | **$15,184** | **$91,890** |
| 12 | INSTRUMENTATION & CONTROL | | | | | | | | | | |
| 12.1 | IGCC Control Equipment | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.2 | Combustion Turbine Control | w/6.1 | $0 | w/6.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.3 | Steam Turbine Control | w/8.1 | $0 | w/8.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.4 | Other Major Component Control | $520 | $0 | $347 | $0 | $0 | $867 | $82 | $43 | $149 | $1,141 |
| 12.5 | Signal Processing Equipment | w/12.7 | $0 | w/12.7 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.6 | Control Boards,Panels & Racks | $345 | $0 | $221 | $0 | $0 | $566 | $54 | $28 | $130 | $778 |
| 12.7 | Computer & Accessories | $5,524 | $0 | $177 | $0 | $0 | $5,701 | $523 | $285 | $651 | $7,160 |
| 12.8 | Instrument Wiring & Tubing | $0 | $2,798 | $5,720 | $0 | $0 | $8,518 | $722 | $426 | $2,417 | $12,083 |
| 12.9 | Other I & C Equipment | $4,118 | $0 | $2,000 | $0 | $0 | $6,118 | $576 | $306 | $1,050 | $8,050 |
| | **SUBTOTAL 12.** | **$10,507** | **$2,798** | **$8,465** | **$0** | **$0** | **$21,770** | **$1,957** | **$1,088** | **$4,396** | **$29,211** |

## Exhibit 5-39 Case 7 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 13 | IMPROVEMENTS TO SITE | | | | | | | | | | |
| 13.1 | Site Preparation | $0 | $141 | $3,021 | $0 | $0 | $3,162 | $314 | $0 | $1,043 | $4,519 |
| 13.2 | Site Improvements | $0 | $2,514 | $3,341 | $0 | $0 | $5,855 | $578 | $0 | $1,930 | $8,362 |
| 13.3 | Site Facilities | $4,505 | $0 | $4,754 | $0 | $0 | $9,259 | $913 | $0 | $3,051 | $13,223 |
| | **SUBTOTAL 13.** | **$4,505** | **$2,655** | **$11,115** | **$0** | **$0** | **$18,276** | **$1,804** | **$0** | **$6,024** | **$26,104** |
| 14 | BUILDINGS & STRUCTURES | | | | | | | | | | |
| 14.1 | Combustion Turbine Area | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 14.2 | Steam Turbine Building | $0 | $2,599 | $3,703 | $0 | $0 | $6,302 | $580 | $0 | $1,032 | $7,914 |
| 14.3 | Administration Building | $0 | $870 | $630 | $0 | $0 | $1,500 | $134 | $0 | $245 | $1,879 |
| 14.4 | Circulation Water Pumphouse | $0 | $160 | $90 | $0 | $0 | $250 | $22 | $0 | $41 | $312 |
| 14.5 | Water Treatment Buildings | $0 | $494 | $482 | $0 | $0 | $975 | $88 | $0 | $160 | $1,223 |
| 14.6 | Machine Shop | $0 | $445 | $305 | $0 | $0 | $750 | $67 | $0 | $123 | $939 |
| 14.7 | Warehouse | $0 | $726 | $464 | $0 | $0 | $1,190 | $105 | $0 | $194 | $1,489 |
| 14.8 | Other Buildings & Structures | $0 | $514 | $400 | $0 | $0 | $914 | $82 | $0 | $199 | $1,195 |
| 14.9 | Waste Treating Building & Str. | $0 | $1,149 | $2,196 | $0 | $0 | $3,346 | $312 | $0 | $732 | $4,389 |
| | **SUBTOTAL 14.** | **$0** | **$6,957** | **$8,270** | **$0** | **$0** | **$15,227** | **$1,389** | **$0** | **$2,725** | **$19,341** |
| | | | | | | | | | | | |
| | **TOTAL COST** | **$1,390,220** | **$138,640** | **$591,498** | **$0** | **$0** | **$2,120,358** | **$196,659** | **$126,234** | **$421,228** | **$2,864,478** |
| | | | | | | | | | | | |
| | **Owner's Costs** | | | | | | | | | | |
| | **Preproduction Costs** | | | | | | | | | | |
| | 6 Months All Labor | | | | | | | | | | $22,173 |
| | 1 Month Maintenance Materials | | | | | | | | | | $3,941 |
| | 1 Month Non-fuel Consumables | | | | | | | | | | $615 |
| | 1 Month Waste Disposal | | | | | | | | | | $937 |
| | 25% of 1 Months Fuel Cost at 100% CF | | | | | | | | | | $1,559 |
| | 2% of TPC | | | | | | | | | | $57,290 |
| | **Total** | | | | | | | | | | **$86,515** |
| | **Inventory Capital** | | | | | | | | | | |
| | 60 day supply of fuel and consumables at 100% CF | | | | | | | | | | $13,495 |
| | 0.5% of TPC (spare parts) | | | | | | | | | | $14,322 |
| | **Total** | | | | | | | | | | **$27,817** |
| | | | | | | | | | | | |
| | **Initial Cost for Catalyst and Chemicals** | | | | | | | | | | $17,540 |
| | **Land** | | | | | | | | | | $900 |
| | **Other Owner's Costs** | | | | | | | | | | $429,672 |
| | **Financing Costs** | | | | | | | | | | $77,341 |
| | **Total Overnight Costs (TOC)** | | | | | | | | | | **$3,504,262** |
| | TASC Multiplier | | | | | | | | | | 1.201 |
| | **Total As-Spent Cost (TASC)** | | | | | | | | | | **$4,208,619** |

# Exhibit 5-40 Case 7 Initial and Annual O&M Costs

| | | | | | | |
|---|---|---|---|---|---|---|
| INITIAL & ANNUAL O&M EXPENSES | | | | | Cost Base (Jun): | 2007 |
| Case 7 - Siemens Quench SNG Production w/o CO2 | | | | | MWe-net: | 46 |
| SNG (MMbtu/hr): 6353 | | | | | Capacity Factor (%): | 90 |
| OPERATING & MAINTENANCE LABOR | | | | | | |
| Operating Labor | | | | | | |
| Operating Labor Rate(base): | 34.65 | $/hour | | | | |
| Operating Labor Burden: | 30.00 | % of base | | | | |
| Labor O-H Charge Rate: | 25.00 | % of labor | | | | |
| | | | Total | | | |
| Operating Labor Requirements(O.J.)per Shift: | 1 unit/mod. | | Plant | | | |
| Skilled Operator | 2.0 | | 2.0 | | | |
| Operator | 12.0 | | 12.0 | | | |
| Foreman | 1.0 | | 1.0 | | | |
| Lab Tech's, etc. | 3.0 | | 3.0 | | | |
| TOTAL-O.J.'s | 18.0 | | 18.0 | | | |
| | | | | | Annual Cost | Annual Unit Cost |
| | | | | | $ | $/MMBtu |
| Annual Operating Labor Cost | | | | | $7,102,696 | $0.142 |
| Maintenance Labor Cost | | | | | $28,373,931 | $0.566 |
| Administrative & Support Labor | | | | | $8,869,157 | $0.177 |
| Property Taxes and Insurance | | | | | $57,289,567 | $1.144 |
| **TOTAL FIXED OPERATING COSTS** | | | | | **$101,635,351** | **$2.029** |
| VARIABLE OPERATING COSTS | | | | | | |
| | | | | | | $/MMBtu |
| **Maintenance Material Cost** | | | | | **$42,560,897** | **$0.84968** |
| Consumables | Consumption Initial Fill | Consumption /Day | Unit Cost | Initial Fill Cost | | |
| **Water(/1000 gallons)** | 0 | 3,183 | 1.08 | **$0** | **$1,131,059** | **$0.02258** |
| **Chemicals** | | | | | | |
| MU & WT Chem. (lb) | 0 | 18,964 | 0.17 | $0 | $1,078,164 | $0.02152 |
| Carbon (Mercury Removal) (lb) | 216,740 | 371 | 1.05 | $227,614 | $128,033 | $0.00256 |
| COS Catalyst (m3) | 0 | 0 | 2,397.36 | $0 | $0 | $0.00000 |
| Water Gas Shift Catalyst (ft3) | 10,021 | 8.58 | 498.83 | $4,998,731 | $1,405,893 | $0.02807 |
| ZnO Sorbent (ton) | 63 | 0.22 | 12,574.00 | $792,321 | $891,361 | $0.01779 |
| Methanation Catalyst (ft3) | 11,443 | 8.71 | 440.00 | $5,035,097 | $1,258,774 | $0.02513 |
| Selexol Solution (gal) | 484,076 | 154 | 13.40 | $6,485,766 | $676,901 | $0.01351 |
| SCR Catalyst (m3) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Aqueous Ammonia (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Claus Catalyst (ft3) | w/equip | 1.63 | 131.27 | $0 | $70,246 | $0.00140 |
| **Subtotal Chemicals** | | | | **$17,539,529** | **$5,509,373** | **$0.10999** |
| **Other** | | | | | | |
| Supplemental Fuel (MBtu) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for consumption) (M\ | 0 | 0 | 61.60 | $0 | $0 | $0.00000 |
| Gases,N2 etc. (/100scf) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| L.P. Steam (/1000 pounds) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| **Subtotal Other** | | | | **$0** | **$0** | **$0.00000** |
| **Waste Disposal** | | | | | | |
| Spent Mercury Catalyst (lb.) | 0 | 371 | 0.42 | $0 | $50,848 | $0.00102 |
| Spent ZnO Sorbent (ton) | 0 | 0.22 | 16.23 | $0 | $1,150 | $0.00002 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Slag (ton) | 0 | 1,889 | 16.23 | $0 | $10,069,904 | $0.20103 |
| **Subtotal-Waste Disposal** | | | | **$0** | **$10,121,902** | **$0.20207** |
| **By-products & Emissions** | | | | | | |
| Sulfur (tons) | 0 | 116 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for sale) (MWh) | 0 | 1,105 | 58.00 | $0 | -$21,061,948 | -$0.42048 |
| **Subtotal By-Products** | | | | **$0** | **-$21,061,948** | **-$0.42048** |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$17,539,529** | **$38,261,284** | **$0.76384** |
| **Fuel (ton)** | 0 | 18,779 | 10.92 | **$0** | **$67,361,980** | **$1.34480** |

# Exhibit 5-41 Case 8 Total Plant Cost Details

| | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | Client: | USDOE/NETL | | | | | | | | Report Date: | 2010-Feb-18 |
| | Project: | Cost & Performance Baseline for Fossil Energy Plants Vol.2 | | | | | | | | | |
| TOTAL PLANT COST SUMMARY | | | | | | | | | | | |
| | Case: | Case 8 - Siemens Quench SNG Production w/ CO2 | | | | | | | | | |
| | Plant Size: | -3.43 | MW,net | Estimate Type: | | Conceptual | | | Cost Base (Jun) | 2007 | ($x1000) |
| Acct | | Equipment | Material | Labor | | Sales | Bare Erected | Eng'g CM | Contingencies | | Total Plant Cost |
| No. | Item/Description | Cost | Cost | Direct | Indirect | Tax | Cost $ | H.O.& Fee | Process | Project | $ |
| 1 | COAL & SORBENT HANDLING | | | | | | | | | | |
| 1.1 | Coal Receive & Unload | $7,733 | $0 | $3,779 | $0 | $0 | $11,512 | $1,031 | $0 | $2,509 | $15,052 |
| 1.2 | Coal Stackout & Reclaim | $9,993 | $0 | $2,423 | $0 | $0 | $12,415 | $1,088 | $0 | $2,701 | $16,204 |
| 1.3 | Coal Conveyors & Yd Crush | $9,291 | $0 | $2,397 | $0 | $0 | $11,688 | $1,026 | $0 | $2,543 | $15,256 |
| 1.4 | Other Coal Handling | $2,431 | $0 | $555 | $0 | $0 | $2,985 | $261 | $0 | $649 | $3,896 |
| 1.5 | Sorbent Receive & Unload | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.6 | Sorbent Stackout & Reclaim | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.7 | Sorbent Conveyors | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.8 | Other Sorbent Handling | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 1.9 | Coal & Sorbent Hnd.Foundations | $0 | $5,472 | $13,683 | $0 | $0 | $19,155 | $1,836 | $0 | $4,198 | $25,189 |
| | **SUBTOTAL 1.** | **$29,447** | **$5,472** | **$22,836** | **$0** | **$0** | **$57,755** | **$5,243** | **$0** | **$12,600** | **$75,598** |
| 2 | COAL & SORBENT PREP & FEED | | | | | | | | | | |
| 2.1 | Coal Crushing & Drying | $127,899 | $0 | $18,636 | $0 | $0 | $146,535 | $13,308 | $0 | $31,969 | $191,811 |
| 2.2 | Prepared Coal Storage & Feed | $6,058 | $1,450 | $950 | $0 | $0 | $8,457 | $723 | $0 | $1,836 | $11,017 |
| 2.3 | Dry Coal Injection System | $199,367 | $2,314 | $18,515 | $0 | $0 | $220,197 | $18,965 | $0 | $47,832 | $286,994 |
| 2.4 | Misc.Coal Prep & Feed | $3,331 | $2,424 | $7,268 | $0 | $0 | $13,024 | $1,197 | $0 | $2,844 | $17,065 |
| 2.5 | Sorbent Prep Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.6 | Sorbent Storage & Feed | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.7 | Sorbent Injection System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.8 | Booster Air Supply System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 2.9 | Coal & Sorbent Feed Foundation | $0 | $12,948 | $10,631 | $0 | $0 | $23,579 | $2,184 | $0 | $5,153 | $30,916 |
| | **SUBTOTAL 2.** | **$336,655** | **$19,136** | **$56,001** | **$0** | **$0** | **$411,792** | **$36,377** | **$0** | **$89,634** | **$537,802** |
| 3 | FEEDWATER & MISC. BOP SYSTEMS | | | | | | | | | | |
| 3.1 | Feedwater System | $3,454 | $1,605 | $1,947 | $0 | $0 | $7,006 | $633 | $0 | $1,528 | $9,168 |
| 3.2 | Water Makeup & Pretreating | $656 | $69 | $367 | $0 | $0 | $1,092 | $104 | $0 | $359 | $1,555 |
| 3.3 | Other Feedwater Subsystems | $3,750 | $1,134 | $1,855 | $0 | $0 | $6,739 | $612 | $0 | $1,470 | $8,821 |
| 3.4 | Service Water Systems | $376 | $774 | $2,685 | $0 | $0 | $3,834 | $374 | $0 | $1,262 | $5,470 |
| 3.5 | Other Boiler Plant Systems | $2,016 | $781 | $1,936 | $0 | $0 | $4,733 | $449 | $0 | $1,036 | $6,218 |
| 3.6 | FO Supply Sys & Nat Gas | $102 | $210 | $188 | $0 | $0 | $500 | $48 | $0 | $110 | $658 |
| 3.7 | Waste Treatment Equipment | $918 | $0 | $560 | $0 | $0 | $1,477 | $144 | $0 | $486 | $2,108 |
| 3.8 | Misc. Power Plant Equipment | $1,639 | $219 | $841 | $0 | $0 | $2,700 | $261 | $0 | $888 | $3,848 |
| | **SUBTOTAL 3.** | **$12,911** | **$4,792** | **$10,379** | **$0** | **$0** | **$28,081** | **$2,625** | **$0** | **$7,140** | **$37,845** |
| 4 | GASIFIER & ACCESSORIES | | | | | | | | | | |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries | $325,357 | $0 | $150,637 | $0 | $0 | $475,994 | $42,277 | $71,399 | $88,451 | $678,121 |
| 4.2 | Syngas Cooling | w/4.1 | $0 | w/ 4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.3 | ASU/Oxidant Compression | $246,271 | $0 | w/equip. | $0 | $0 | $246,271 | $23,871 | $0 | $27,014 | $297,156 |
| 4.4 | LT Heat Recovery & FG Saturation | $51,285 | $0 | $19,002 | $0 | $0 | $70,287 | $6,733 | $0 | $15,404 | $92,424 |
| 4.5 | Misc. Gasification Equipment | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.6 | Flare Stack System | $0 | $2,260 | $920 | $0 | $0 | $3,180 | $305 | $0 | $697 | $4,182 |
| 4.7 | CO2 Solid Feed System Compressors | $8,698 | $2,087 | $3,131 | $0 | $0 | $13,916 | $1,392 | $0 | $3,062 | $18,369 |
| 4.8 | Major Component Rigging | w/4.1&4.2 | $0 | w/4.1&4.2 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 4.9 | Gasification Foundations | $0 | $26,928 | $15,365 | $0 | $0 | $42,293 | $3,872 | $0 | $11,541 | $57,706 |
| | **SUBTOTAL 4.** | **$631,611** | **$31,276** | **$189,055** | **$0** | **$0** | **$851,942** | **$78,449** | **$71,399** | **$146,169** | **$1,147,958** |

## Exhibit 5-41 Case 8 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 5A | GAS CLEANUP & PIPING | | | | | | | | | | |
| 5A.1 | Double Stage Selexol | $130,462 | $0 | $110,700 | $0 | $0 | $241,162 | $23,323 | $48,232 | $62,543 | $375,261 |
| 5A.2 | Elemental Sulfur Plant | $8,848 | $1,764 | $11,416 | $0 | $0 | $22,028 | $2,140 | $0 | $4,834 | $29,001 |
| 5A.3 | Mercury Removal | $2,347 | $0 | $1,786 | $0 | $0 | $4,134 | $399 | $207 | $948 | $5,687 |
| 5A.4 | Shift Reactors | $10,263 | $0 | $4,131 | $0 | $0 | $14,393 | $1,380 | $0 | $3,155 | $18,928 |
| 5A.5 | Methanation | $26,350 | $10,540 | $15,810 | $0 | $0 | $52,700 | $5,270 | $5,270 | $12,648 | $75,888 |
| 5A.6 | SNG Purification & Compression | $18,600 | $7,440 | $11,160 | $0 | $0 | $37,200 | $3,720 | $0 | $8,184 | $49,104 |
| 5A.7 | Fuel Gas Piping | $0 | $1,630 | $1,141 | $0 | $0 | $2,772 | $257 | $0 | $606 | $3,634 |
| 5A.9 | Process Interconnects | $0 | $12,000 | $18,000 | $0 | $0 | $30,000 | $3,000 | $0 | $6,600 | $39,600 |
| 5A.10 | HGCU Foundations | $0 | $1,651 | $1,064 | $0 | $0 | $2,715 | $249 | $0 | $889 | $3,854 |
| 5A.11 | Zinc Oxide Guard Bed | $758 | $0 | $140 | $0 | $0 | $897 | $90 | $0 | $197 | $1,185 |
| | **SUBTOTAL 5A.** | **$197,628** | **$35,024** | **$175,348** | **$0** | **$0** | **$408,000** | **$39,828** | **$53,709** | **$100,604** | **$602,141** |
| 5B | CO2 COMPRESSION | | | | | | | | | | |
| 5B.1 | CO2 Removal System | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 5B.2 | CO2 Compression & Drying | $37,246 | $0 | $21,934 | $0 | $0 | $59,180 | $5,698 | $0 | $12,976 | $77,853 |
| | **SUBTOTAL 5B.** | **$37,246** | **$0** | **$21,934** | **$0** | **$0** | **$59,180** | **$5,698** | **$0** | **$12,976** | **$77,853** |
| 6 | COMBUSTION TURBINE/ACCESSORIES | | | | | | | | | | |
| 6.1 | Combustion Turbine Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.2 | Syngas Expander | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.3 | Compressed Air Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 6.9 | Combustion Turbine Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 6.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 7 | HRSG, DUCTING & STACK | | | | | | | | | | |
| 7.1 | Heat Recovery Steam Generator | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.2 | HRSG Accessories | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.4 | Stack | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| | **SUBTOTAL 7.** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** | **$0** |
| 8 | STEAM TURBINE GENERATOR | | | | | | | | | | |
| 8.1 | Steam TG & Accessories | $31,587 | $0 | $5,482 | $0 | $0 | $37,069 | $3,557 | $0 | $4,063 | $44,689 |
| 8.2 | Turbine Plant Auxiliaries | $171 | $0 | $504 | $0 | $0 | $675 | $66 | $0 | $74 | $815 |
| 8.3a | Condenser & Auxiliaries | $3,352 | $0 | $1,548 | $0 | $0 | $4,899 | $470 | $0 | $537 | $5,907 |
| 8.3b | Air Cooled Condenser | $30,717 | $0 | $6,158 | $0 | $0 | $36,875 | $3,687 | $0 | $8,112 | $48,675 |
| 8.4 | Steam Piping | $5,449 | $0 | $3,833 | $0 | $0 | $9,283 | $797 | $0 | $2,520 | $12,600 |
| 8.9 | TG Foundations | $0 | $1,090 | $1,843 | $0 | $0 | $2,934 | $278 | $0 | $964 | $4,175 |
| | **SUBTOTAL 8.** | **$71,275** | **$1,090** | **$19,369** | **$0** | **$0** | **$91,735** | **$8,857** | **$0** | **$16,270** | **$116,862** |

## Exhibit 5-41 Case 8 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 9 | COOLING WATER SYSTEM | | | | | | | | | | |
| 9.1 | Cooling Towers | $7,817 | $0 | $1,422 | $0 | $0 | $9,239 | $880 | $0 | $1,518 | $11,637 |
| 9.2 | Circulating Water Pumps | $2,078 | $0 | $88 | $0 | $0 | $2,166 | $182 | $0 | $352 | $2,700 |
| 9.3 | Circ.Water System Auxiliaries | $168 | $0 | $24 | $0 | $0 | $192 | $18 | $0 | $32 | $242 |
| 9.4 | Circ.Water Piping | $0 | $7,009 | $1,817 | $0 | $0 | $8,826 | $798 | $0 | $1,925 | $11,548 |
| 9.5 | Make-up Water System | $360 | $0 | $515 | $0 | $0 | $875 | $84 | $0 | $192 | $1,150 |
| 9.6 | Component Cooling Water Sys | $827 | $990 | $704 | $0 | $0 | $2,521 | $236 | $0 | $551 | $3,309 |
| 9.9 | Circ.Water System Foundations | $0 | $2,566 | $4,363 | $0 | $0 | $6,929 | $657 | $0 | $2,276 | $9,862 |
| | **SUBTOTAL 9.** | **$11,250** | **$10,565** | **$8,933** | **$0** | **$0** | **$30,748** | **$2,855** | **$0** | **$6,845** | **$40,448** |
| 10 | ASH/SPENT SORBENT HANDLING SYS | | | | | | | | | | |
| 10.1 | Slag Dewatering & Cooling | $49,571 | $0 | $48,892 | $0 | $0 | $98,463 | $9,540 | $0 | $10,800 | $118,804 |
| 10.2 | Gasifier Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.3 | Cleanup Ash Depressurization | w/10.1 | w/10.1 | w/10.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.4 | High Temperature Ash Piping | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.5 | Other Ash Recovery Equipment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 10.6 | Ash Storage Silos | $1,072 | $0 | $1,167 | $0 | $0 | $2,239 | $217 | $0 | $368 | $2,825 |
| 10.7 | Ash Transport & Feed Equipment | $1,438 | $0 | $347 | $0 | $0 | $1,785 | $167 | $0 | $293 | $2,245 |
| 10.8 | Misc. Ash Handling Equipment | $2,221 | $2,722 | $813 | $0 | $0 | $5,756 | $548 | $0 | $946 | $7,250 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $95 | $119 | $0 | $0 | $214 | $20 | $0 | $70 | $304 |
| | **SUBTOTAL 10.** | **$54,303** | **$2,817** | **$51,338** | **$0** | **$0** | **$108,458** | **$10,492** | **$0** | **$12,477** | **$131,427** |
| 11 | ACCESSORY ELECTRIC PLANT | | | | | | | | | | |
| 11.1 | Generator Equipment | $435 | $0 | $430 | $0 | $0 | $866 | $83 | $0 | $95 | $1,043 |
| 11.2 | Station Service Equipment | $6,156 | $0 | $555 | $0 | $0 | $6,711 | $619 | $0 | $733 | $8,062 |
| 11.3 | Switchgear & Motor Control | $11,381 | $0 | $2,070 | $0 | $0 | $13,451 | $1,248 | $0 | $2,205 | $16,903 |
| 11.4 | Conduit & Cable Tray | $0 | $5,287 | $17,441 | $0 | $0 | $22,728 | $2,198 | $0 | $6,231 | $31,157 |
| 11.5 | Wire & Cable | $0 | $10,101 | $6,637 | $0 | $0 | $16,738 | $1,216 | $0 | $4,489 | $22,443 |
| 11.6 | Protective Equipment | $0 | $807 | $2,936 | $0 | $0 | $3,743 | $366 | $0 | $616 | $4,724 |
| 11.7 | Standby Equipment | $317 | $0 | $309 | $0 | $0 | $626 | $60 | $0 | $103 | $789 |
| 11.8 | Main Power Transformers | $10,063 | $0 | $172 | $0 | $0 | $10,234 | $776 | $0 | $1,652 | $12,662 |
| 11.9 | Electrical Foundations | $0 | $186 | $489 | $0 | $0 | $675 | $65 | $0 | $222 | $962 |
| | **SUBTOTAL 11.** | **$28,352** | **$16,381** | **$31,039** | **$0** | **$0** | **$75,772** | **$6,629** | **$0** | **$16,345** | **$98,747** |
| 12 | INSTRUMENTATION & CONTROL | | | | | | | | | | |
| 12.1 | IGCC Control Equipment | w/4.1 | $0 | w/4.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.2 | Combustion Turbine Control | w/6.1 | $0 | w/6.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.3 | Steam Turbine Control | w/8.1 | $0 | w/8.1 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.4 | Other Major Component Control | $516 | $0 | $345 | $0 | $0 | $860 | $81 | $43 | $148 | $1,133 |
| 12.5 | Signal Processing Equipment | w/12.7 | $0 | w/12.7 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 12.6 | Control Boards,Panels & Racks | $343 | $0 | $220 | $0 | $0 | $562 | $53 | $28 | $129 | $772 |
| 12.7 | Computer & Accessories | $5,483 | $0 | $176 | $0 | $0 | $5,659 | $519 | $283 | $646 | $7,107 |
| 12.8 | Instrument Wiring & Tubing | $0 | $2,778 | $5,678 | $0 | $0 | $8,456 | $717 | $423 | $2,399 | $11,995 |
| 12.9 | Other I & C Equipment | $4,088 | $0 | $1,985 | $0 | $0 | $6,074 | $571 | $304 | $1,042 | $7,991 |
| | **SUBTOTAL 12.** | **$10,430** | **$2,778** | **$8,403** | **$0** | **$0** | **$21,611** | **$1,943** | **$1,081** | **$4,364** | **$28,998** |

## Exhibit 5-41 Case 8 Total Plant Cost Details (continued)

| Acct No. | Item/Description | Equipment Cost | Material Cost | Labor Direct | Labor Indirect | Sales Tax | Bare Erected Cost $ | Eng'g CM H.O.& Fee | Contingencies Process | Contingencies Project | Total Plant Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 13 | IMPROVEMENTS TO SITE | | | | | | | | | | |
| 13.1 | Site Preparation | $0 | $141 | $3,021 | $0 | $0 | $3,162 | $314 | $0 | $1,043 | $4,519 |
| 13.2 | Site Improvements | $0 | $2,514 | $3,341 | $0 | $0 | $5,855 | $578 | $0 | $1,930 | $8,362 |
| 13.3 | Site Facilities | $4,505 | $0 | $4,754 | $0 | $0 | $9,259 | $913 | $0 | $3,051 | $13,223 |
| | **SUBTOTAL 13.** | **$4,505** | **$2,655** | **$11,115** | **$0** | **$0** | **$18,276** | **$1,804** | **$0** | **$6,024** | **$26,104** |
| 14 | BUILDINGS & STRUCTURES | | | | | | | | | | |
| 14.1 | Combustion Turbine Area | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| 14.2 | Steam Turbine Building | $0 | $2,623 | $3,737 | $0 | $0 | $6,360 | $585 | $0 | $1,042 | $7,987 |
| 14.3 | Administration Building | $0 | $870 | $630 | $0 | $0 | $1,500 | $134 | $0 | $245 | $1,879 |
| 14.4 | Circulation Water Pumphouse | $0 | $160 | $90 | $0 | $0 | $250 | $22 | $0 | $41 | $312 |
| 14.5 | Water Treatment Buildings | $0 | $549 | $535 | $0 | $0 | $1,084 | $98 | $0 | $177 | $1,359 |
| 14.6 | Machine Shop | $0 | $445 | $305 | $0 | $0 | $750 | $67 | $0 | $123 | $939 |
| 14.7 | Warehouse | $0 | $726 | $464 | $0 | $0 | $1,190 | $105 | $0 | $194 | $1,489 |
| 14.8 | Other Buildings & Structures | $0 | $514 | $400 | $0 | $0 | $914 | $82 | $0 | $199 | $1,195 |
| 14.9 | Waste Treating Building & Str. | $0 | $1,149 | $2,196 | $0 | $0 | $3,346 | $312 | $0 | $732 | $4,389 |
| | **SUBTOTAL 14.** | **$0** | **$7,036** | **$8,358** | **$0** | **$0** | **$15,394** | **$1,404** | **$0** | **$2,752** | **$19,550** |
| | | | | | | | | | | | |
| | **TOTAL COST** | **$1,425,614** | **$139,022** | **$614,108** | **$0** | **$0** | **$2,178,744** | **$202,203** | **$126,189** | **$434,199** | **$2,941,335** |
| | | | | | | | | | | | |
| | **Owner's Costs** | | | | | | | | | | |
| | **Preproduction Costs** | | | | | | | | | | |
| | 6 Months All Labor | | | | | | | | | | $22,166 |
| | 1 Month Maintenance Materials | | | | | | | | | | $3,939 |
| | 1 Month Non-fuel Consumables | | | | | | | | | | $767 |
| | 1 Month Waste Disposal | | | | | | | | | | $937 |
| | 25% of 1 Months Fuel Cost at 100% CF | | | | | | | | | | $1,559 |
| | 2% of TPC | | | | | | | | | | $58,827 |
| | **Total** | | | | | | | | | | **$88,196** |
| | **Inventory Capital** | | | | | | | | | | |
| | 60 day supply of fuel and consumables at 100% CF | | | | | | | | | | $13,457 |
| | 0.5% of TPC (spare parts) | | | | | | | | | | $14,707 |
| | **Total** | | | | | | | | | | **$28,164** |
| | | | | | | | | | | | |
| | **Initial Cost for Catalyst and Chemicals** | | | | | | | | | | $16,258 |
| | **Land** | | | | | | | | | | $900 |
| | **Other Owner's Costs** | | | | | | | | | | $441,200 |
| | **Financing Costs** | | | | | | | | | | $79,416 |
| | **Total Overnight Costs (TOC)** | | | | | | | | | | **$3,595,468** |
| | TASC Multiplier | | | | | | | | | | 1.201 |
| | **Total As-Spent Cost (TASC)** | | | | | | | | | | **$4,318,158** |

# Exhibit 5-42 Case 8 Initial and Annual O&M Costs

| | | | | | | |
|---|---|---|---|---|---|---|
| INITIAL & ANNUAL O&M EXPENSES | | | | | Cost Base (Jun): | 2007 |
| Case 8 - Siemens Quench SNG Production w/ CO2 | | | | | MWe-net: | -3 |
| SNG (MMbtu/hr): | 6349 | | | | Capacity Factor (%): | 90 |
| OPERATING & MAINTENANCE LABOR | | | | | | |
| Operating Labor | | | | | | |
| Operating Labor Rate(base): | 34.65 | $/hour | | | | |
| Operating Labor Burden: | 30.00 | % of base | | | | |
| Labor O-H Charge Rate: | 25.00 | % of labor | | | | |
| | | | Total | | | |
| Operating Labor Requirements(O.J.)per Shift: | 1 unit/mod. | | Plant | | | |
| Skilled Operator | 2.0 | | 2.0 | | | |
| Operator | 12.0 | | 12.0 | | | |
| Foreman | 1.0 | | 1.0 | | | |
| Lab Tech's, etc. | 3.0 | | 3.0 | | | |
| TOTAL-O.J.'s | 18.0 | | 18.0 | | | |
| | | | | | Annual Cost | Annual Unit Cost |
| | | | | | $ | $/MMBtu |
| Annual Operating Labor Cost | | | | | $7,102,696 | $0.142 |
| Maintenance Labor Cost | | | | | $28,362,952 | $0.567 |
| Administrative & Support Labor | | | | | $8,866,412 | $0.177 |
| Property Taxes and Insurance | | | | | $58,826,693 | $1.175 |
| **TOTAL FIXED OPERATING COSTS** | | | | | **$103,158,753** | **$2.061** |
| VARIABLE OPERATING COSTS | | | | | | |
| | | | | | | $/MMBtu |
| **Maintenance Material Cost** | | | | | **$42,544,428** | **$0.84994** |
| Consumables | Consumption | | Unit | Initial Fill | | |
| | Initial Fill | /Day | Cost | Cost | | |
| **Water(/1000 gallons)** | 0 | 3,694 | 1.08 | **$0** | **$1,312,704** | **$0.02622** |
| **Chemicals** | | | | | | |
| MU & WT Chem.(lb) | 0 | 22,010 | 0.17 | $0 | $1,251,314 | $0.02500 |
| Carbon (Mercury Removal) (lb) | 214,355 | 367 | 1.05 | $225,110 | $126,624 | $0.00253 |
| COS Catalyst (m3) | 0 | 0 | 2,397.36 | $0 | $0 | $0.00000 |
| Water Gas Shift Catalyst (ft3) | 7,580 | 6.49 | 498.83 | $3,780,914 | $1,063,382 | $0.02124 |
| ZnO Sorbent (ton) | 61 | 0.21 | 12,574.00 | $772,815 | $869,417 | $0.01737 |
| Methanation Catalyst (ft3) | 11,347 | 8.64 | 440.00 | $4,992,625 | $1,248,156 | $0.02494 |
| Selexol Solution (gal) | 484,145 | 154 | 13.40 | $6,486,691 | $676,998 | $0.01352 |
| SCR Catalyst (m3) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Aqueous Ammonia (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Claus Catalyst (ft3) | w/equip | 1.63 | 131.27 | $0 | $70,295 | $0.00140 |
| **Subtotal Chemicals** | | | | **$16,258,155** | **$5,306,187** | **$0.10601** |
| **Other** | | | | | | |
| Supplemental Fuel (MBtu) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for consumption) (M | 0 | 82 | 61.60 | $0 | $1,665,795 | $0.03328 |
| Gases,N2 etc.(/100scf) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| L.P. Steam(/1000 pounds) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| **Subtotal Other** | | | | **$0** | **$1,665,795** | **$0.03328** |
| **Waste Disposal** | | | | | | |
| Spent Mercury Catalyst (lb.) | 0 | 367 | 0.42 | $0 | $50,289 | $0.00100 |
| Spent ZnO Sorbent (ton) | 0 | 0.21 | 16.23 | $0 | $1,122 | $0.00002 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Slag (ton) | 0 | 1,889 | 16.23 | $0 | $10,069,904 | $0.20117 |
| **Subtotal-Waste Disposal** | | | | **$0** | **$10,121,315** | **$0.20220** |
| **By-products & Emissions** | | | | | | |
| Sulfur (tons) | 0 | 116 | 0.00 | $0 | $0 | $0.00000 |
| Supplemental Electricity (for sale) (MWh) | 0 | 0 | 58.00 | $0 | $0 | $0.00000 |
| **Subtotal By-Products** | | | | **$0** | **$0** | **$0.00000** |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$16,258,155** | **$60,950,429** | **$1.21765** |
| **Fuel (ton)** | 0 | 18,779 | 10.92 | **$0** | **$67,361,980** | **$1.34574** |

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# 6. SUMMARY

## 6.1 PERFORMANCE

The nominal thermal input for the Siemens gasifier used for all cases in this study is 500 MW, but actual input varies with coal type. The bituminous coal cases in this study had a thermal input of 549 MW per gasifier, the PRB cases 527 MW and the lignite cases 506 MW. For the co-production cases, a portion of the syngas was used to produce 2,000 tonnes/day (2,204 tons/day) of ammonia with the remaining syngas producing approximately 1.2 $BNm^3$/year (42 Bscf/year) of SNG at 90 percent CF.

The performance summaries for all cases are shown in Exhibit 6-1. The HHV efficiencies are shown in Exhibit 6-2. The higher and lower heating values (LHVs) of the SNG production are shown in Exhibit 6-3.

- The cases using PRB coal have the highest net conversion efficiency at 63.1 percent and the highest SNG product heating value at 973 Btu/scf (HHV) and 876 Btu/scf (LHV). The lower nitrogen content of the lower rank PRB and lignite coals aids in boosting the HHV of the SNG end-product, and consequently the net conversion efficiency.
- The remaining six cases have essentially constant conversion efficiency (61.3 to 61.5 percent). The addition of carbon sequestration has only a minimal impact on conversion efficiency. The sequestration impact is seen in the reduced available export power relative to the non-sequestration case.

## Exhibit 6-1 Performance, Emissions, and Cost Results

| | Illinois No. 6 Coal | | | | Rosebud PRB | | North Dakota Lignite | |
|---|---|---|---|---|---|---|---|---|
| | SNG Only | | SNG and Ammonia | | SNG Only | | SNG Only | |
| **PERFORMANCE** | **Case 1** | **Case 2** | **Case 3** | **Case 4** | **Case 5** | **Case 6** | **Case 7** | **Case 8** |
| **$CO_2$ Sequestration** | No | Yes | No | Yes | No | Yes | No | Yes |
| **SNG Production (Bscf/year)[1]** | 57 | 56 | 42 | 42 | 55 | 55 | 52 | 52 |
| **Ammonia Production (TPD)[1]** | N/A | N/A | 2,204 | 2,204 | N/A | N/A | N/A | N/A |
| **HHV Conversion Efficiency, %** | 61.4% | 61.3% | 61.5% | 61.4% | 63.1% | 63.1% | 61.5% | 61.5% |
| **Gross Power Output ($kW_e$)** | 308,000 | 310,600 | 292,300 | 292,300 | 302,000 | 302,000 | 305,800 | 310,500 |
| **Auxiliary Power Requirement ($kW_e$)** | 216,350 | 262,090 | 263,600 | 316,030 | 247,570 | 300,190 | 259,740 | 313,930 |
| **Net Power Output ($kW_e$)** | 91,650 | 48,510 | 28,700 | -23,730 | 54,430 | 1,810 | 46,060 | -3,430 |
| **Coal Flowrate (lb/hr)** | 964,752 | 964,752 | 964,000 | 964,000 | 1,259,331 | 1,259,331 | 1,564,932 | 1,564,932 |
| **HHV Thermal Input ($kW_{th}$)** | 3,298,455 | 3,298,455 | 3,295,885 | 3,295,885 | 3,160,745 | 3,160,745 | 3,034,796 | 3,034,796 |
| **Raw Water Withdrawal (gpm)** | 7,169 | 7,123 | 7,434 | 7,458 | 4,853 | 5,509 | 4,421 | 5,131 |
| **Process Water Discharge (gpm)** | 1,462 | 1,451 | 1,455 | 1,461 | 1,071 | 1,218 | 1,071 | 1,230 |
| **Raw Water Consumption (gpm)** | 5,708 | 5,672 | 5,979 | 5,997 | 3,783 | 4,291 | 3,350 | 3,901 |
| **$CO_2$ Emissions (lb/MMBtu)[2]** | 128 | 5 | 147 | 10 | 140 | 0.7 | 147 | 0.9 |
| **$SO_2$ Emissions (lb/MMBtu)[2]** | 0.0274 | 0.0003 | 0.0229 | 0.0006 | 0.0218 | 0.0000 | 0.0239 | 0.0000 |
| **NOx Emissions (lb/MMBtu)[2]** | Negligible | Negligible | Negligible | Negligible | Negligible | Negligible | Negligible | Negligible |
| **PM Emissions (lb/MMBtu)[2]** | 0.007 | 0.007 | 0.007 | 0.007 | 0.007 | 0.007 | 0.007 | 0.007 |
| **Hg Emissions (lb/MMBtu)[2]** | 5.71E-07 | 5.71E-07 | 5.71E-07 | 5.71E-07 | 3.51E-07 | 3.51E-07 | 5.60E-07 | 5.60E-07 |
| **COST** | | | | | | | | |
| **Total Plant Cost ($ x 1,000)** | 2,628,754 | 2,692,997 | 3,048,463 | 3,119,611 | 2,741,044 | 2,813,258 | 2,864,478 | 2,941,335 |
| **Total Overnight Cost ($ x 1,000)** | 3,235,262 | 3,312,740 | 3,742,411 | 3,829,817 | 3,354,442 | 3,442,310 | 3,504,262 | 3,595,468 |
| **Total As-spent Capital ($ x 1,000)** | 3,885,549 | 3,978,601 | 4,494,636 | 4,599,610 | 4,028,684 | 4,134,215 | 4,208,619 | 4,318,158 |
| **SNG FYCOP ($/MMBtu)[1]** | 19.27 | 20.95 | 15.82 | 17.85 | 19.15 | 21.01 | 21.27 | 23.24 |
| ***$CO_2$ TS&M Costs*** | *0.00* | *0.91* | *0.00* | *1.29* | *0.00* | *0.96* | *0.00* | *1.03* |
| ***Fuel Costs*** | *2.67* | *2.67* | *1.89* | *1.89* | *1.41* | *1.41* | *1.34* | *1.35* |
| ***Variable Costs*** | *1.03* | *1.04* | *0.79* | *0.82* | *1.02* | *1.03* | *1.18* | *1.18* |
| ***Fixed Costs*** | *1.76* | *1.79* | *1.41* | *1.45* | *1.84* | *1.86* | *2.03* | *2.06* |
| ***Electricity Costs*** | *-0.77* | *-0.41* | *-0.17* | *0.15* | *-0.46* | *-0.02* | *-0.42* | *0.03* |
| ***Capital Costs*** | *14.58* | *14.94* | *11.90* | *12.25* | *15.34* | *15.76* | *17.13* | *17.59* |
| **Ammonia FYCOP[3] ($/ton)** | - | - | 799 | 828 | - | - | - | - |

[1] Based on a capacity factor of 90 percent for all cases

[2] Based on coal thermal input

[3] Ammonia price is correlated to historic natural gas costs

**Exhibit 6-2 Net Plant Conversion Efficiency (HHV)**

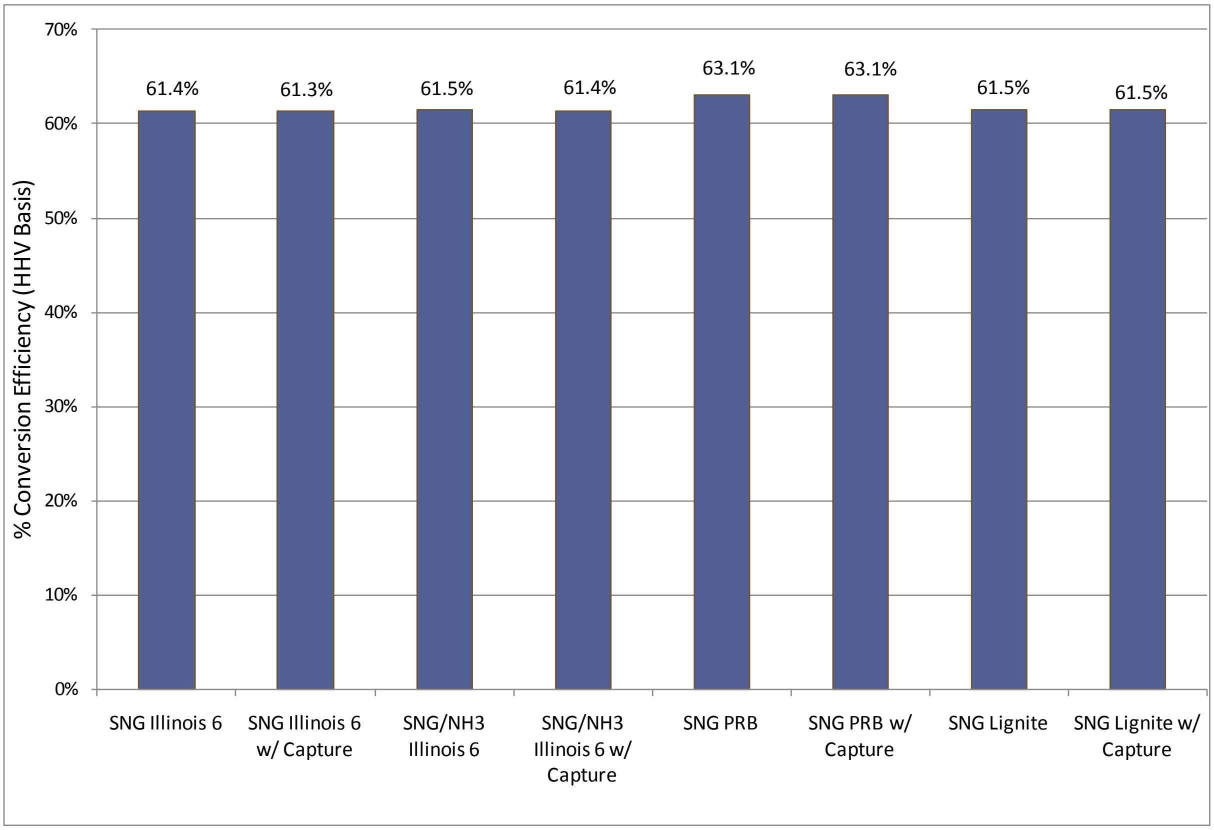
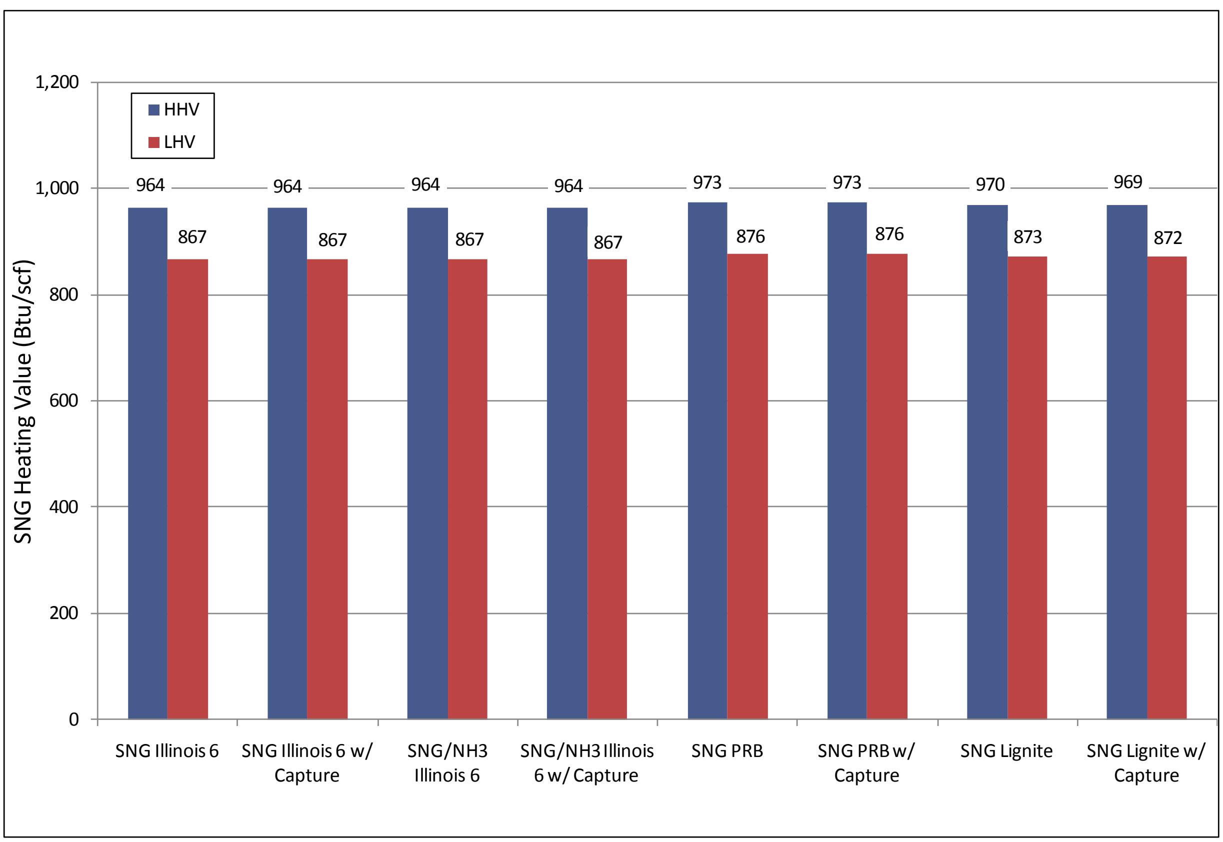



**Exhibit 6-3 SNG Heating Values**



## 6.2 ECONOMICS

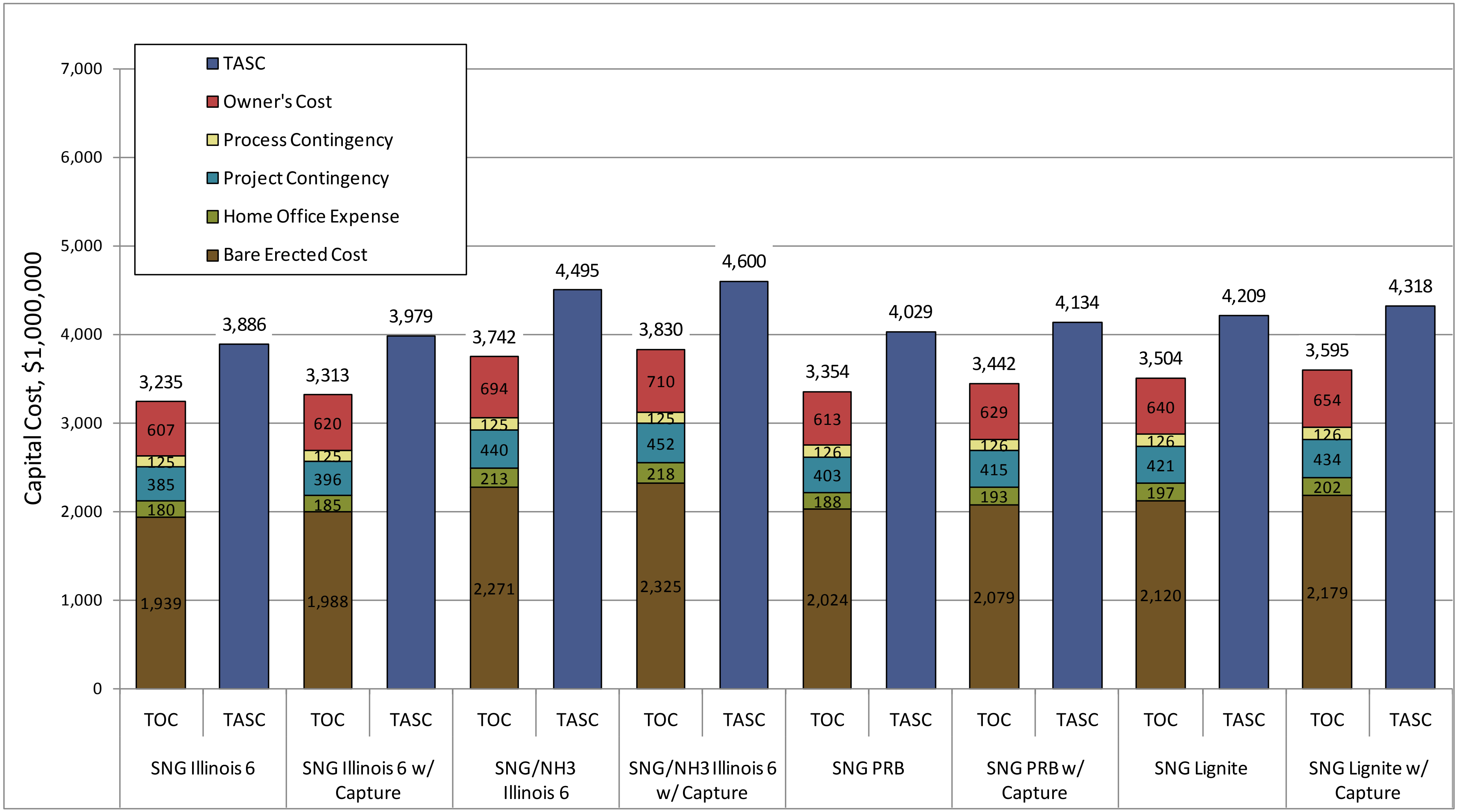
The TOCs are shown in Exhibit 6-4. For reference, Total As-spent Capital (TASC) is also shown. TASC is the sum of TOC, escalation during the capital expenditure period, and interest on debt during the capital expenditure period. Observations related to TOC include:

- In the SNG only cases, the TOC is lowest for the Illinois No. 6 cases followed by the PRB cases, and then by the lignite cases.
- The TOC increase to add carbon sequestration is approximately 2.4 percent for the Illinois No. 6 SNG only case, 2.4 percent for the co-production case, 2.6 percent for the PRB coal SNG only case, and 2.6 percent for the SNG only lignite case.
- The TOC is increased by 16 percent with the addition of ammonia co-production for both the Illinois No. 6 coal non-sequestration and sequestration cases.
- The lignite cases have 4 percent higher TOC compared to the PRB cases.

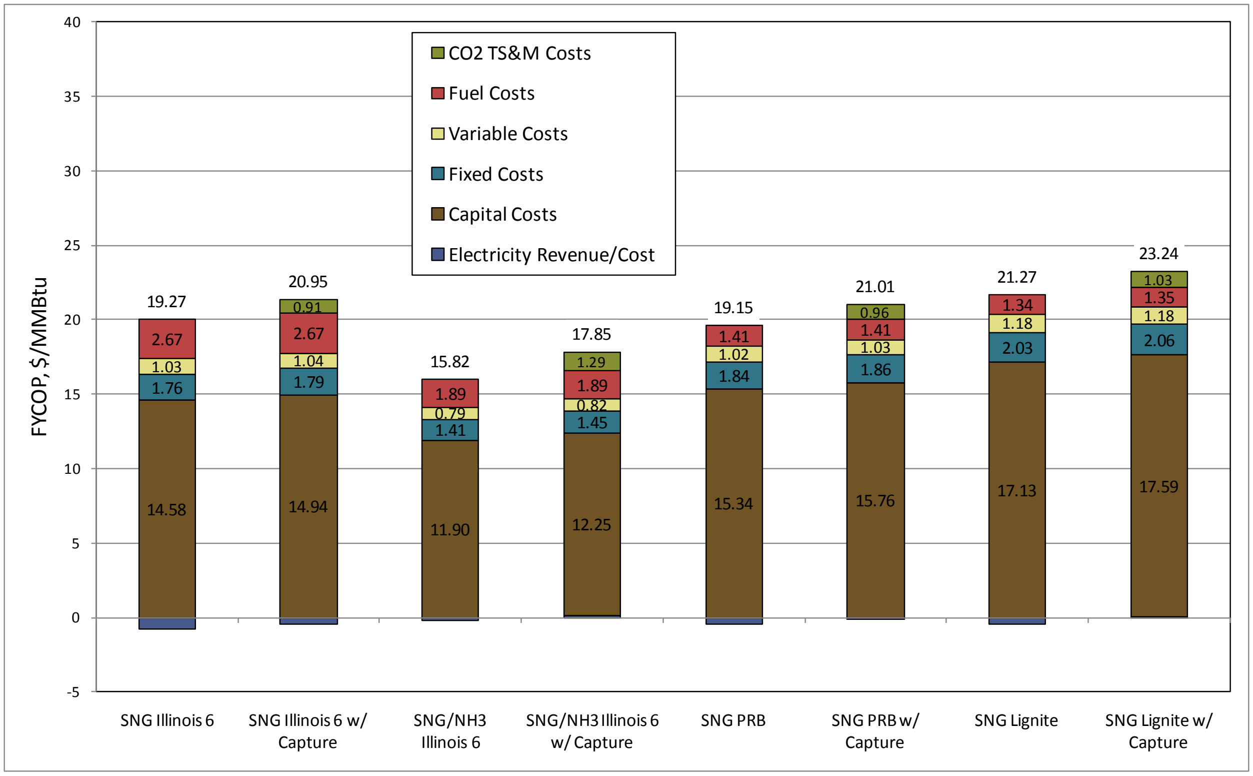
The FYCOPs are shown in Exhibit 6-5. The FYCOP components and SNG/ammonia product revenues are shown in tabular form in Exhibit 6-6. The following observations can be made:

- The FYCOP is dominated by capital charges in all cases. For the SNG only cases, the capital charges (which include owner's costs) range from 71 to 81 percent of the total. For the co-production cases, the capital costs are 75 and 69 percent of the FYCOP for the non-sequestration and sequestration cases, respectively.
- The fuel cost component is highest for the Illinois No. 6 coal cases. For the SNG only cases the fuel represents 13 percent of the total cost for Illinois No. 6 coal, 7 percent for PRB coal, and 6 percent for lignite coal.
- The TS&M component is relatively minor for all cases, ranging from 4 percent to 7 percent.

**Exhibit 6-4 Total Overnight Cost and Total As-spent Capital**



## Exhibit 6-5 First Year Cost of Production




## Exhibit 6-6 First Year Cost of Production Components

| FYCOP Component[1] | SNG Bituminous No Capture (Case 1) | SNG Bituminous w/ Capture (Case 2) | SNG and Ammonia Bituminous No Capture (Case 3) | SNG and Ammonia Bituminous w/ Capture (Case 4) | SNG PRB No Capture (Case 5) | SNG PRB w/ Capture (Case 6) | SNG Lignite No Capture (Case 7) | SNG Lignite w/ Capture (Case 8) |
|---|---|---|---|---|---|---|---|---|
| Capital, $/MMBtu | 14.58 | 14.94 | 11.90 | 12.25 | 15.34 | 15.76 | 17.13 | 17.59 |
| Fixed O&M, $/MMBtu | 1.76 | 1.79 | 1.41 | 1.45 | 1.84 | 1.86 | 2.03 | 2.06 |
| Variable O&M, $/MMBtu | 1.03 | 1.04 | 0.79 | 0.82 | 1.02 | 1.03 | 1.18 | 1.18 |
| Fuel, $/MMBtu | 2.67 | 2.67 | 1.89 | 1.89 | 1.41 | 1.41 | 1.34 | 1.35 |
| Electricity, $/MMBtu | -0.77 | -0.41 | -0.17 | 0.15 | -0.46 | -0.02 | -0.42 | 0.03 |
| $CO_2$ TS&M, $/MMBtu | 0.00 | 0.91 | 0.00 | 1.29 | 0.00 | 0.96 | 0.00 | 1.03 |
| **Total, $/MMBtu** | **19.27** | **20.95** | **15.82** | **17.85** | **19.15** | **21.01** | **21.27** | **23.24** |
| SNG First Year Revenues, MM$ | 1,047 | 1,137 | 639 | 721 | 1,025 | 1,124 | 1,065 | 1,163 |
| Ammonia First Year Revenues, MM$ | - | - | 578 | 599 | - | - | - | - |

[1] Based on heating value of SNG

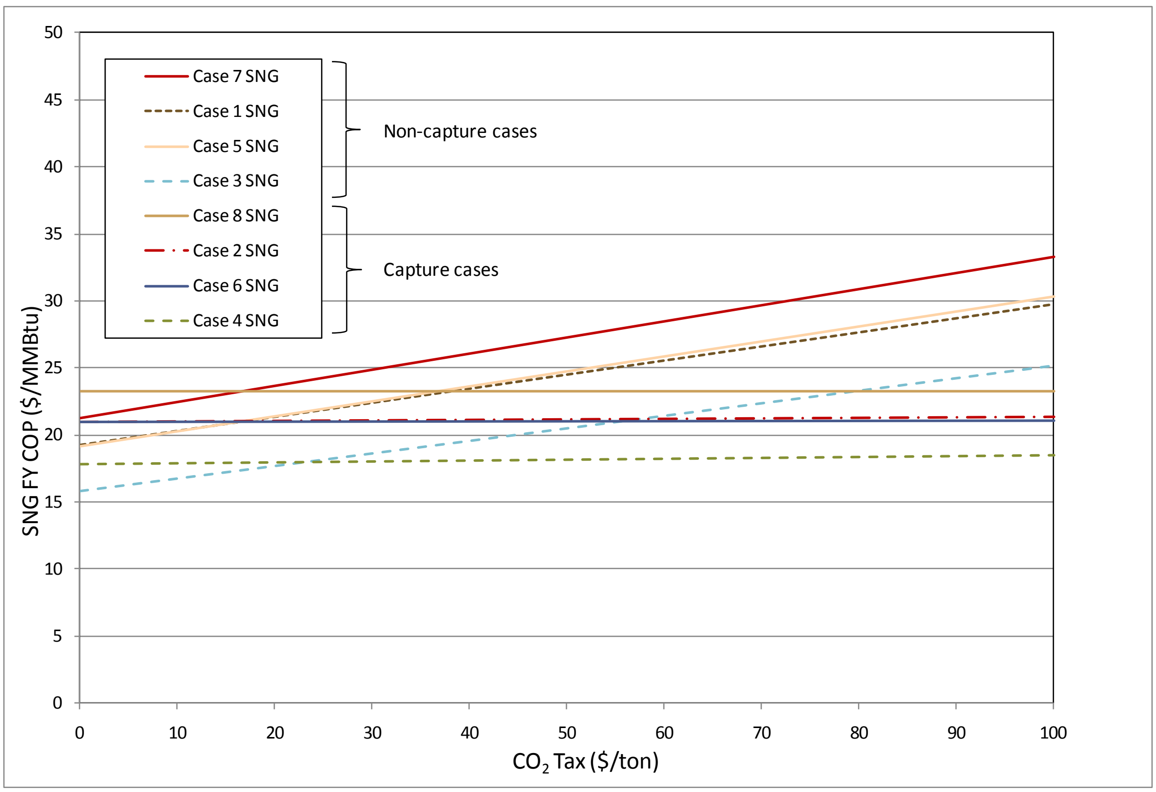
The impact of various levels of $CO_2$ tax on the FYCOP of SNG as well as the FYCOP of ammonia in the co-production cases is shown in Exhibit 6-7. In the co-production cases the cost of ammonia varies with the cost of SNG according to the following relationship:

$$\text{Ammonia Price Paid by Farmers}\left(\frac{\$}{\text{ton}}\right) = 39.78 * \text{Cost of Industrial NG}\left(\frac{\$}{\text{MMBtu}}\right) + 169.29$$

Cases without sequestration are shown as solid lines while those with sequestration are shown as dashed lines. Taxation of plant $CO_2$ emissions causes a significant increase in FYCOP in the non-sequestration cases while the increase is greatly diminished in the sequestration cases. It should be noted that end-use $CO_2$ emissions from the SNG product are not accounted for.

Exhibit 6-7 shows that a carbon tax of $15-20/ton of $CO_2$ is required to make SNG produced in a sequestration case less expensive than that produced in the corresponding non-sequestration case for SNG only cases. The carbon cost must increase to about $25/ton before the same is true in the co-production case.

**Exhibit 6-7 Impact of $CO_2$ Tax on FYCOP**

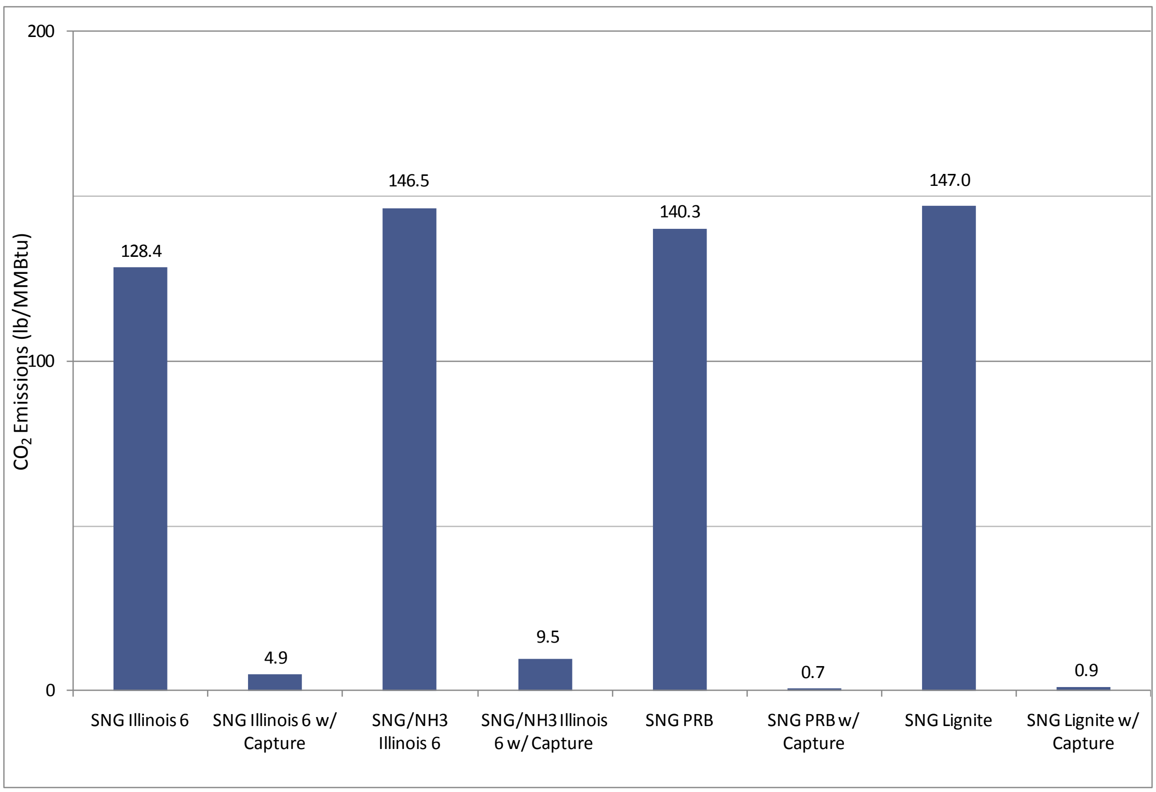



## 6.3 CARBON DIOXIDE EMISSIONS/MITIGATION

Carbon dioxide emissions are not currently regulated nationwide; however, there is increasing momentum for establishing carbon limits. $CO_2$ emissions are presented in Exhibit 6-8 for each case, normalized by coal thermal input to enable comparison between the technology types (SNG only versus co-production). The following conclusions can be drawn:

- In cases with no carbon sequestration, the SNG only case using bituminous coal emits 12 percent less $CO_2$ than the co-production case, 13 percent less than the lignite case, and 8 percent less than the PRB case.
- The $CO_2$ emissions for the SNG only non-sequestration cases are proportional to the carbon intensity of the design coal. Based on fuel compositions used in this study, bituminous coal contains 55 pounds of carbon per million Btu, PRB and lignite coal contain 58 and 60 pounds per million Btu, respectively.
- The normalized $CO_2$ emissions in the sequestration cases are lower for the low rank coal cases largely because the coal drying method utilized does not require the burning of syngas. The normalized $CO_2$ emissions for the Illinois No. 6 co-production cases (both non-sequestration and sequestration) are higher than the equivalent SNG only cases because more $CO_2$ is produced in the shift section of the ammonia plant due to the hydrogen requirements for ammonia production that is not subsequently captured.

**Exhibit 6-8 $CO_2$ Emissions Normalized by Thermal Input**



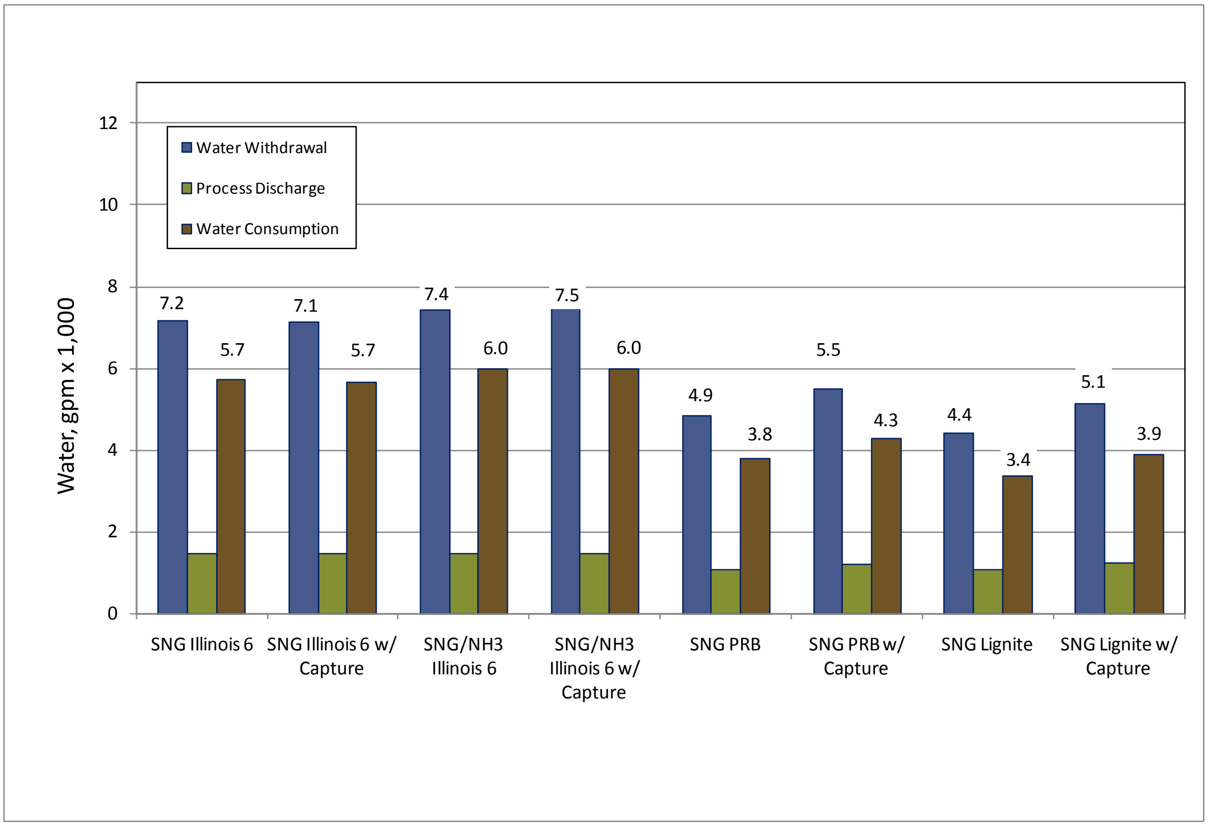

## 6.4 RAW WATER WITHDRAWAL AND CONSUMPTION

Water withdrawal, process discharge, and water consumption are shown in Exhibit 6-9 for each case. The primary conclusions that can be drawn are:

- In all cases, the primary water consumer is cooling tower makeup, which ranges from 60 to 80 percent of the total raw water withdrawal.

- The SNG and ammonia co-production cases (Cases 3 and 4) consume approximately 5 percent more water than the SNG only cases using bituminous coal (Cases 1 and 2). This is largely due to the excess shift steam required to produce the hydrogen necessary for ammonia production.
- For the SNG only non-sequestration cases, the case using lignite coal requires the least amount of water withdrawal followed by the cases using PRB coal and bituminous coal. The relative raw water withdrawal for the SNG only cases is 1.6:1.1:1.0 (bituminous:PRB:lignite). The trend for the SNG only cases with carbon sequestration is 1.4:1.1:1.0 (bituminous:PRB:lignite).

**Exhibit 6-9 Raw Water Withdrawal and Consumption**



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