Appendix C

Task 3 Gasification Alternatives for Industrial Applications

DOE Contract No
DE-AC26-99FT40342
Subtask 3.4 – Lignite-Fueled IGCC Power Plant

Prepared For:
United States Department of Energy / National Energy Technology Laboratory

Nexant

In association with:

gti

May 2005

Task 3 Gasification Alternatives for Industrial Applications

Subtask 3.4
Lignite-Fueled IGCC Power Plant

DOE Contract
No. DE-AC26-99FT40342

May 2005

Prepared For:

United States Department of Energy / National Energy Technology Laboratory

Authors
Nexant: Samuel Tam, Alan Nizamoff, Sheldon Kramer, Scott Olson
GTI: Francis Lau, Mike Roberts, Robert Zabransky
NETL: Jeffrey Hoffmann, Erik Shuster, Nelson Zhan

101 Second Street, 11th Floor, San Francisco, California 94105-3672, USA
Tel: +1 415 369 1000 Fax: +1 415 369 9700

In association with:

Disclaimer

Contents

Figures		Page

Nexant, Inc. completed Tasks 1 and 2 of the *Gasification Plant Cost and Performance Optimization Study* for the U.S. Department of Energy (DOE), the National Energy Technology Laboratory (NETL) in 2003. These tasks focused on the use of the E-GAS™ gasification technology (now owned by ConocoPhillips). NETL has expanded this effort to evaluate *Gasification Alternatives for Industrial Applications.* For this effort the GTI fluidized bed gasifier (U-GAS®) was selected as the basis for the gasification portion for the plant design. This technology is well suited for use on an industrial scale to replace coal-fired boilers and power applications.

This project is defined as Task 3 of the *Gasification Plant Cost and Performance Optimization Study* and focuses on *Gasification Alternative for Industrial Applications*. This task has two basic objectives. The first objective focused on smaller scale industrial systems (here industrial scale is considered to be less than 100 MW) suitable for the coproduction of power and heat which can supplement or replace current on-site utility equipment, increase efficiency, reduce pollution, lower operating costs, and/or improve the steam/power balance of the entire plant. Subtask 3.2 developed a base case design and Subtask 3.3 improved this design further. Subtasks 3.2 and 3.3 did not consider applications for a grass-roots plant, but rather as a retrofit situation that uses part of the existing industrial facility's infrastructure.

The second objective was to examine the application of a GTI fluidized bed gasifier in a larger, grass-roots, stand-alone lignite-fueled integrated gasification combined-cycle (IGCC) power plant. Subtask 3.4 developed a base case design for this scenario. This design consisted of a single oxygen-blown gasification train producing sufficient syngas to fully load a single GE 7FB combustion turbine. The plant will be fueled by North Dakota lignite and will be located at an unspecified generic North Dakota site.

This report describes the work performed on Subtask 3.4, which developed a base case design for a grass roots lignite-fueled IGCC power plant that will produce about 251 MW of export power. This lignite-fueled IGCC power plant uses Gas Technology Institute's U-GAS® fluidized bed gasification process. This report examines a single oxygen-blown gasifier coupled with a single GE 7FB (or similar sized) gas turbine, a single heat recovery steam generator (HRSG), and a single steam turbine to produce power in a stand-alone power plant. Optimization of this case is beyond the scope of the current contract.

BENEFITS OF LIGNITE GASIFICATION

Investigation into wider use of lignite is important since over 25% of the total U.S. coal reserves is comprised of lignite. Lignite primarily is found in the Northern Great Plains (North Dakota, South Dakota and Montana) and along Gulf Coast in Texas and Louisiana. Generally, lignite is located near the surface and is surface mined, giving lignite a cost advantage over other coal types since surface mining is the most efficient and cheapest method of mining coal.

In general, lignite contains a substantial amount of moisture, ranging from 30 to 70%. Consequently, the mined lignite has a low higher heating value, ranging from 6,300 to 8,300 Btu/lb (HHV basis). When compared to bituminous and subbituminous coals, lignite is lower cost on a Btu basis, yet it is less efficient to use lignite in a power plant because significant energy is consumed in drying the fuel. As a result, lignite is generally consumed relatively close to the mine to minimize shipping costs. Opportunities may exist to use lignite as a cost effective feed to IGCC power plants sited in lignite producing regions.

DESIGN RESULTS AND ECONOMICS

At design conditions, the plant consumes 2,558 tpd of moisture-free lignite feed and produces 251 MW of export electric power. It also produces 277 tpd of by-product sulfur. Because the as-received lignite contains 32.2% moisture, the lignite crushing and drying area processes 3,775 tpd of wet lignite. In addition, the plant requires 2,763 gpm of makeup river water and 8.93 MBtu/hr of natural gas. This small amount of natural gas is used in the Claus plant, and since this usage is so small, it is being considered as an operating cost under the catalyst and chemical classification and not as a plant feed.

At design conditions, the plant exports 251 MW of power with a net electrical efficiency of 36.5%. The efficiency is a direct function of the moisture content of the lignite feed. A substantial amount of low-level, waste heat is used to supply the heat that is required for drying the lignite from the as-received moisture content of 32.2% to the 20% moisture content of the gasifier feed. In the gasifier, this residual moisture is vaporized and heated to 1600°F, thereby consuming a substantial amount of heat, most of which is recovered in downstream processing. Table 1.1 summarizes the major input and output streams along with some key operating parameters.

The Nexant developed IGCC Financial Model Version 3.01 was used to obtain the results described in this study. A financial analysis showed that the addition of a second, spare gasification train (from coal feeding up to and including the particulate filters) improves the financial performance of the facility. The payout period for a spare gasifier is 4.8 years. The increased availability gained from the spare gasification train (90.76% versus 85.32% without scheduled maintenance) outweighs the 25 M$ increase in capital costs. The estimated EPC cost of the grass roots facility (including the spare gasification train) is 410.5 M$ (second quarter 2004 dollars), or about 1,635 $/kW of design export power.

The Lignite-Fueled IGCC Power Plant has an expected return on investment (ROI) of 19.4%, with a net present value (NPV) of 175.6 M$ at a 10% discount rate over a 20 year project life. As expected, Subtasks 3.2 and 3.3, the 25 MW subbituminous industrial gasification facilities, have higher installed costs (2,700-3,100 $/kW) because of the economy of scale disadvantage. However, studies of larger IGCC designs (450 MW) have been able to capture even greater economy of scale benefits, with installed costs of 1,300 to 1,650 $/kW[1]. The installed cost of this case is approaching the cost of the large IGCC facilities by taking advantage of a greater economy of scale. The results point to the possibility that a larger design may be able to reduce installed costs further.

Table 1.1 Overall Plant Summary

	Lignite-Fueled IGCC Power Plant
Design Inputs	
Lignite Feed, moisture-free tpd	2,558
Lignite Feed, moisture-free lb/hr	213,160
Fuel (Natural Gas), million Btu/hr	8.93
Makeup Water, gpm	1,920
Design Outputs	
Export Power, MW	251.0
Sulfur, lb/hr	1,557
Ash, lb/hr	23,729
EPC Cost, M$*	410.5
Plant EPC Cost, $/kW	1,635
Plant Energy Input, k$/million Btu/hr	174.7
Plant Energy Output, k$/million Btu/hr	478.2
Cold Gas Efficiency, % (HHV basis)**	84.0
Net Electrical Efficiency, % (HHV basis)***	36.5

* EPC cost is on second quarter 2004 dollars at the North Dakota location. Contingency, taxes, fees, and owners costs are excluded

** Cold Gas Efficiency is defined as the energy in the syngas leaving the gasifier relative to the energy of the feed coal (HHV basis)

*** Net Electrical Efficiency is defined as the export electrical energy from the turbine relative to the energy of the feed coal (HHV basis)

Table 1.2 outlines the rate of return (ROI), NPV, payback year, and required electricity selling price to obtain a 12% ROI with all other entries fixed. The ash and sulfur produced in the plant accounts for all additional revenue beyond electricity tariffs.

Table 1.2 Financial Cost Summary for Lignite-Fueled IGCC Power Plant

ROI (%)	19.4
NPV (M$) (10% Discount Rate)	175.6
Payback Year	2014
Electricity Selling Price for 12% ROI (cents/kWh)	4.7

[1] Analysis of 4 different IGCC technologies without CO_2 capture, "Gasification Process Selection—Tradeoffs and Ironies", EPRI, presented at the Gasification Technologies Conference 2004, October 2004.

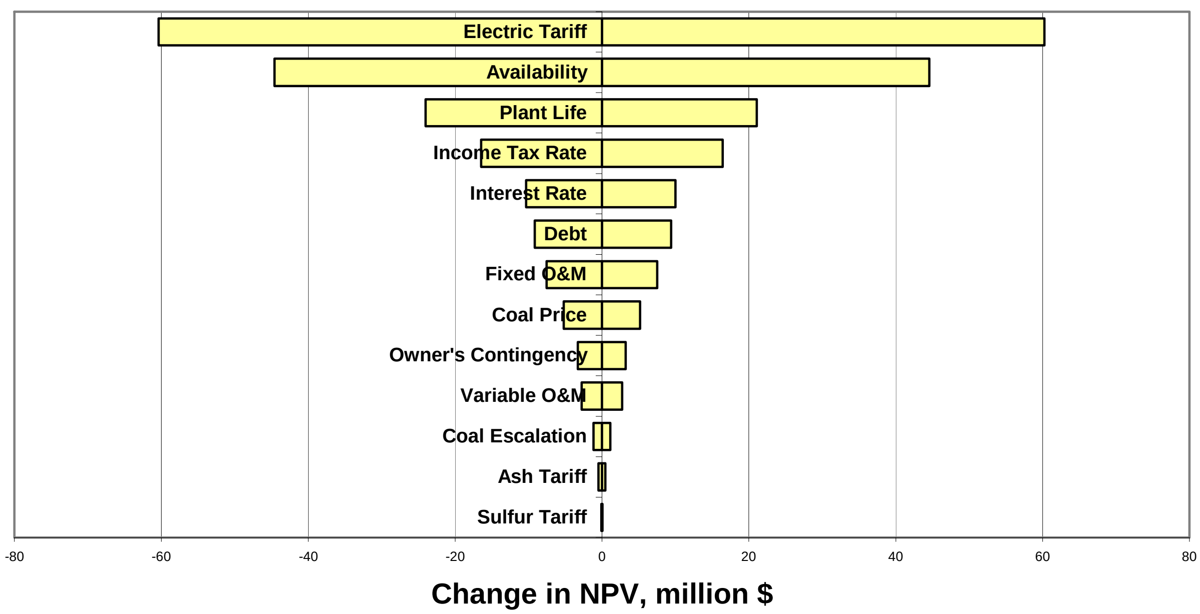
The investment cost accuracy for this study ranges from -15% to +30%, and results in a ROI range of 23.6% to 12.8%.

The results reported in Table 1.2 do not include any credits for the environmental benefits gained from the use of IGCC technology. In order to properly compare this design versus other power generation technologies using lignite, a life cycle analysis also should be performed. Quantification of the environmental differences will provide a more level playing field by which alternate technologies can be evaluated. A project developer must consider alternative compliance costs to meet new emission rules versus the cost of the IGCC plant.

Subtask 3.4 represents a case focused on commercially proven technologies. Future analysis should be performed to consider additional potential cost savings more closely. Section 7 describes potential areas for technology and design improvements that may be able to reduce the total project cost further.

Figure 1.1 shows the impacts of selected variables on the NPV at a discount rate of 10%. In all of the cases, the input parameter is varied by ±10%, and the NPV changes from the base case are shown. 10% changes were used to give a common ground by which all variables were evaluated. However, the range of realistic possibilities for each variable could differ significantly. For example, 10% changes in the availability or income tax rate would capture the majority of long-term variations. This would not be the case with variables such as coal price and electricity tariff that could vary by much more than 10%. The relative significance and range of possible values were considered in determining which variables have the most impact on the model.

Figure 1.1 Comparisons of a +/-10% Change in Selected Inputs on Project NPV
(Discount Rate = 10%)



The electricity tariff has the greatest impact on the plant net present value; increasing it by 10% increases the net present value by more than 60 M$. In this case, “Electricity Tariff” refers to the sales price of the electricity that the plant generates. This variable was also the most significant in Subtasks 3.2 and 3.3. The significance is more pronounced in this design since, unlike Subtasks 3.2 and 3.3, there is no steam export. Also very significant in importance is the availability (annual average on-stream time). By reducing the availability by 10%, the net present value is reduced by more than 45 M$. All other variables associated with the amount of time the plant is operating (e.g., operating hours, and plant life) also have a significant impact on plant economics.

The remainder of the input variables impacted the plant economics to a significantly lower extent. It is interesting to note that the interest rate, amount of debt financing, and the plant fixed O&M cost have a greater impact on the economics than in Subtasks 3.2 and 3.3. This is due in large part to the higher EPC cost of Subtask 3.4. Changes in these variables will impact the early cash flow to a greater extent than in the industrial gasification case. Income tax rate also has a greater impact than in Subtasks 3.2 and 3.3 due to the positive cash flows throughout the operational life of the project. Coal prices could change fairly significantly without changing the overall economics to a great extent. If the coal price is increased to 12 $/ton (a nearly 30% rise), the NPV is only decreased by 15.3 M$, a 0.8 percentage point change in the return on investment.

Figure 1.2 shows the relationship between the electricity tariff and ROI assuming a 10% discount rate. The model relies most heavily on the electricity tariff for the economic outcome due to electricity being the main product. Even with the relatively low electricity prices that exist in North Dakota, the plant still demonstrates positive economics. If the electric price used for upstate New York for Subtasks 3.2 and 3.3 of nearly 8 cents/kW-hr were applied here, the plant would have a return of over 27%. Regardless of the tariff value assumed, any U.S. electricity market could obtain positive returns with this facility, all other plant inputs being equal. The fluctuating marginal prices for electricity from other feedstocks make coal based gasification a competitive option.

Figure 1.2 Effect of Electricity Tariff on Investment Return

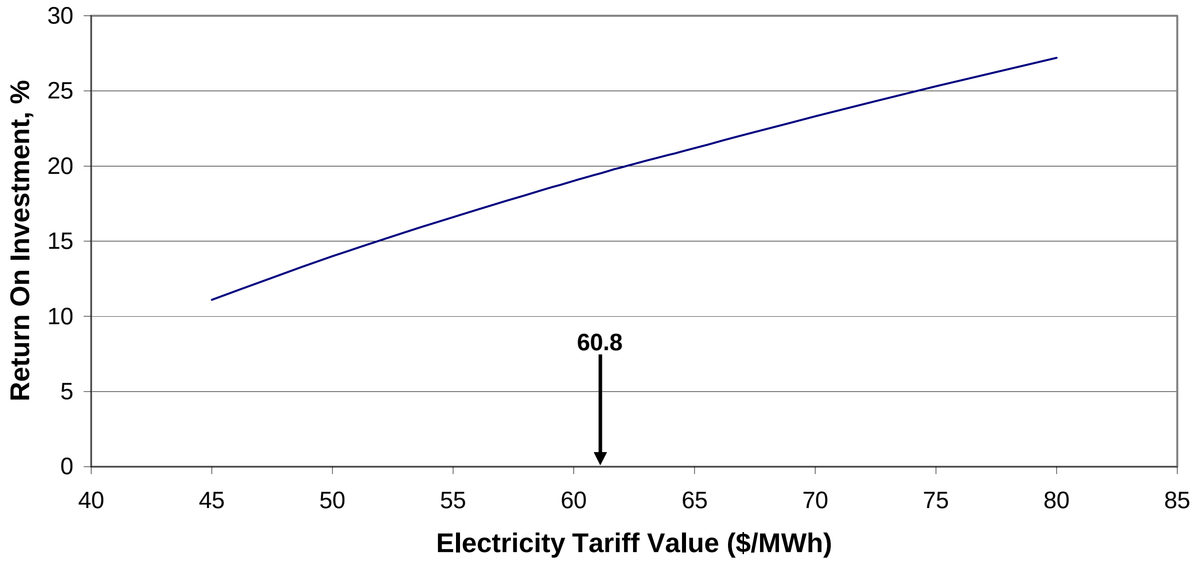
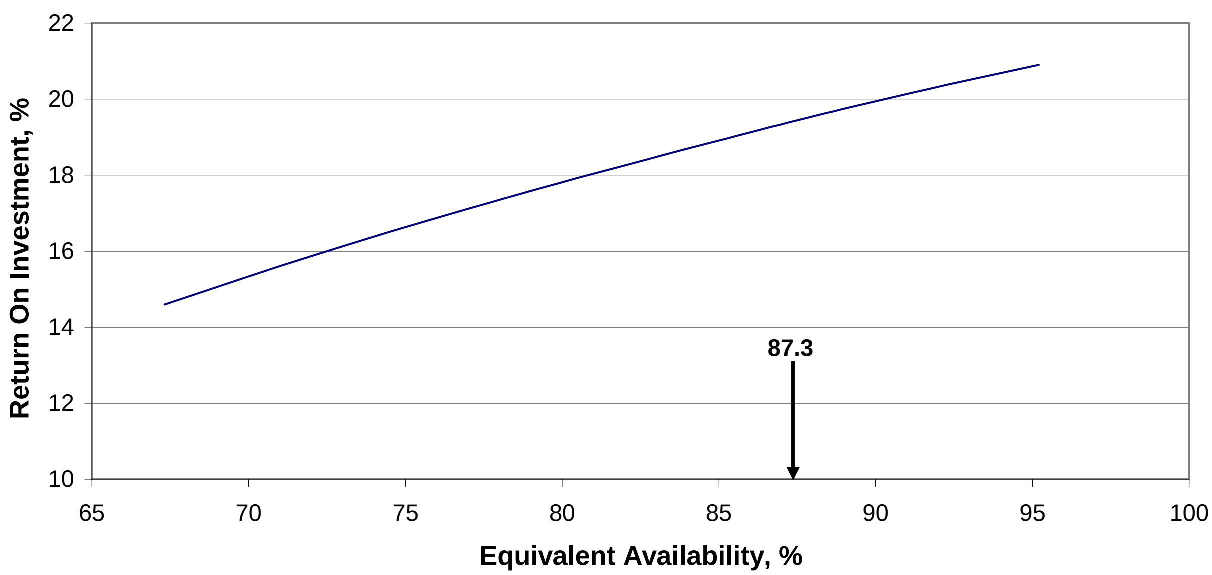


Figure 1.3 shows the relationship that varying the guaranteed availability has on the ROI assuming a 10% discount rate.

Figure 1.3 Effect of Availability on Investment Return

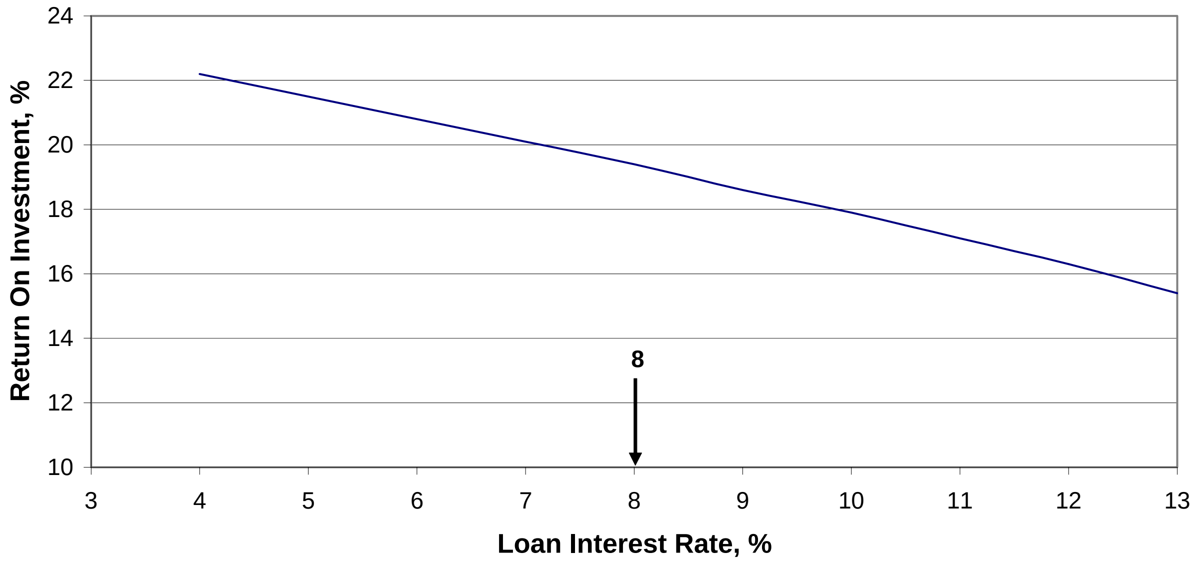


The impact that availability has on the plant economics comes as little surprise. Reliable operation is very important to assure that the cost of project development and construction can be recovered. Long downtimes throughout the life of the project will significantly hurt overall project economics given a 20-year project life. The impact of availability on overall plant economics is similar to that of Subtasks 3.2 and 3.3. Since a plant of this size should be built to have a lifetime longer than the 20 years used in the model, greater consideration of plant life should be made during the project development phases. Figure 1.4 makes this point more clearly. A certain economic life is required in order to pay off the debt incurred during project construction. Once this debt has been paid and construction costs recouped, the steady cash flow will lead to a stable rate of return.

Figure 1.4 Effect of Plant Life on Investment Return

The interest rate for debt financing plays a larger role in this case than in Subtasks 3.2 and 3.3. Monthly interest payments will be significantly higher than in the industrial gasification case. However, interest rate variations do not have a relatively greater significance than either availability or electricity price, as shown in Figure 1.5.

Figure 1.5 Effect of Interest Rate on Investment Return



As with Subtasks 3.2 and 3.3, availability and electricity tariff should receive the most attention when considering the range of financial outcomes. Other parameters, while important to a complete picture of a facility's financial potential, do not have the impact of these two factors. Electricity tariff is especially important in this case due to the lack of other important plant outputs. The increase in capital costs in Subtask 3.4 makes the net plant investment of higher significance than in Subtasks 3.2 and 3.3. The conditions under which the plant is financed become more important with the capital cost increasing by a factor of four.

One key result of the sensitivity analysis is that positive investment returns were found for the entire range of variables that were analyzed. This demonstrates that the model and economics are robust—even with large changes in the financial parameters required to establish a very "conservative" case, plant returns are still positive. The economic results can be stated with confidence that even if changes are made in some of the key financial parameters, the base case still provides a close estimate of economic performance. This range of outputs needs to be reconciled with the risk tolerance of the project developers.

The results of this analysis should not be applied to every facility considering lignite gasification. While these results are valid for the current site and timeframe, others interested in gasification applications must consider their own unique circumstances to develop a proper financial analysis. However, this sensitivity analysis can provide insight into the outcome for plants with somewhat different base assumptions.

CONCLUSIONS AND RECOMMENDATIONS

This study has shown that:

- There are commercially available processes and technologies available for the design of a lignite-fueled IGCC power plant based on the U-GAS® gasification technology that could provide reliable, long-term operation.
- A ROI of 19.4% is achievable at the current average industrial price of electricity in North Dakota. Future optimization of this plant design will identify several enhancements that will further improve the economics of IGCC power plants.
- Results of a sensitivity analysis point to capital investment, availability and electricity tariff as the most sensitive parameters.
- During development of this design a spare gasification train was justified because it increases the overall availability even though it increases the capital investment.
- As a result of this study, a list of potential enhancements (see Section 7) have been identified that should provide additional cost savings as they are researched, developed and implemented, such as.
    - The Stamet solids feeding system
    - Warm sulfur and mercury removal systems
    - Improved particulate removal systems
    - Optimization of the makeup water purification system
    - Combined ash removal systems
    - Improved heat integration
- As a result of this study, a list of R&D needs have been identified including:
    - Further study of lignite drying techniques
    - Investigate the effect that moisture content has on the U-GAS® gasifier operation
    - Update the database for gasification reactivity of the desired coal
    - Further study of the ash characteristics associated with the char
- Based on the simulations prepared for this study the design should meet emission targets established by the DOE in their roadmap for 2010 (see Section 5.3).

Technology development will be the key to the long-term commercialization of gasification technologies and integration of this environmentally superior solid fuel technology into the existing mix of power plants. The following areas are recommended for further development:

- Additional optimization work is required for lignite coals. This may entail changes to the plant configuration to include improved heat integration and sulfur recovery methods.

- Demonstration of the warm gas clean-up technologies so that cooling of the syngas (i.e., below 300°F) can be eliminated, increasing overall efficiency.

- Develop a R&D program that will address the critical issues such as:

    - Developing a technique for drying lignite feedstock that doesn't lead to spontaneous combustion and at the same time increases the overall efficiency.
    - Commercialize the Stamet feed pump at high pressure and high capacity to reduce the cost of the gasifier feed system.
    - Prove the availability of the gasification system and various sub-systems.
    - Determine the combustion turbine performance on the design syngas (both output and emissions) in order to prepare for commercialization.

- Although it is known that reducing the moisture content of the coal feed to the gasifier is more efficient than evaporating the moisture in the gasifier, it has not been established that 20% is the optimum moisture content of the gasifier feed. This needs to be more thoroughly investigated.

- The physical characteristics and properties of lignite must be further studied in order to better predict the gasification systems. These include:

    - Determination of the gasification reactivity of the desired coal.
    - Determine the ash characteristics associated with the char
    - Characterize the particulate properties
    - Characterize hydrocarbon content of the syngas to confirm the design of the sour water stripper and the effluent water treatment facilities

- Determination of the cyclone performance at higher temperatures.

    - During a visit to a gasification facility in China it was noted that at temperatures above 1000°F the cyclone efficiency drops off sharply. This was confirmed by Emtrol, a world leader in cyclone design.

Nexant, Inc. completed Tasks 1 and 2 of the *Gasification Plant Cost and Performance Optimization Study* for the U.S. Department of Energy (DOE), the National Energy Technology Laboratory (NETL) in 2003. These tasks focused on the use of the E-GAS™ gasification technology (now owned by ConocoPhillips). NETL has expanded this effort to evaluate *Gasification Alternatives for Industrial Applications.* For this effort the GTI fluidized bed gasifier was selected as the basis for the gasification portion for the plant design. This technology is well suited for use on an industrial scale to replace coal-fired boilers and power applications.

This project is defined as Task 3 of the *Gasification Plant Cost and Performance Optimization Study* and focuses on Gasification Alternative for Industrial Applications. Subtasks 3.2 and 3.3 focused on smaller scale industrial systems suitable for the coproduction of power and heat which can supplement or replace current on-site utility equipment, increase efficiency, reduce pollution, lower operating costs, and/or improve the steam/power balance of the entire plant. Subtasks 3.2 and 3.3 did not consider applications for grass-roots plants or larger power plants, but rather as a retrofit situation that uses part of the existing industrial facility's infrastructure.

Subtask 3.4 considered a larger, grass-roots, stand-alone power plant consisting of an oxygen-blown gasification train producing sufficient syngas to fully load a single GE 7FB combustion turbine. This plant uses low-level waste heat to dry the lignite that otherwise would be rejected to the atmosphere. The use of oxygen allows the potential for capture and sequestration of CO_2 from gasification systems. The plant is fueled by North Dakota lignite and will be located at an unspecified generic North Dakota site.

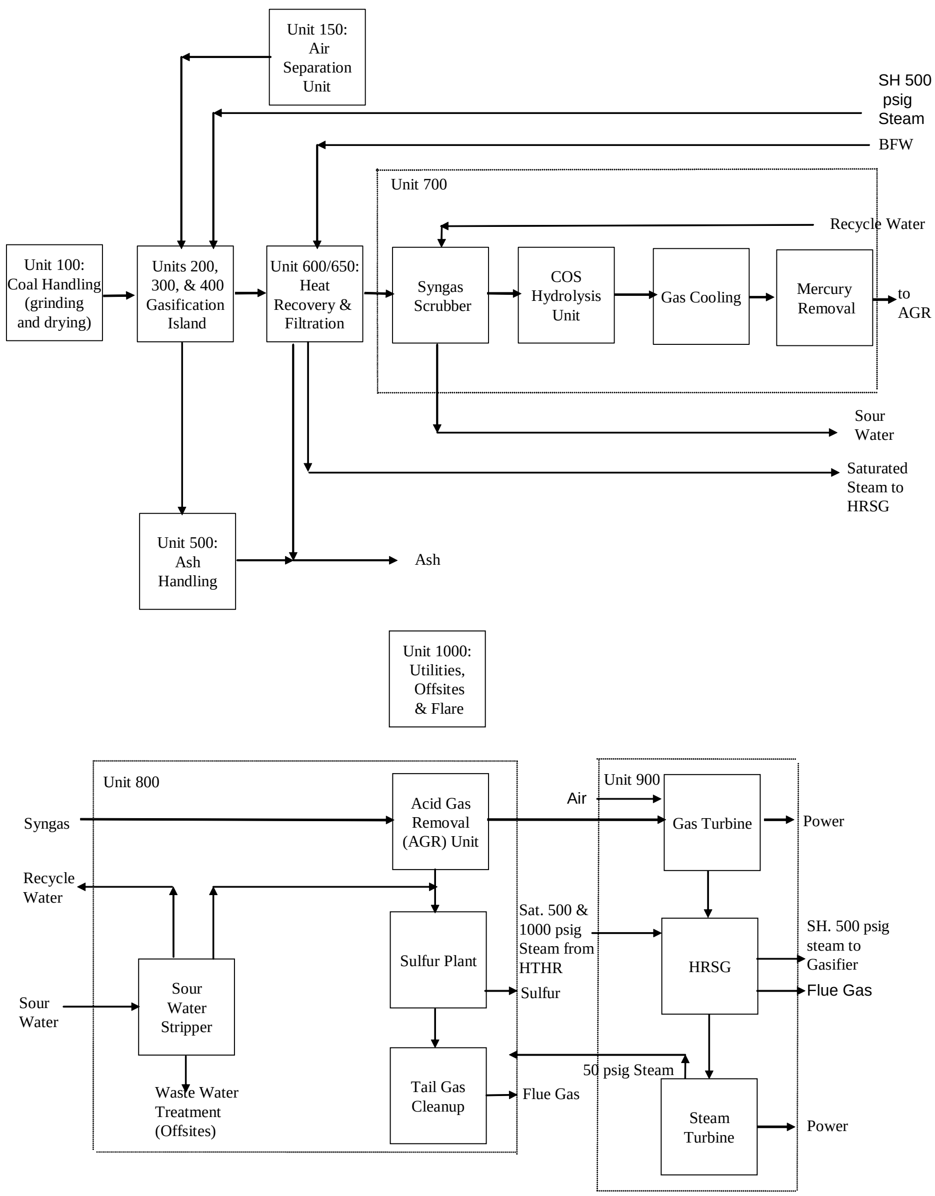
This task has two specific objectives. The first objective was to examine the application of a GTI fluidized bed U-GAS® gasifier at an industrial application in upstate New York using a Southeastern Ohio coal (here industrial scale is considered to be less than 100 MW). Subtask 3.2 developed a base case design and Subtask 3.3 improved this design further. (Subtask 3.1 covers management activities.) The second objective was to examine the application of a GTI fluidized bed U-GAS® gasifier at a stand-alone lignite-fueled IGCC power plant in North Dakota. Subtask 3.4 developed a base case design for this scenario. This report describes the work performed during Subtask 3.4.

Figure 2.1 is a simplified block flow diagram of the facility. The inside battery limits (ISBL) plant contains one air separation unit (oxygen production facility) that feeds a single GTI U-GAS® gasifier. The syngas leaving the gasifier is cooled in the high temperature heat recovery area. Any remaining particles are removed in the metallic filters. The syngas goes to the low temperature heat removal area. The cooled syngas is sent to a cleanup section that consists of a water scrubber, sulfur recovery, and mercury removal. The cleaned syngas then goes to the power block that consists of a single GE 7FB combustion turbine with a dedicated heat recovery steam generator (HRSG), and a single steam turbine.

This plant is a stand-alone facility that will be located at an unspecified generic site in North Dakota with the assumption that it a level and cleared site that will be accessible to other utilities, such as potable water, sanitary sewer, etc. River water (for process use) and electrical transmission facilities shall be available near the facility.

The determination of the exact lignite feed rate was a part of this study. The lignite feed rate was selected so that the plant will produce the required amount of syngas to fully load one GE 7FB combustion turbine (CT).

Figure 2.1 Simplified Block Flow Diagram



3.1 STUDY OBJECTIVES

The primary objective of Task 3 is to investigate Gasification Alternatives for Industrial Applications. This is the third of the three topical reports defined as subtasks under Task 3 of this DOE contract (DOE contract no. DE-AC26-99FT40342). The first topical report presented the capital and operating costs for a preliminary design of an industrial-size, IGCC coal-fired gasification project. An existing industrial site that is considering replacement of outdated steam boilers is used as the site model. IGCC will reduce emissions, increase efficiency, and reduce operating costs at the facility. The use of combined heat and power (CHP) at industrial facilities using coal can contribute to a significant increase in distributed generation (DG) for improving local power grid security. The third topical report discussed an alternate air-blown design for this case.

In Subtask 3.4 (third topical report) a stand-alone power plant configuration consisting of a single oxygen-blown GTI fluidized bed U-GAS® gasifier coupled with a single General Electric 7FB (or similar sized) gas turbine, heat recovery steam generator (HRSG), and a steam turbine was investigated. This plant uses low-level waste heat to dry the lignite that otherwise would be rejected to the atmosphere. The use of oxygen allows the potential for capture and sequestration of CO_2 from gasification systems. The design for this evaluation includes equipment and technology that are commercially available to allow construction at the present time. It is believed that a concentrated stream of CO_2 will make CO_2 capture and sequestration more economic than that for a stream diluted with nitrogen as in an air-blown design. The plant will be fueled by North Dakota lignite and will be located at a generic North Dakota site.

Several NETL employees are working on this project directly with Nexant personnel. They are assisting with the execution of this project. As an outcome of this participation, NETL will develop and enhance its systems analysis expertise from the initial stage of developing the strategy for an appropriate level systems study, through the analysis of technical and economic feasibility, performing sensitivity analyses, and finally, the presentation of results.

Specifically, the NETL employees participating in this activity have been directly involved in or exposed to the following tasks:

- Participated in strategy meetings and brainstorming sessions to enhance their "systems perspective"
- Developed an appropriate approach commensurate to the level of results needed
- Used spreadsheets, ASPEN, and other software models to analyze a system or concept
- Determined economic and technical feasibility (developing cost estimates, project financing, mass and energy balances, etc.)

- Conducted sensitivity analyses to determine the primary variables that affect cost and/or performance
- Evaluated trade-offs for optimization
- Developed summary tables, flowcharts, written documentation, and presentation materials that effectively report the project objective, approach, and results

3.2 BACKGROUND

In late 1999, the National Energy Technology Laboratory awarded Nexant Inc. (a Bechtel Technology & Consulting Company) and Global Energy, Inc. (which acquired the gasification related assets of Dynegy Inc., of Houston, Texas including the E-GAS™ gasification technology, formerly the Destec Gasification Process) a contract to optimize IGCC plant performance.[1] During the performance of this contract, ConocoPhillips purchased the E-GASTM gasification technology. This contract was divided into three tasks. Task 1 developed two optimized IGCC plant configurations: (1) petroleum coke gasification for electric power with the coproduction of hydrogen and industrial-grade steam, and (2) coal gasification for electric power generation or hydrogen production. Task 2 developed two different optimized IGCC plant configurations: (1) petroleum coke gasification for electric power with the co-production of liquid transportation fuel precursors, and (2) coal gasification for electric power with the co-production of liquid transportation fuel precursors. In September 2003, a Final Report [for Tasks 1 and 2] was published.[2] The Tasks 1 and 2 Topical Reports are an integral part of this report.[3,4]

In late 2003, the contract was modified to add a new task (Task 3) to consider "Gasification Alternatives for Industrial Applications."[5] This task was designed to develop smaller gasification plants for industrial applications using Gas Technology Institute's (GTI's) U-GAS® fluidized bed gasifier. Task 3 is divided into three technical subtasks. Subtask 3.2 investigated a brownfield design modeled after the requirements of a specific industrial site in upper New York State that co-produces both power and steam.[6] Both air and oxygen-blown gasification systems were considered. Subtask 3.3 will developed an alternate plant design based on the air-blown configuration from Subtask 3.2. Subtask 3.4 developed a design of a nominal 251 MW power plant using North Dakota lignite.

---

1 Contract No. DE-AC26-99FT40342, "Gasification Plant Cost and Performance Optimization"

2 "Final Report – [Tasks 1 and 2]" Gasification Plant Cost and Performance Optimization, United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, September 2003.

3 "Topical Report – Task 1 Topical Report, IGCC Plant Cost Optimization," Gasification Plant Cost and Performance Optimization, United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, May 2002, http://www.netl.doe.gov/coalpower/gasification/projects/systems/docs/40342R01.PDF.

4 "Topical Report – Task 2 Topical Report, Coke/Coal Gasification With Liquids Coproduction," Gasification Plant Cost and Performance Optimization, United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, September 2003.

5 Contract modification November 21, 2003

6 "Topical Report – Task 3: Gasification Alternatives for Industrial Applications, Subtask 3.2: Preliminary Design for Eastern Coal Case," Gasification Plant Cost and Performance Optimization, United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, November 2004.

This document is the Topical Report for Subtask 3.4.

3.3 METHODOLOGY

Task 3, *Gasification Alternatives for Industrial Applications*, shifts the focus from the scope of the study in Tasks 1 and 2. The objective of Subtasks 3.2 and 3.3 examined smaller scale systems suitable for the coproduction of power and heat which can supplement or replace current on-site utility equipment, increase efficiency, reduce pollution, lower operating costs, and/or improve the steam/power balance of the entire plant. Subtasks 3.2 and 3.3 did not consider applications for a grass-roots plant, but rather as a retrofit situation that uses part of the existing industrial facility's infrastructure.

Subtask 3.4 considered a grass-roots, stand-alone power plant consisting of a single oxygen-blown gasification train producing sufficient syngas to fully load a single GE 7FB combustion turbine. The plant is fueled by North Dakota lignite and will be located at a generic North Dakota site.

The U-GAS® gasification technology system developed by the Gas Technology Institute was the basis for this study. This system is based on a non-slagging, fluidized bed gasifier. The total of knowledge gained from previous GTI gasifier designs using this technology on coal (e.g. Shanghai Coking and Chemical Corp.) has been studied to compile relevant information for this project.

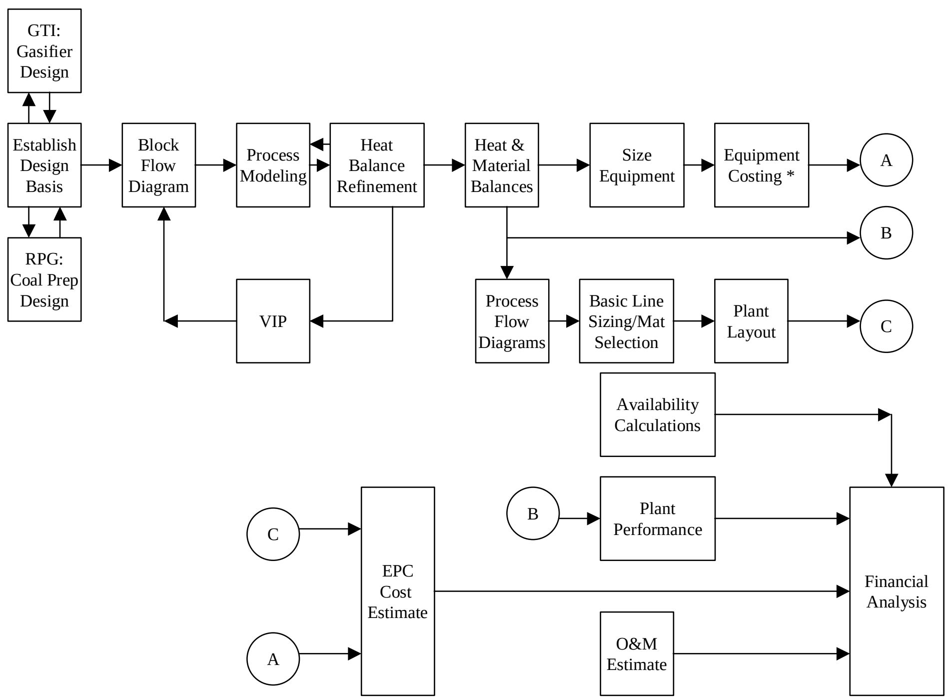
Figure 3.1 is a schematic diagram of the steps involved in developing the design, cost and economics for a specific case. Addendum F contains the technical work plans for the Subtask 3.4 design basis. Based on these design bases, input from GTI's gasifier model and input from various vendors, an elementally balanced process simulation model of the gasifier was developed using Aspen Plus®, a commercially available process simulation program. This is a very detailed process simulation program that simulates the various heat exchange and steam generation steps within the gasification area. Process simulations were also developed for the syngas cooling and cleanup portion of the plant and the sour water stripper. The resulting heat and material balances provided the feed to the GateCycle simulation program for a detailed simulation of the power block. This report and its appendices contain sufficient information for verification of the carbon, slag, sulfur, and heat balances.

Process flow diagrams (PFDs), sized equipment lists, line sizings, and other information necessary to calculate the plant costs were developed based on the model results. The mid-year 2004 plant cost was built up based on prior cost information selected equipment quotes, information from similar Nexant projects, and from commercially available cost estimating software.

In addition the study for the Subtask 3.4 plants was done employing the structured Value Improving Practices (VIP) Program promoted by Independent Project Analysis, Inc.

Availability analyses were calculated based on the design configuration to determine the annual production rates (capacity factors). The cost and capacity information along with operating and maintenance costs, contingencies, feed and product prices, and other pertinent economic data were entered in a discounted cash flow economic model. This model then was used to generate the return on investment (ROI), net present value (NPV), and sensitivities.

Figure 3.1 Task Development Methodology



* All critical process equipment costs in Gasifier Train, gas turbine, HRSG & ASU are derived from budgetary quotes

In some cases, such as in the development of the spare gasification train cases, iterations were made back to the to the block flow diagrams to examine the effects of replicated equipment and the addition of a spare gasification train.

The initial basis for the Subtask 3.4 design required that the coal be dried to 10% moisture content and used two parallel gasification and gas turbine trains. These preliminary results showed a relatively low net electrical efficiency. The basis was reconsidered with items critical to increasing the efficiency and improving the design reexamined. The material and energy (M&E) balances were recalculated using GateCycle and ASPEN except for the sour water stripper, which was simply scaled from the initial design. Utilities that were not developed using the simulation tools were scaled based on throughput. Cost estimates were prorated based on throughput from the initial design using an exponent of 0.65 for all equipment.

3.4 AVAILABILITY ANALYSIS

The common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period, all are dependent on the project cash flow. The net cash flow is the sum of all project revenues and expenses. Depending upon the detail of the financial analysis, the cash flow streams usually are computed on either an annual or quarterly bases. For most projects, the net cash flow is negative in the early years during construction and only turns positive when the project starts generating revenues by producing saleable products. However, a plant is generating revenue only when it is operating and not when it is shut down for forced outages, scheduled maintenance, or repairs. Therefore, the yearly production (total annual production) is a key parameter in determining the financial performance of a project.

Although the design capacity is the major factor influencing the annual production, other factors including scheduled maintenance, forced outages, equipment reliability, and redundancy influence it. To develop a meaningful financial analysis, an availability analysis that considers all of the above factors must be performed to predict the annual production and revenue streams. On this basis, availability analyses were performed to determine the applicable revenue streams and the ROI.

3.4.1 Availability Analysis Basis

In Table 5.0A of the Final Report for the Wabash River Repowering Project, Global Energy reported downtime and an availability analysis of each plant system for the final year of the Demonstration Period.[7] During this March 1, 1998 through February 28, 1999 period, the plant was operating on coal for 62.37% of the time. There were three scheduled outages for 11.67% of the time (three periods totaling 42 days), and non-scheduled outages accounted for the remaining 25.96% of the time (95 days). After some adjustments, the EPRI recommended procedure was used to calculate availability estimates for each case.[8]

Recent data presented at the 2002 Gasification Technologies Council conference by Clifton Keeler show further reliability improvements in the on-stream performance of the Wabash River Repowering Project.[9] However, the following availability and financial analyses are based on the data reported in the final repowering project report. Thus, the following financial analysis is somewhat conservative.

The objective of this availability study is to determine the projected annual revenue stream. With respect to the annual revenue stream, it is immaterial whether the plant is off line because of a forced outage as the result of an equipment malfunction or whether it is off line because of a scheduled outage for normal maintenance or refractory

7 "Wabash River Coal Gasification Repowering Project, Final Technical Report," U. S. Department of Energy, Contract Agreement DE-FC21-92MC29310, http://www.lanl.gov/projects/cctc/resources/pdfs/wabsh/Final%20_Report.pdf, August 2000.

8 Research Report AP-4216, Availability Analysis handbook for Coal gasification and Combustion Turbine-based Power Systems, Research Project 1800-1, Electric Power Research Institute, 3412 Hillview Avenue, Palo Alto, CA 94304, August 1985.

9 Clifton G. Keeler, Operating Experience at the Wabash River Repowering Project, 2002 Gasification Technologies Council Conference, San Francisco, CA, October 28, 2002.

replacement. This study calculated annual expected plant availabilities for the period that the facility is scheduled to be operating. These values were then adjusted for the scheduled maintenance outages to determine the annual feed and product rates.

This study also assumes a mature facility, as compared to a first-of-a-kind (FOAK) plant that is subject to lower availabilities in its early operational history.

3.4.2 Use of Natural Gas

In certain situations, sufficient amounts of syngas may not be available to fully load all available gas turbines. Under these conditions an auxiliary fuel may be used to supplement the available syngas to fire the combustion turbine to maximize power production. Natural gas is preferred for these applications. However, it was assumed not to be available for this study. When this situation occurs, the power output from the turbines is reduced and the internal power consumption is reduced. The net effect of this combination of events is that there is a reduction in export power.

The decision of whether or not to use backup natural gas to supplement power production should be a "real time" decision that considers the relative prices of natural gas and power, expected length of the syngas shortage and power demand. For example, the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction uses purchased power to maintain Fischer-Tropsch liquids production during periods when the combustion turbine is unavailable.

Although natural gas is used in most of the Task 1 and Task 2 subtasks to increase production when sufficient syngas is not available, no natural gas is used for this purpose in Subtask 3.4. However, some natural gas is used during startup for heat up, refractory conditioning, etc. This gas usage is considered to be an O&M cost and not a feedstock cost.

3.5 COMMODITY PRICING

The initial basis for the commodity prices into the gasification model came from information provided by U.S. government agencies. This includes data from the DOE's Energy Information Administration (EIA) Annual Energy Outlook 2004[10] for commercial electricity values and coal, and from the U.S. Geological Survey (USGS) for sulfur. The gasifier bottoms value was estimated using previous values for Nexant gasification studies. Each value was normalized where necessary to reflect the current nominal value, using a 3% inflation rate. The preliminary financial model runs were made using these inputs. Table 3.1 below lists the major assumptions for commodity prices. The financial sensitivities (Section 6) show that the price of the ash and sulfur by-products has almost no influence on the plant economics, but are included here for completeness.

[10] U. S. Department of Energy, Energy Information Administration, "Annual Energy Outlook 2004 with Projections to 2025", January 2004, www.eia.doe.gov/oiaf/aeo.

Table 3.1 Basic Economic Parameters

Feeds	Price	Escalation (%/yr)
Lignite (32.24% moisture)	9.29 $/short ton	2.0
Products		
Electric Power	6.08 cents/kW-hr	3.0
Sulfur	26.52 $/short ton	3.0
Ash	10.00 $/short ton	3.0

The prices for coal and electric power were based on 2002 EIA reported values for North Dakota. Each value was normalized to reflect expected 2005 prices. The coal price is for wet lignite delivered to the gasification facility. Sulfur and ash are from the EIA and USGS estimates used in Subtask 3.2.

To stay consistent with the financial analysis performed in Subtask 3.2, the escalation and other financial model entries outside of coal and electricity prices were kept the same. All escalation rates used for this study were maintained throughout the life of the gasification facility.

In the electricity market, the EIA has predicted a slight decrease in real electricity prices through 2011, then a slight increase through 2025. The net impact for the timeframe of this project is for electricity prices to escalate with the overall rate of inflation. Therefore, the inflation factor used by the EIA, 3%, also was used for the electricity price. It is expected that North Dakota will follow a similar trend in price escalation.

In keeping with previous Nexant studies and expectations of oversupply in the coal industry, the escalation rate for the lignite price was kept to 2%, below the 3% that is expected for future general inflation rates. This falls between current EIA estimates and escalation factors used in previous Nexant studies. While there may be an additional downside to coal prices as some in the industry have suggested, this study wanted to stay away from significant speculation by keeping it close to government predictions and previously published technical reports. It is unclear if lignite will have an escalation rate that is different from the estimate used for coal in Subtask 3.2. EIA estimates show Western coal with a price trend similar to that of general coal prices. For this reason and for consistency with Subtask 3.2, the lignite price escalation rate was kept at 2%.

The ash product (gasifier bottoms and fly ash) can be used for cement and asphalt production. Using previous studies as a basis, it was assumed that this product could be sold for 10 $/ton. This assumption will be tested in the sensitivity analysis due to the volatile nature of this price.

3.6 FINANCIAL ANALYSIS METHODOLOGY

The results reported for rate of return and discounted cash flow come from the NETL IGCC Financial Model Version 3.01 developed by Nexant. This version of the model was developed in May 2002 specifically for NETL under a task order from NETL on-site support contractor E^2S. The model has been used in previous gasification studies, and has undergone critical scrutiny by NETL and other technical experts. It is a robust discounted cash flow model that takes into account all major financial and scenario assumptions in developing the key economic outputs.

In order to develop the appropriate financial assumptions for the facility under consideration, a number of sources were reviewed and conversations were held with team experts. For the most part, the values used were kept the same as those in Subtask 3.2, with the exception of scheduled downtime and construction period. The main sources used as the input basis were NETL's "Quality Guidelines for Energy System Studies", an industry study analyzing the potential for gasification in the U.S. refining market, and previous gasification optimization studies performed by Nexant, namely Tasks 1 and 2 of the current DOE contract. Details of the financial assumptions made can be found in Addendum C. A few of the major assumptions and some of the areas that will be explored via sensitivity analysis are listed below:

- + 30/-15% accuracy assumed for this phase of the analysis
- A 15% project contingency applied across the entire plant with the exception of the gasifier block. For the gasifier block, a separate 25% process contingency was used to reflect the higher uncertainty in this unit's cost estimate.
- Scheduled downtime for 14 calendar days due to gasifier requirements. This is lower than that used for Subtask 3.2 because of the inclusion of a spare gasification train.
- 8% cost of capital
- Total operation and maintenance (O&M) costs of 5% of EPC costs per year (fixed and variable)
- 42-month construction period. This is longer than Subtask 3.2 due to the size of the facility.
- 20 year plant life
- Fees were added to the EPC costs to capture project development, start-up, licensing/permitting, spares, training, construction management, commissioning, transportation, and owner's costs.

Specific plant performance and operating data were entered into the financial model from the design basis. The material and energy balance provided by GTI and verified by Nexant/DOE, along with the subsequent design work by Nexant and NETL, set the entries for items such as power output, sulfur production, and quantities of ash produced. The plant EPC cost used for the model analysis was determined by establishing installed cost estimates for all major unit operations, off-sites, and balance-of-plant items. The basis for installed costs came from a combination of GTI input for the gasifier block, vendor quotes for major unit operations, process design software, and team expertise for the remaining pieces of equipment. A more rigorous explanation of how these numbers were developed is given in Section 6.2 (Plant Costs). Appropriate scale-up factors used in previous gasification projects allowed any additional equipment to be properly estimated.

4.1 STUDY BASIS

This study investigated the cost for installation and operation of a stand-alone integrated gasification combined cycle (IGCC) power plant that will use North Dakota lignite as fuel. The goal of the study was to develop a base case design for the facility and to identify potential improvements for reducing operating costs and lowering plant emissions associated with power generation. This plant will be located at a generic mine-mouth North Dakota site.

The Design Criteria for Subtask 3.4 are highlighted below.

- The fuel will be North Dakota lignite with the following properties:

Table 4.1 North Dakota Lignite Properties

	Proximate Analysis, wt%		**Ash Analysis**	**Wt%**
	As-Received	Dry	SiO_2	18.4
Moisture	32.24	0	Al_2O_3	10.22
Ash	6.59	9.72	Fe_2O_3	8.0
Volatile Matter	30.45	44.94	TiO_2	0.48
Sulfur	0.54	0.80	CaO	24.72
Fixed Carbon	30.18	44.54	MgO	7.48
HHV, kJ/kg	17,338	25,588	Na_2O	7.76
HHV, Btu/lb	7,454	11,001	K_2O	0.94
LHV, kJ/kg	15,894	24,625	SO_3	17.55
LHV, Btu/lb	6,833	10,587	P_2O_5	0.48
			BaO	0.84
	Ultimate Analysis, wt%		MnO	0.14
	As-Received	Dry	SrO	1.12
Moisture	32.24	0	Total	98.13
Carbon	44.62	65.85		
Hydrogen	2.95	4.36	Oxidizing Ash	
Nitrogen	0.70	1.04	Fusion Temperatures, F	
Chlorine	0.03	0.04	IT	2329
Sulfur	0.54	0.80	ST	2393
Ash	6.59	9.72	HT	2425
Oxygen	12.32	18.19	FT	2460
	Sulfur Analysis, wt%		Reducing Ash	
	As-Received	Dry	Fusion Temperatures, F	
Pyritic	---	0.14	IT	2246
Sulfate	---	0.03	ST	2310
Organic	---	0.63	HT	2349
			FT	2394

The mercury content of the dry lignite is 0.14 ppm by weight.

Deposits of lignite are plentiful in the Dakotas region. The lignite is low priced because it contains a high level of moisture, and for that reason, it is not shipped very far from the source. North Dakota lignite was selected for this study because a fluidized bed gasifier appears to be able to handle it very well.

There is an energy penalty (and therefore reduced power output) for drying the high moisture lignite to the low moisture content necessary for reliable feeding via lockhoppers and pneumatic conveying. In order to provide this reliability, the feed system employed herein is 100% spared. Drying of the lignite to very low levels of moisture (e.g., 10%) requires a significant amount of energy and has a detrimental affect on the overall plant efficiency. GTI has specified that the lignite be dried to 20% moisture. However, this is not necessarily an optimum moisture level.

- 2 GTI U-GAS® fluidized bed gasifiers (1 operating and a spare)
- 1 GE 7FB gas turbine @ ~210 MW
- 1 steam turbine @ ~90 MW

Performance will be based on the DOE target emission and performance goals established in their roadmap for 2010. These targets are:

- Sulfur > 99% removal
- NOx < 0.05 lb/MBtu
- Particulates < 0.005 lb/MBtu
- Mercury > 90% removal
- Net Electrical Efficiency = 45-50%
- Capacity factor = 85%

4.1.1 Plant Description

The stand-alone U-GAS® IGCC power plant consists of the following process blocks and subsystems:

- Unit 100: Coal Preparation – Handling, Sizing and Drying
- Unit 150: Air Separation Unit
- Unit 200: Solids Feeding System
- Unit 300: Gasification
- Unit 400: Fines Separation

- Unit 500: Ash Handling
- Unit 600: High Temperature Heat Recovery
- Unit 650: Particulate Removal
- Unit 700: Syngas Scrubber, COS Hydrolysis Reactor, Low Temperature Heat Recovery and Mercury Removal
- Unit 800: Sulfur Removal and Recovery, Sour Water Stripper (SWS)
- Unit 900: Power Block including a single General Electric 7FB combustion turbine (GT) with a heat recovery steam generator (HSRG), and a single steam turbine.
- Unit 1000: Utilities (e.g., instrument and plant air, cooling water systems, firewater system) and other offsites (e.g., flare, DCS, plant roads, buildings, chemical storage)

Figure 4.1 is a block flow diagram of the plant.

Figure 4.1 Block Flow Diagram
Part 1- Syngas Generation and Processing

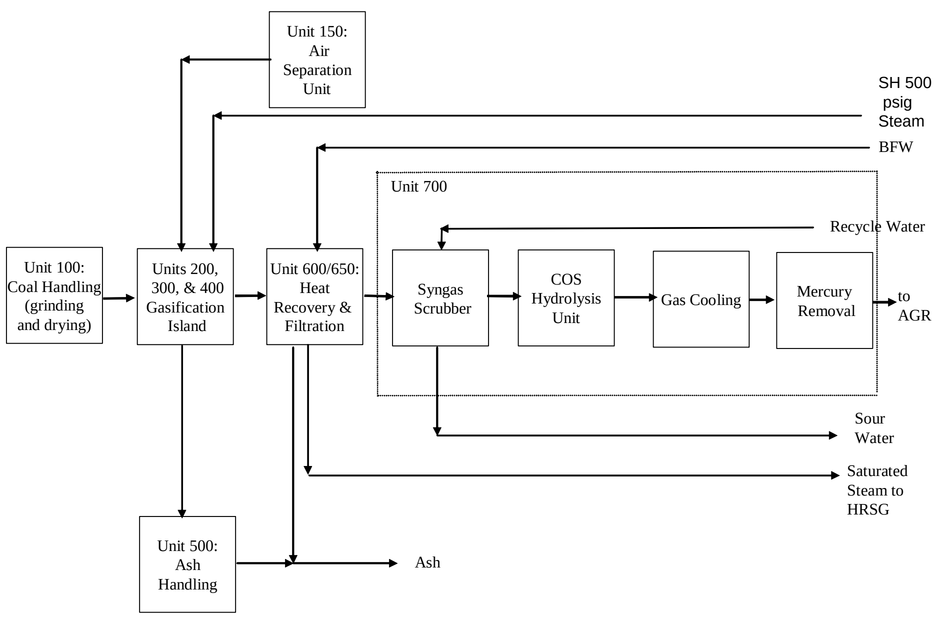
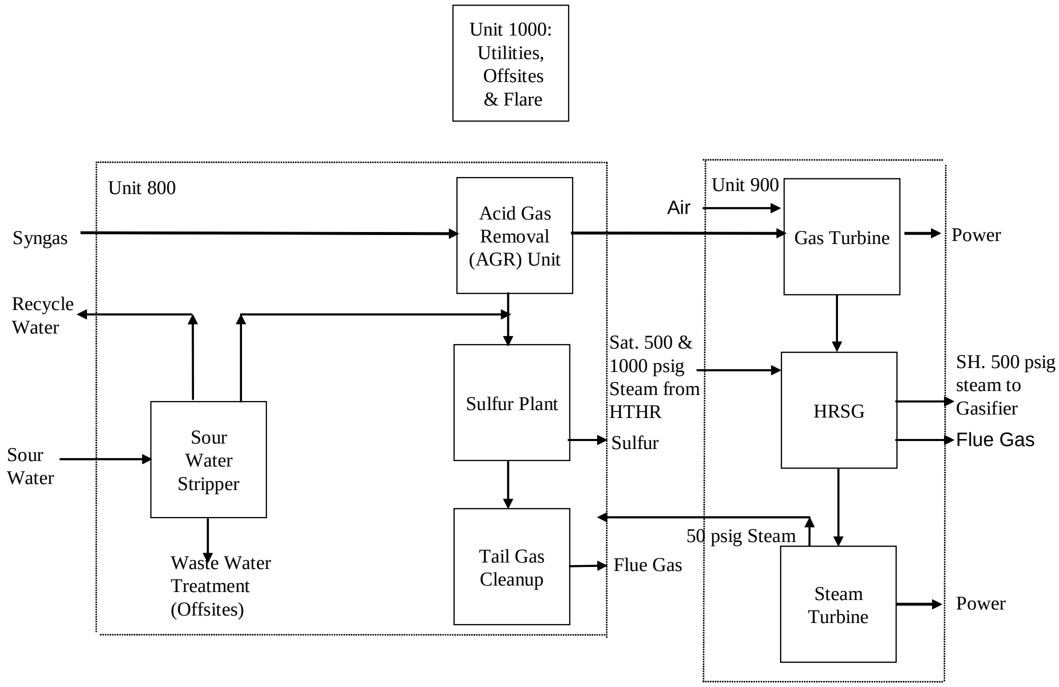


Figure 4.1 (continued) Block Flow Diagram
Part 2 – Sulfur Removal, Sulfur Recovery, SWS, and Power Block



4.1.2 Site Selection

The plant will be situated at a generic mine-mouth location in North Dakota. The site shall be assumed to be level and cleared, and it shall be of sufficient size. Normal infrastructure (electric power, potable water, sanitary sewer, etc.) shall be available at the plant boundary. River water shall be available. Table 4.2 shows the site conditions.

Table 4.2 Plant Site Conditions

Elevation, feet		1,900
Air Temperature, °F		
	Maximum	110
	Average Monthly Maximum	83
	Annual Average	40.5
	Minimum	-50
Relative Humidity, %		
	Maximum	84
	Minimum	36
	Normal	74
Maximum Hourly Rain Fall, in		2.5
Seismic Zone		0
Wind Speed, mph		
	Mean / Maximum	10 / 70

4.1.3 Design Considerations Specific to Lignite (i.e., Spontaneous Combustion)

Lignite (or brown coal) is a class of coals that has a HHV of less than 8,300 but greater than 6,300 Btu/lb on an as-mined basis. Lignite constitutes more than 25% of the U. S. coal reserves. Most of the U.S. low rank coals are concentrated in the Northern Great Plains, in states such as North Dakota, Montana, and Wyoming. Smaller deposits exist in the Rocky Mountain States as well as the Pacific and Gulf Coast States.

In addition to the high moisture content (30 to 70%), lignite typically has a high oxygen content, generally over 20% on a devolatized ash-free basis. It also has a relatively high volatiles content. Substantial quantities of lignite and brown coals occur near the surface in many parts of the world where mining them is relatively easy. It is generally uneconomical to transport lignite over long distances because of its high moisture content and low specific energy content. Thus, lignite is mainly used for mine-mouth power generation or at locations that are reasonably close to the mine. Due to its large surface area and rich volatiles content, precautions must be taken to prevent spontaneous combustion from partially dried lignite.

Coals consist of pore systems intermeshing with the continuous coal structure. The younger coals, such as lignite, where the coalification process is not fully completed can contain a substantial amount of moisture. Most of this moisture is present in the pore structure in contrast to bituminous coals where most of the moisture is present on the surface. Surface moisture and moisture which is contained in the macropores is relatively easy to remove simply by heating. However, the removal of moisture trapped in the micropores tends to be more difficult. It is reported that a portion of the water forms stronger bonding with the coal than between water molecules.[1]

Lignite is a complex system of condensed aromatic rings to which various functional groups are attached. Most of the groups contain oxygen (carboxyl, hydroxyl, methoxyl, ether, and carbonyl).[2] During drying, depending upon the temperature reached, the functional groups, notably the carboxyl groups, are reduced. Thus, to an extent, moisture removal is related to coalification and causes changes in both the chemical structure and surface properties of the material.

Dried lignite has a tendency to undergo spontaneous combustion. The key parameters that influence spontaneous combustion are oxygen content, airflow velocity, particle size, moisture content, and humidity of the air. In a dryer where air flows over the lignite at high velocities, it is rare for the coal to spontaneously combust because the flowing air prevents the development of "hot spots". When dry air flows over relatively moist coal, moisture is removed from the coal through evaporation, resulting in a decrease in temperature. On the other hand, when moist air contacts relatively dry lignite, the lignite adsorbs moisture from the air causing its temperature to rise; thus promoting spontaneous combustion. This means that the relative moisture content of the air has

---

1 Gordon, R Couch, "Lignite Upgrading," IEA Coal Research, IEACR/23, May 1990.

2 Foul, J., Lugscheider, W., and Wallner F., "Entferen von Wasser aus der Braunkohle," Braunkohle: 39(3) 46-57 (March 1987).

an important role in the spontaneous combustion of lignite. Consequently, this study has given special attention to the design of the drying equipment and storage of the dried lignite.

4.2 PROJECT OVERVIEW

This topical report is the second in a series of studies of designs for applications of the U-GAS® gasifier. Subtasks 3.2 and 3.3 are based on a set of criteria that can be applied to a wide cross section of industrial facilities across the United States. Subtask 3.2 established a baseline design for a facility that can be constructed with currently available technology using low-cost coal available in the Eastern United States. Subtask 3.3 examined a variety of alternatives for improving plant costs, using new contaminant removal technologies and examined the use of different plant configurations.

The objective of Subtask 3.2 was to design a first-pass industrial-size, CHP coal-fired gasification system. For the next phase of the study, Subtask 3.3, the objective was to improve the base design from Subtask 3.2 by applying improvements in technologies that are expected to be achieved over the next decade. The goal of this program is to provide guidance to persons interested in replacement or expansion of existing power and heating systems at industrial sites. There are three compelling reasons for considering the use of coal based IGCC plants for this purpose:

- The use of IGCC increases efficiency and reduces operating costs compared to use of premium fuels or continued upgrading of old coal facilities to meet tighter emission standards.
- IGCC is the cleanest means of providing power and steam from coal, thereby reducing emissions from the utility facilities at an industrial site.
- The use coal enhances energy stability and security at the facility. Use of IGCC and/or Combined Heat and Power (CHP) at industrial facilities using coal can contribute to a significant increase in distributed generation for improving site and local power grid security.

The second objective of this study (i.e., to apply the GTI fluidized bed U-GAS® gasifier technology at a stand-alone lignite-fueled IGCC power plant in North Dakota) was achieved with Subtask 3.4. For Subtask 3.4 a base case design was developed for a 251 MW IGCC power plant using lignite at a generic mine-mouth North Dakota site. This design only includes equipment and technology that are commercially available to allow construction at the present time. Some lessons learned during the development of the previous subtasks are included in this base case design, such as the inclusion of metallic filters to clean up the syngas instead of relying upon a scrubber column.

4.3 HEAT INTEGRATION

Extensive heat integration to recover the maximum amount of sensible heat from the facility can improve efficiency of the process. However, this requires more capital investment and can create operational problems when a process (e.g., U-GAS®) is not mature. The philosophy that was used for the design of this plant was to maximize availability by keeping the design as simple as possible. Thus, integration between the various sections of the plant was minimized. Admittedly, this design philosophy results in a less efficient design, but it should produce a design that is less troublesome and has a higher operating factor. Consequently, there is no direct heat integration between the gasification block, coal preparation, and gas turbine/HRSG sections of the plant. In addition, only commercially proven technologies are employed in cleaning the syngas.

The syngas cooling section of the plant is designed to minimize deposition and erosion problems as a result of dust carried in the syngas. Therefore, only one heat exchanger is used before the filter system. This single heat exchanger is a fire tube boiler design that cools the hot syngas leaving the third stage cyclone from about 1600°F to about 650°F by producing saturated steam at 1,000 psig and 546°F. This steam is superheated to 650°F in the HRSG. For maximum electrical efficiency, the 1,000 psig steam should be superheated with the hot syngas before it enters the steam boiler, but this would add another exchanger to the syngas cooling train and would increase the potential for additional deposition and erosion problems.

Upstream of the syngas scrubber, the syngas is used to produce saturated steam at 500 psig and at the same time cooling the syngas to about 360°F. The syngas scrubber removes water-soluble materials from the syngas and simultaneously cools the syngas to about 267°F. It also acts as a final trap to remove any particulates that may have passed through the filter system. The cleaned syngas now is cooled by producing 500 psig steam and hot boiler feed water before the remaining contaminants (such as sulfur compounds, mercury and ammonia) are removed. Removal of the COS is accomplished by passing the syngas through a hydrolysis reactor to convert the COS to H_2S. The H_2S is removed downstream in the acid gas removal unit. The syngas leaving the hydrolysis reactor then is cooled in a series of three heat exchangers. The first exchanger cools the syngas to 236°F by preheating boiler feed water. The second is an air cooler, and the third is a water cooler that cools the syngas to 110°F.

In the above processing scheme, the only interaction between the gasification area and the remainder of the facility is the exchange of steam and boiler feed water. There is no direct heat exchange.

4.4 TECHNOLOGY DRIVERS

There are three primary drivers in terms of energy media selection: cost, emissions compliance, and reliability.

4.4.1 Cost Drivers

Over the past thirty years, natural gas has been generally low cost and certainly the cleanest fossil fuel available for delivering the energy needs to industry. Natural gas delivery is reliable most of the time, although increasing demand for natural gas and a lagging improvement in delivery infrastructure require increasing needs for "back-up" fuels at industrial facilities.

Over the past four years, natural gas prices have risen dramatically. With the price of natural gas currently selling for over 6.00 $/MBtu, many companies are worried about their energy supply costs as near term gas prices are expected to continue higher. The rise in gas prices is forcing industry to critically examine their energy supply choices. Recent articles in the press[3] highlight the closure of chemical companies in the US that rely on natural gas as a raw material, and they are moving overseas where natural gas is less costly. Chemical industry employment is down 7.3% over the past 8 years. Although the Energy Information Administration (EIA) predicts stabilizing gas prices in the next several years, it is important to provide new options for industry to remain competitive and avoid further loss of industrial facilities in this country.

EIA's long term cost projections for delivered industrial natural gas are for prices to decrease (2003 dollars) to 4.37 $/MBtu by 2010 and then slowly increase to 5.47 $/MBtu in 2025.[4] This represents a 4.5% escalation rate in natural gas price, higher than the predicted inflation rate of 3%. In nominal dollars, this rate of increase suggests natural gas prices nearly 8.00 $/MBtu by 2025. Natural gas prices have demonstrated significant volatility over the past few years. This is not expected to change considerably since these variations are based on changes in U.S. supply, demand options, and future world events.

Gas prices paid by industry are not fully reported on EIA databases to retain confidentiality of sensitive company data. Typically only about 12% of industrial pricing is reported. Industry payments for gas vary widely; a key determinant in price variation is whether a company is in a position to bypass the local distribution company (LDC) for gas purchases and buy gas directly via a pipeline. When gas is available bypassing the LDC, the price is about 0.90 $/MBtu above the wellhead price, on average nationally. However, data reported in key industrial states like New York, Pennsylvania, and Ohio, indicates that industries pay as much as 1.00 to 2.20 $/MBtu in transportation cost to the LDC. This could raise the long-term expectation for natural gas prices above those reported by the EIA for delivered cost.

Coal can play a greater role for many industrial facilities. In contrast to natural gas, coal prices have remained stable over the past decade. Coal prices to industrial users are typically between $1.25 to $2.00 per MBtu (highly dependent on fuel type and delivery cost). Furthermore, coal prices are projected by the EIA to remain flat over the next 20

3 Malita Marie Garze, Chicago Tribune, Energy Costs an Offshore Factor, 4/25/2004.

4 Annual Energy Outlook 2005 with Projections to 2025, www.eia.doe.gov/oiaf/aco/economic.html

years. This nets a fuel cost differential in favor of coal of roughly 3.00 to 5.00 $/MBtu, depending on the specific fuel transportation factors to a given facility.

An alternative for industry would be to use coal gasification to convert low cost coal to a fuel gas to take advantage of high-efficiency IGCC technology for generation of heat and power for their facilities. This study suggests that the costs for conversion of coal to syngas for an IGCC application is about 4.10 $/MBtu. However, conversion of a solid fuel to gas is capital intensive, and the cost is high. Thus, the critical decision for implementation of this technology lies in the long term differential fuel costs between coal and natural gas including the attendant emission controls associated with their use. Although not every industrial facility can benefit from coal gasification on a purely price basis, there are clearly many facilities that can justify a serious evaluation of this technology as long term solution to meeting its energy needs.

The most likely target facilities for early adoption of coal gasification would be: 1) facilities that cannot buy gas directly from national pipelines or; 2) old, inefficient coal fired boilers that may be able to reduce energy costs through the use of gasification.

4.4.2 Emission Drivers

Natural gas has been the industrial fuel of choice for the past 30 years. Natural gas is flexible, clean and convenient. For many years, natural gas was available at a cost lower than either liquid or solid fuels. Gas was chosen for many installations because it allowed conversion of existing coal fired boilers and avoided the added cost of installing emission control equipment for sulfur and NOx control. Some facilities have switched back to using coal in recent years as natural gas prices have increased.

By the end of 2005 the U.S. EPA is planning to release new standards for emission controls at industrial plants that will require essentially all sites with combustion facilities rated at over 10 MBtu/hr to apply state-of-the-art emission controls. Emission control will be required for sulfur, NOx, particulates, mercury, and possibly chlorides. Post combustion control for all these emissions will require significant expense for industrial utility systems that are in many cases 40 to 60 years old. Replacement of old coal fired equipment with new systems at an industrial scale is relatively expensive. This application of IGCC technology has been demonstrated to be environmentally superior to post combustion emission controls and can be applied to industrial facilities in a cost effective manner.

4.4.3 Reliability Drivers

Reliability is a tangible factor for industrial applications; however the value of reliability can only be quantified by each facility individually. EPRI and others have studied costs associated with the loss of manufacturing and industrial productivity. These studies reflect the importance of an uninterrupted supply of electric power and steam to an industrial facility. Often, a brief electrical outage of only a few minutes can cause a shutdown of units that can result in hours or days of lost production. For this reason, many companies have invested in emergency backup generators to provide power to critical applications in the event of an outage. These units are typically only used for backup (when using liquid fuels) and are limited in the annual number of hours for which they can be used.

For many years, industry was able to purchase electricity and gas from their local suppliers on an "interruptible" contract basis. This allowed the local utility to call the company in times of short supply to curtail their energy use. This ensured reliability to the entire community by reducing the energy use of several large consumers. This was acceptable as long as operations were not interrupted frequently, and the cost of lost production was significantly less than the purchase of "firm" energy delivery from the utility. This type of service has become less acceptable to industry because they are now operating at much higher use factors; lost production is more costly, and secondly, utilities are more apt to enforce interruptible contracts than they were in the past.

Many industrial facilities have found that for reliability and economy it is most effective to self generate all or part of their electrical needs with steam. This provides a reliable source of electric power as well as thermal energy to meet the heating and cooling demands of their facilities. Such combined heat and power (CHP) facilities are common across the country; however there are many facilities that do not take full advantage of their ability to maximize efficiency with CHP. This is largely due to the low cost energy that was available from suppliers many years ago when these plants were built.

Coal based IGCC facilities can be a secure source of energy for industrial plants. Self generated electricity and steam can provide the bulk of a facilities power and thermal needs, while coal stored on site can provide fuel to the plant on an uninterruptible basis.

4.5 PLANT SIZE

The plant consists of a single GE 7FB gas turbine, one HRSG and a steam turbine with an export power output of 251 MW. This output size was selected because it is about the minimum size for a base load power plant. Furthermore, multiples of this size equipment can be readily developed to provide facilities of a larger scale.

Syngas to power the gas turbine is supplied by a single gasifier train using GTI's U-GAS® fluidized bed gasifiers. For the purposes of the study, the GE 7FB engine was selected for the gas turbine. Each turbine requires 1,801.7 MBtu/hr to produce 210.78 MW (net). Waste heat from the turbine and the gasification system is used to produce about 570 klb/hr of steam; a portion of which is used internally, and the rest is used for additional power generation.

The gasification system contains several subsystems:

- Coal Preparation – Handling, Sizing and Drying
- Gasifier Island

- Syngas Cooling
- Syngas Cleaning (including sulfur removal & recovery)
- Power Island
- Auxiliary Systems

A generic overall plot plan for the Subtask 3.4 facility is shown in Figure 4.2.

A project construction schedule is shown in Addendum E.

4.6 STUDY PERCEPTIONS AND STRATEGIC MARKETING CONSIDERATIONS

This study is directed at a large audience, which has many viewpoints, expectations and objectives. This study results are presented in a format that addresses these perceptions and strategic marketing considerations. If an in-depth evaluation of any specific project or projects is required, a gasification technology vendor, such as GTI, should be contacted. The following is a list of what we believe to be our reader's major points of interest.

Promotion (or Planning Studies) – This report basically describes what is a series of planning studies for various coal fueled IGCC applications. General economics were developed using a discounted cash flow model. These general results should allow prospective IGCC project developers to consider the merits of further evaluations of IGCC technology on a project specific basis.

Precision – Using cost information from Price and Delivery Quoting Service for Chemical Process Equipment (PDQ$®), vendor quotes and previous designs allowed the cost estimates to have a high degree of confidence or expressed differently, a minimum amount of uncertainty.

Potential – This study addresses the potential of GTI's gasification technology to reduce the cost and improve the efficiency of IGCC plants. Further cost savings have been identified for study, but not yet quantified.

Place (location) – The North Dakota location seems to be the best location for a lignite evaluation because of the costs and safety issues associated with the transportation of lignite.

Product (or Market Penetration) – Currently lignite-fueled IGCC plants have a feedstock price advantage over eastern coal fueled facilities.

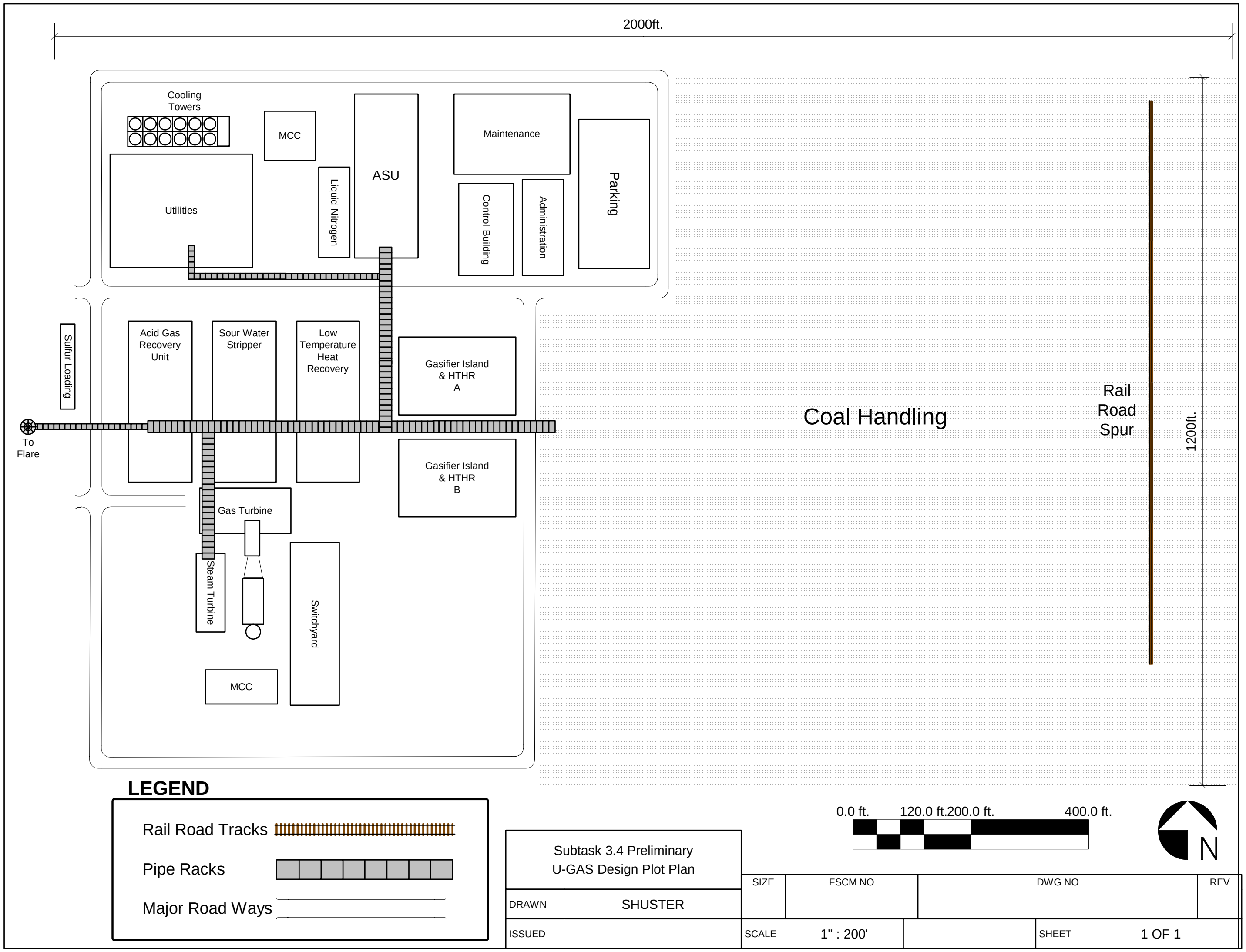
Proliferation – As more IGCC plants are built, their costs will decrease, availability will improve, and companies will be more willing to proceed with the construction of additional IGCC plants.

Promise – IGCC plants can have higher efficiencies than pulverized coal (PC) facilities with the potential of further increased efficiencies coupled with lower costs. The potential of very low SO_2 and NOx emissions coupled with CO_2 capture are possible in the near future.

Promote – This study promotes the development and implementation of IGCC by demonstrating that it is possible to build an IGCC plant that can produce electricity at competitive prices.

Prospectus – IGCC project development requires detailed analysis and planning on a project specific basis. Study performance may not be indicative of or adequate to quantify future revenues.

Figure 4.2 Overall Plot Plan for Subtask 3.4

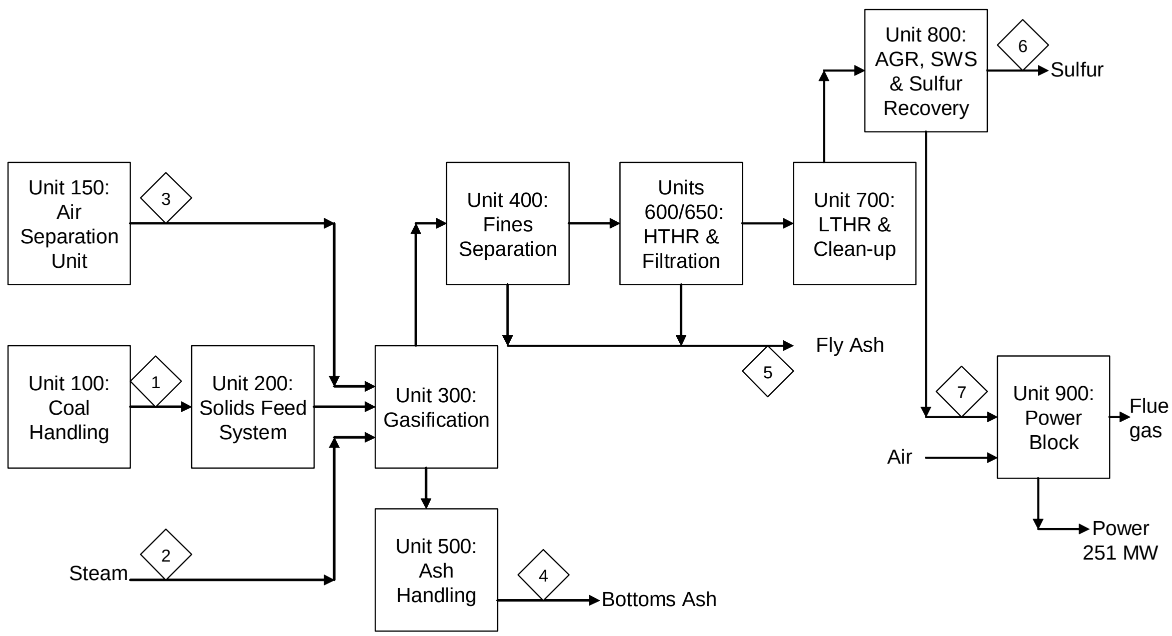


5.1 INTRODUCTION

This section describes the base case design of the lignite fueled IGCC power plant.

Figure 5.1 contains a simplified Block Flow Diagram and material balance of the lignite-fueled facility. The complete material balance is shown in the Addendum D.

Figure 5.1 Simplified Block Flow Diagram and Material Balance



Stream Number	1	2	3	4	5	6	7
Description	**Coal to Gasifier**	**Steam to Gasifier**	**Oxidant to Gasifier**	**Bottom Ash**	**Fly Ash**	**Sulfur Product**	**Clean Syngas to Gas Turbine**
Pressure, psia	14.7	500	500	14.7	14.7	NA	375
Temperature, F	70	550	590	1600	1600	NA	110
Flow Rate, lb/hr	266,450	62,968	127,462	18,399	5,329	1,577	345,865

5.2 PLANT CONFIGURATION

5.2.1 Lignite Preparation – Handling, Sizing and Drying (Unit 100)

North Dakota lignite will be delivered by railcars to the site, handled and processed for feeding the GTI gasifiers (one operating and one spare).

The Lignite Coal Handling System starts at the unloading area where unit trains unload the lignite coal one car at a time to under track hoppers. The unloading area includes an 80 foot long x 30 foot wide x 20 foot high building that admits a 100 ton capacity rail car. The building is provided with wall mounted infrared radiant heaters and track heaters for thawing carloads of frozen coal. The building includes a car puller and positioner for quick turnaround time and a car shaker to loosen the frozen coal to provide effective unloading at the desired rate. Each car would be indexed over the pit and a Kinergy car shaker would index against the side of the car, then the load would be dumped into a two compartment pit. The site will receive one train unit, consisting of a minimum of 100 railcars, each car having a 100 ton capacity for a total of 10,000 tons. The power plant is responsible for breaking the train into segments and spotting the coal cars at the unloading facility. The facility is provided with railroad tracks for moving railcars in and out of the unloading building and around the plant site. The facility also includes a 2,000 hp diesel locomotive switch engine that will be used to move the rail cars on the power plant's property.

Coal delivery is made every 2nd work day. Coal will be unloaded from the rail car at a rate of 1,250 tph (1,500 tph maximum) and transferred to the active pile or to the inactive (reserved) storage by way of the Transfer Tower belt conveyors. This Transfer Tower will include a metal detector and magnetic separator to remove and collect tramp irons. The wet lignite at the Transfer Tower will be discharged using a diverter valve to convey the flow either to the 30 day storage pile or to the 7 day storage pile. The active pile requires 7 days storage or 29,842 tons based on a design rate of 4,263 tpd at 40% moisture at the crusher inlet. The reserve pile will provide 30 day storage or 127,900 tons. Once the lignite reaches the 7-day storage pile, it will be reclaimed utilizing three vibrating storage pile reclaimers and vibrating feeders to draw down the pile and discharge the material onto a belt conveyor. This belt conveyor will be in a tunnel underneath the 7-day storage pile. The active pile stacker and reclaim conveyor will transfer the coal first to a diverter valve which will then discharge to another belt conveyor that feeds one of the 100% Heyl & Patterson crushers inside the coal handling building.

The lignite coal handling system will require one 100% crusher and three 33.33% dryer units (based on the maximum size available), to meet the 100% plant capacity. The plant will be provided with one additional 100% crusher and one additional 33.33% dryer as stand-by units.

The gasifier feed coal specification calls for 98% passing a ¼" screen and no more than 10% less than 100 mesh (fines). The coal is dried to 20% moisture (refer to Section 4.1). (A 40% moisture content of as-received lignite is used for dryer design; normally the lignite moisture content of the as-received lignite will be about 32.2%.) To achieve these specifications, one of the two 100% crushers receives coal with the largest lump size of 2" x 0" and breaks the coal to the top size of ¼". The oversized coal is recirculated to the inlet of crushers until the ¼" top size is met. The ¼" sized coal and fines are then conveyed to a common surge bin which discharges to three screw feeders. Each screw feeder delivers crushed coal to associated fluidized-bed dryers which utilize low level waste heat and some 50 psig steam as the heat sources for drying the lignite. The annual average air temperature is 40°F and must be heated to 245°F for drying purposes. The low-level waste heat sources are the cooling water return, stripped water from the sour water stripper unit and sour water from the syngas scrubber. Steam (50 psig) is used to meet the final air outlet temperature. The start-up boiler is used to supplement the steam system when high moisture lignite (40%) is used. The dried coal is blown with the hot air to the cyclone separator for gas-solid separation. The hot air from the separator is passed to the baghouse for particulate removal. The product lignite at ¼" top size and 20% moisture by weight is discharged from the fluidized-bed dryer and mixed with the fines from cyclone separator before conveying to the silo.

The manufacturers of the crusher and dryer units have indicated the equipment availability at 97%, and the Mean Time to Repair is typically one week. The crusher/dryer manufacturers also have indicated that the equipment will have negligible adverse emissions to the environment.

The dried coal is discharged to a vibrating screen where any coal greater than ¼" is separated and recirculated to the crusher. To ensure that no more than 10% of the gasifier feed is fines (less than 100 mesh), the coal is applied to a 120 mesh screen. Fines passing through the 120 mesh screen are expected to be of a small amount and will be rejected and collected in a dust collector. Adjustments in grinding can be performed if fines become significant. If fines cannot be reduced, a pneumatic transport system can be installed to collect the excess fines for transport to offsite boilers. This collection and conveying system will not be required if grab sample analyses indicate that total amount of fines are less than 10%, which means that all of the coal discharged from the vibrating screen will be used and transported to the silo.

The 24 hour storage silo is approximately 800 feet from the railcar unloading area. The proposed coal handling building is 125 ft. long x 110 ft. wide x 75 ft. high for the crusher–dryer units. The building will contain the coal handling equipment, including the proposed two 100% crusher and four 33.33% dryer units. The fugitive dust emissions inside the building (from the crusher-dryer units) are negligible. The rest of the coal handling equipment after the vibratory screen discharge feeder will be located outside the building. The start-up coke handling equipment from the delivery truck to the coke silo discharge feeder also will be located outside the coal handling building. All equipment located outdoors will be weather protected and tightly sealed to prevent dust leaks.

The dried coal from each dryer unit is conveyed by a screw feeder at the design rate of 1,066 tpd at 20% moisture by weight to a common bucket elevator. The common elevator takes the coal to the top of the 24 hour primary silo for storage at the design rate of 3,197 tpd. The primary lignite coal silo is about 44 feet in diameter with a cylindrical height of 100 feet. The silo top is approximately 150 feet above grade.

Dried coal at 20% moisture is discharged from the 24 hour silo at a design rate of 133 tph either to the primary screw feeder during normal operation or to the redundant screw feeder as a back up when the primary feeder is out of service. The primary screw feeder discharges to a bucket elevator, which takes the coal to approximately 120 feet above grade, conveys and transfers it to a surge hopper, and finally to a common distribution feeder that supplies the four gasifier weigh feeders. The redundant screw feeder takes the silo coal to a redundant elevator which functions similar to the primary elevator, except that the redundant elevator also is used by the start-up coke handling system as described below.

The distribution screw feeder, which takes the coal or start-up coke from either the primary or redundant elevator, discharges the coal to the weigh feeder at approximately 67 tons (30 minutes of lignite coal feed) each. The distribution screw conveyor includes a grab sampling port before the first weigh feeder opening for coal sampling and analysis. If the fines exceed the 10% maximum limit, the primary silo will be scheduled for cleaning at a predetermined coal storage level. Cleaning will be performed in accordance with the plant maintenance procedure.

The start-up coke handling system is designed for outdoor installation and is provided for gasifier start-up. Coke is delivered by truck and unloaded to a hopper that feeds the belt conveyor that transfers the coke to a bucket elevator. The bucket elevator takes the coke to the top of the 12 hour coke storage silo. The coke silo stores 350 tons and is approximately 20 feet in diameter by 80 feet cylindrical height. Coke is discharged at the bottom and conveyed to the redundant elevator that takes it to the common distribution screw conveyor for supplying the gasifier weigh feeders. The coke is fed to the gasifiers at a rate of 5 to 30 tph.

5.2.2 Air Separation Unit with Nitrogen Compression (Unit 150)

The Air Products cryogenic air separation unit (ASU) provides 1,530 tons per day of a 95% purity oxygen stream (1,462 tpd of contained oxygen) at about 90°F and 500 psia. It also provides 4,958 tpd of a 99.3% purity nitrogen stream at 219°F and 500 psia and contains storage for 250 tpd of nitrogen for internal use. The plant consumes 40.9 MW of power, 4,450 lb/hr of 50 psig steam, and requires 9,400 gpm of cooling water.

Figure D.2 in Addendum D is a schematic diagram of the air separation unit. This ASU is similar to the one that Air Products installed at Tampa Electric's Polk power station except that the oxygen compressor has been replaced by an air booster compressor / pumped liquid oxygen system.

Ambient air is compressed in the main air compressor, before it is sent to a pretreatment section where impurities such as H_2O and CO_2 are removed in an absorption section. The absorption section is a two-bed system with one bed always in service, and the other is on regeneration. Regeneration is done with dry nitrogen from the main nitrogen production stream.

The air then enters the main heat exchanger, where it is cooled by the oxygen and nitrogen product streams. A compressor/expander (Compander) provides refrigeration for the system. Cryogenic distillation occurs in a standard two column arrangement: one column operates at an elevated pressure and the other at a reduced pressure.

5.2.3 Gasification Island (Units 200-500)

The gasification system is enclosed in a building.

The gasifier train consists of the following elements:

- Coal Lockhopper Feed System
- Gasifier
- Startup Heater
- Dust Cyclones
- Dust Removal System
- Ash Removal System

5.2.3.1 Coal Lockhopper Feed System (Unit 200)

The dried and crushed coal is fed to the gasifier via a lockhopper system that transfers the coal from atmospheric pressure to the operating pressure of the gasifier. Each gasifier has two lockhopper feed trains, each sized to deliver 100% of the design feed rate to the gasifier. This allows for complete redundancy in the event of disruption of the fuel feed in one of the feed trains. Each lockhopper is designed for four cycles per hour, but is capable of operating at up to eight cycles per hour.

5.2.3.2 Gasification (Unit 300)

The gasifier vessels are refractory lined. An outer layer minimizes the heat loss, and the inner layer is made of abrasion resistant material to withstand the rigorous environment of the gasifier. A grid supports the gasifier bed. Oxygen and steam enter the gasifier below the grid. Fuel is fed to the gasifier just above the grid. Solids collected from the first- and second-stage cyclones are returned to the gasifier bed just above the grid. The bed of solids in the gasifier is maintained at a sufficient depth to ensure adequate residence time for high carbon conversion and to minimize tar/oil formation in the gasifier. The gasifier is approximately 45 feet tall that is of sufficient height for the grid, bed, and disengaging zones. The syngas temperature exiting the gasifier when operating on North Dakota lignite is approximately 1600°F. The gasifier operates at 450 psia to provide adequate available pressure through the plant and to the gas turbine.

In operation the gasifier consumes 213,160 lb/hr (2,558 tpd) of moisture-free lignite. Oxygen and saturated 500 psig steam are mixed and fed to the gasifier to react with the coal. At design conditions 127,462 lb/hr of oxygen (95% oxygen) and 62,968 lb/hr of steam are required for gasification.

The product syngas leaving the gasifier has the following composition (mol pct., see Table D.2) as determined by GTI.

CO	33.47%	H_2S	0.26%
CO_2	17.00%	COS	0.01%
H_2	24.93%	NH_3	0.20%
H_2O	15.77%	HCN	0.01%
CH_4	7.05%	N_2	1.30%
		HCl	202 ppm

The steam to carbon weight ratio is 0.45 and the oxygen to carbon weight ratio is 0.87. Small quantities of light oils (primarily benzene and naphthalene based compounds), dust, chlorides, and mercury also are in the syngas stream and must be removed in the downstream cleanup system.

5.2.3.3 Startup

Natural gas or another suitable fuel (e.g., LPG) is required for gasifier startup after an outage. The startup heater is used to heat the gasifier and downstream refractory-lined equipment to about 1200°F. Once the gasifier has stabilized at this temperature, metallurgical coke is introduced to establish a bed of solids and to increase the operating temperature to that required for feeding coal. This startup method reduces the likelihood that tars/oils will build-up in the refractory-lined vessels and equipment when they are cold.

5.2.3.4 Dust Cyclones

A series of three cyclones increase carbon conversion and reduce the contaminant dust concentration in the syngas. Solids separated by the first- and second-stage cyclones are recycled back to the gasifier to maximize carbon conversion and process efficiency. Particulates collected by the third-stage cyclone are discharged via a lockhopper system to the dust collection and removal system.

The cyclones are fabricated of refractory inside a large carbon steel pipe. This avoids the requirement for exotic materials capable of operation at high temperatures. The solids collected by the first- and second-stage cyclones are recycled to the gasifier in a refractory lined pipe.

5.2.3.5 Dust Removal System (Unit 400)

The dust (fly ash) removal system cools the dust and transports it via lockhoppers from the gasifier pressure to storage at atmospheric pressure. A pressurized cooling screw cools the dust to about 500°F to protect the lockhopper valves and to allow the downstream use of carbon steel equipment. The screw rotates when the valve to the lockhopper is open and is stopped when the lockhopper valve is closed. A refractory lined surge hopper collects dust when the screw is not rotating (lockhopper closed). When the lockhopper is full (confirmed by nuclear level detectors) the upper valve is closed, the vessel is depressured to atmosphere, and the discharge valve is opened.

Dust is then transported via a pneumatic system to a day tank from which it can be disposed or sold.

After the lockhopper is emptied, the discharge valve is closed and the vessel is pressurized with nitrogen (see Section 5.2.10.11) to the gasifier pressure. After pressure is attained, the upper fill valve is opened and the screw restarted. The screw operates at a sufficient speed to empty the contents of the surge hopper that accumulate during the cycling of the lockhopper.

5.2.3.6 Bottom Ash Removal System (Unit 500)

The bottom ash removal system cools the ash and transports it via a lockhopper from the gasifier pressure to storage at atmospheric pressure. A pressurized cooling screw cools the bottom ash to about 500°F to protect the lockhopper valves and to allow downstream use of carbon steel equipment. The screw is rotated when the valve to the lockhopper is open and stopped when the lockhopper valve is closed. A refractory lined surge hopper collects ash when the screw is not operating (lockhopper closed). When the lockhopper is full (confirmed by nuclear level detectors) the upper valve is closed, the vessel pressure is lowered to atmospheric, and the discharge valve is opened. Ash is then transported via a pneumatic system to a day tank from which it can be disposed or sold.

After the lockhopper is emptied, the discharge valve is closed and the vessel is pressurized with nitrogen to the gasifier pressure. After pressure is attained, the upper fill valve is opened and the screw restarted. The screw operates at a sufficient speed to empty the contents of the surge hopper that accumulate during the cycling of the lockhopper.

5.2.3.7 Characteristics of Raw Syngas

Table 5.1 lists the major characteristics of the syngas exiting the gasifier. The residual particulates in the syngas stream leaving the gasifier consist of ash, unburned carbon, and small amounts of trace elements. The oils produced in the U-GAS® gasifier are mainly benzene and naphthalene based compounds.

Table 5.1 Major Characteristics of the Syngas Leaving the Gasifiers

Temperature Leaving the Gasifier, °F	1600
Pressure, psia	440
Gas Mass Flow Rate, lb/hr	433,151
Solids, lb/hr	5,329
Total, lb/hr	438,480
Water, lb/hr	56,390
Oils Condensation Temperatures, °F	180 ~ 450
Dew Point, °F (@439.7 psia)	355
Ammonium Chloride Condensation Temperature, °F	~ 540

5.2.4 High Temperature Heat Recovery (Unit 600)

5.2.4.1 Introduction

The high temperature heat recovery system recovers the sensible heat from the syngas by producing saturated 1,000 psig steam, most of which is routed to HRSG to produce 1,000 psig superheated steam. The design objectives are to maximize the syngas sensible heat utilization, to maximize reliability, and to minimize operation difficulties. The syngas leaving the gasifiers contains particulates, light oils, chlorides, ammonia, etc. Each of these undesirables alone and in combination significantly impacts the design.

Particulates in the syngas stream have presented challenges to plant designers and operators. The difficulties mainly are related to plugging heat exchanger tubes, equipment damage, and degrading the downstream acid gas removal systems. Due to the presence of the particulates, the syngas is erosive; on the other hand, the syngas flow velocity needs to be maintained relatively high to avoid the particulates from settling inside the heat exchanger equipment. Thus, system design and materials selection are critical to ensuring high reliability.

The light oils (mostly benzene) in the syngas create a different set of challenges. If the syngas temperature is lower than the condensation temperatures of the oils, the oils will adhere to the equipment surfaces; moreover, if particulates are present, they will tend to agglomerate, and thus, intensify the plugging.

Ammonium chloride formed during the coal gasification process starts to condense and deposit on the equipment surfaces between 480°F and 540°F, which could lead to plugging if it is not removed. In addition to the detrimental effects of particulates, oils and ammonium chloride, there are negative effects of chlorides and other acids. If the temperature of the syngas is below the dew point, the acids will dissolve in the condensate, creating a severely corrosive environment.

5.2.4.2 Design Basis

Based on the aforementioned considerations and to ensure a robust and reliable design, the following three principles were established

1. The operating pressure of the clean stream (e.g., steam) should be maintained at a higher level than that of the syngas stream. This reduces the likelihood of particulate laden syngas contamination of the clean stream if a heat exchanger tube breach occurs.

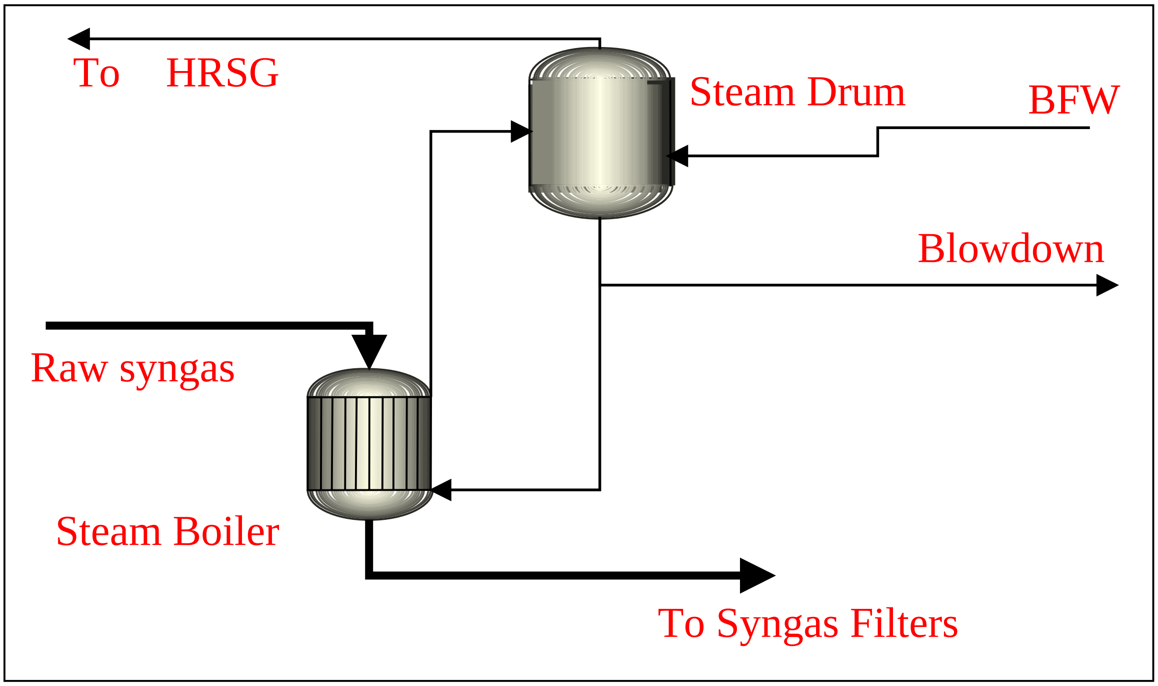
2. The temperature of particulate laden syngas stream in a heat exchanger should always be above its dew point to minimize the potential for condensation of light oils and ammonium chloride.

3. Keep the design simple. Particulates tend to damage equipment and accumulate where syngas flow velocity changes. It is important to have a system with a minimum number of pieces of equipment and geometric changes.

5.2.4.3 System Description

Figure 5.2 shows the schematic flow diagram for the high temperature heat recovery system, which comprises a steam boiler and a steam drum. The raw syngas enters at about 1600°F and is cooled to 650°F. A thermosyphon loop is employed between the steam boiler and the steam drum. Boiler feed water from the boiler feed water preheater enters the steam drum at approximately 350°F and 1,035 psig where it mixes with the steam produced in the steam boiler. The liquid water in the steam drum circulates back to the steam boiler while the saturated steam at 1,025 psig is sent to the heat recovery steam generator (HRSG) in the power block to produce superheated 1,000 psig steam at 650°F. To prevent foreign matter from accumulating in the steam drum, a small percentage of liquid water is extracted as blowdown and sent to the waste water treatment (WWT) unit. For startup, a small pump is required to establish the thermosyphon flow pattern.

Figure 5.2
Schematic Flow Diagram for the High Temperature Heat Recovery System



The steam boiler is a vertical, one-pass shell-and-tube heat exchanger, with the inlet head being refractory lined for erosion protection. Syngas flows downwards on the tube side while the water flows upwards on the shell side. The average syngas flow velocity in the tubes is about 40 ft/sec, and the overall heat transfer coefficient is calculated to be about 65 Btu/hr-ft^2-°F. Inconel is recommended for the tubes for better erosion resistance. The heat exchanger has 195 tubes with a length of about 31 ft long and an inside diameter of two inches. To insulate the tube sheet and protect it from erosion, the gas side of the tube sheet is refractory lined, and the tube inlets are equipped with ferrules to minimize the thermal stresses on the tube sheet/tube joints. The steam boiler is large enough to have a 30-minute liquid hold-up time. Equipment specifications are given in Addendum B.

The design of this heat exchanger is similar in many ways to that at the Wabash River plant, where the syngas exiting the second stage of the gasifier is cooled from about 1900°F to about 700°F. The boiler is a vertical fire tube heat exchanger with the syngas on the tube side. The cooled syngas goes into a particulate removal unit after exiting the syngas cooler. Boiler feed water enters a steam drum that forms a thermosyphon loop with the boiler. The high-pressure steam produced in the syngas cooling system is then superheated in the gas turbine heat recovery system. The experiences gained and lessons learned at the Wabash River plant and Tampa Polk power station served as the basis for this design. The overall heat transfer coefficient in the steam boiler is comparable to those in the exchangers used in the earlier U-GAS® gasifier plants, including the Shanghai plant. Based on these considerations, this system should achieve the design objectives.

A marked benefit of this design is its simplicity. It minimizes the potential of plugging and damage to the equipment by particulates. In the steam boiler, the syngas flows downwards towards the bottom head, reducing the likelihood of tube plugging. In addition, the syngas exits the steam boiler at 650°F, preventing any complications caused by the condensation of water, oils, and ammonium chloride. The simplicity of this design translates directly into being low cost. On the other hand, the particulates in the syngas stream may result in increased erosion. Periodic cleaning of the steam boiler is recommended to remove deposits that accumulate on a regular basis (at least once per year). Experience indicates that these issues are at a manageable level and comparable to maintenance issues for traditional combustion systems.

5.2.5 Syngas Filters (Unit 650)

5.2.5.1 Introduction

The syngas filters are porous sintered metal filters that remove the residual particulates that are not captured by the series of cyclones downstream of the gasifier vessel. One filter system is necessary for each gasifier and high temperature heat recovery (HTHR) train. The filter system selected is a Pall Corporation Gas Solid Separation System that will remove >99.99% of the particulates from the syngas. The remaining solids are removed from the syngas in the scrubber column to ensure that a solids-free syngas goes to the gas turbines.

5.2.5.2 Design Basis

The syngas (650°F and 425 psia) exiting the HTHR system contains approximately 2,665 lb/hr of residual particulates or char. To reduce downstream complications due to the presence of the fine solids (e.g., erosion, agglomeration, etc.), a particulate removal system is necessary. Based on operating experience at the Wabash River plant, sintered metal candle filters were selected. The filters remove >99.99% of the char, and leave less than 0.15 lb/hr of particulates remaining in the syngas stream exiting the filters. Maximum pressure drop across the filter assembly is approximately 5 psig.

5.2.5.3 System Description

The candle filter system was designed by Pall Power Generation. Each particulate filter system consists of a single carbon steel vessel configured such that the syngas enters the vessel near the bottom and flows vertically upwards. Inside the vessel, 312 filter elements hang downward, and the char is gathered on the outside of the filter elements. Each vessel is divided into six sections, with each section containing 52 individual filter elements. The elements are cyclically cleaned using nitrogen blowback based on pressure differential, with each blowback cycle duration of 1.3 seconds at a frequency of approximately 30 minutes. Char is collected at the bottom of the vessel and removed by a lockhopper system.

5.2.6 Mid-Temperature Heat Recovery (Unit 600)

5.2.6.1 Introduction

The mid-temperature heat recovery system recovers the sensible heat from the syngas by producing saturated 500 psig steam, most of which is routed to HRSG to produce 500 psig superheated steam. The design objectives are the same as the high temperature heat recovery system, to maximize the syngas sensible heat utilization, to maximize reliability, and to minimize operation difficulties. Although the candle filters upstream of the mid-temperature heat recovery system have removed the particulates, the syngas still contains light oils, chlorides, ammonia, etc. Each of these undesirables alone and in combination significantly impacts the design.

As described in Section 5.2.4, the light oils (mostly benzene) in the syngas create a set of challenges. If the syngas temperature is lower than the condensation temperatures of the oils, the oils will adhere to the equipment surfaces; moreover, if particulates are present, they will tend to agglomerate, and thus, intensify the plugging.

Ammonium chloride formed during the coal gasification process starts to condense and deposit on the equipment surfaces between 480°F and 540°F, which could lead to plugging if it is not removed. In addition to the detrimental effects of oils and ammonium chloride, there are negative effects of chlorides and other acids. If the temperature of the syngas is below the dew point, the acids will dissolve in the condensate, creating a severely corrosive environment. While the mid-temperature heat recovery system is designed to reach syngas temperatures of about 500°F, it is expected that periodic water washing will reduce any negative effects such that design operation and reliability targets will be met.

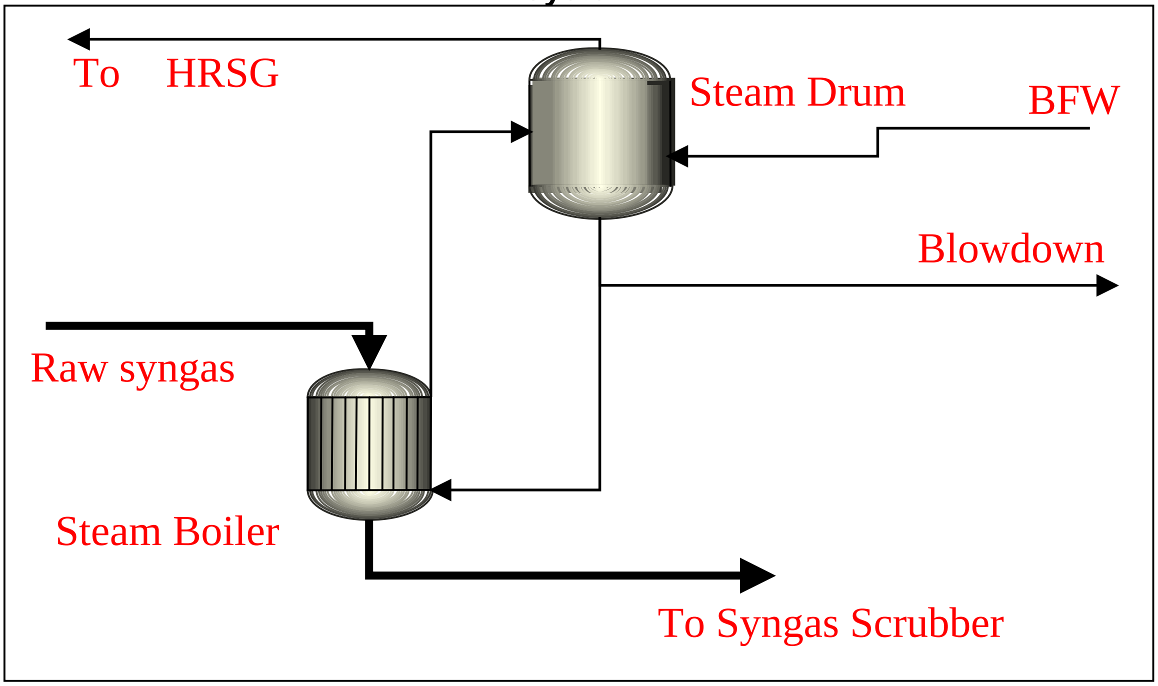
5.2.6.2 Design Basis

As in the case of the high temperature heat recovery section, the operating pressure of the clean stream (e.g., steam) should be maintained at a higher level than that of the syngas stream. In the case of a filter failure, this reduces the likelihood of particulate laden syngas contamination of the steam system if a heat exchanger tube breach occurs.

5.2.6.3 System Description

Figure 5.3 shows the schematic flow diagram for the mid-temperature heat recovery system, which comprises a steam boiler and a steam drum. The raw syngas enters the mid-temperature heat recovery system at 650°F and is cooled to 501°F. A thermosyphon loop is employed between the steam boiler and the steam drum. Boiler feed water enters the steam drum at approximately 350°F and 535 psig upon exiting the low temperature boiler feed water preheater, where it mixes with the steam produced in the steam boiler. The liquid water in the steam drum circulates back to the steam boiler while the saturated steam at 525 psig goes to the HRSG to produce superheated 500 psig steam. To prevent foreign matter from accumulating in the steam drum, a small percentage of liquid water is extracted as blowdown and is sent to the waste water treatment (WWT) unit. For startup, a small pump is used to establish the thermosyphon flow pattern.

Figure 5.3 Schematic Flow Diagram for the Mid-Temperature Heat Recovery System



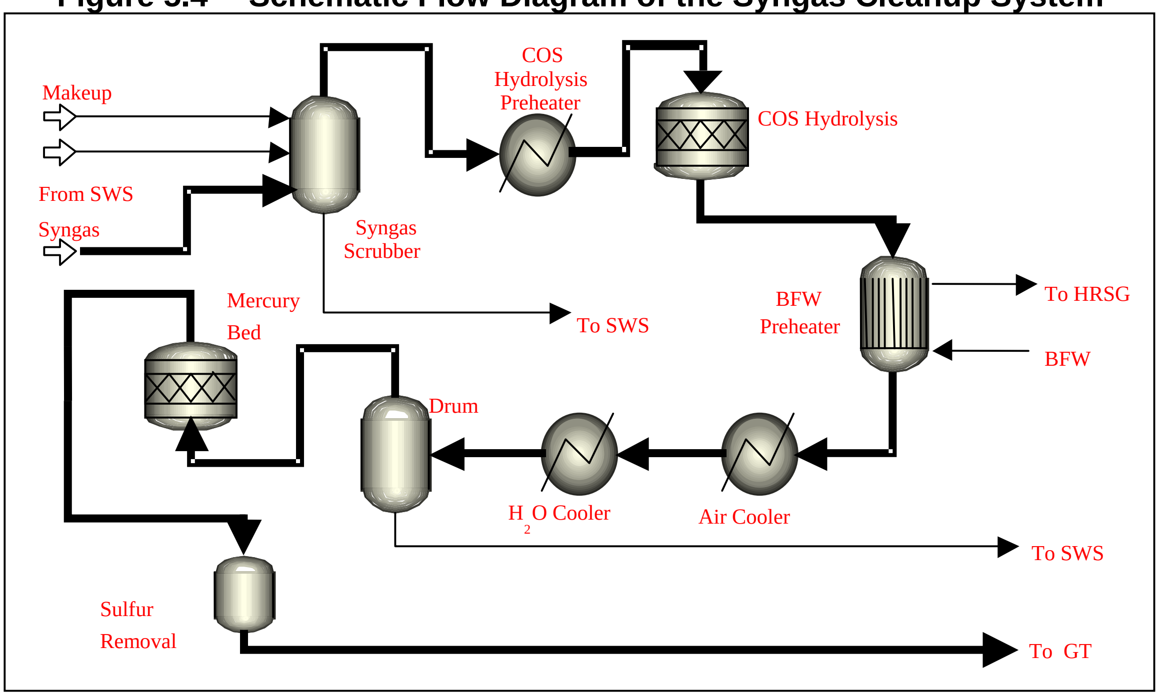
The steam boiler is a vertical, one-pass shell-and-tube heat exchanger, with the inlet head being refractory lined for erosion protection. Syngas flows downward on the tube side while the water flows upwards on the shell side. The average syngas velocity in the tubes is about 18.5 ft/sec, and the overall heat transfer coefficient is calculated to be about 65 Btu/hr-ft^2-°F. Because the particulates were removed upstream in the candle filters, carbon steel can be used for the tubes as opposed to Inconel, which was used in the 1,000 psig steam boiler. The heat exchanger has 380 tubes with a length of about 22 ft and an inside diameter of two inches. To insulate the tube sheet and protect it from erosion, the gas side of the tube sheet is refractory lined, and the tube inlets are equipped with ferrules to minimize the thermal stresses on the tube sheet/tube joints. The steam boiler is large enough to have a 30-minute liquid hold-up time. Equipment specifications are given in Addendum B.

5.2.7 Syngas Cleanup System (Units 700 and 800)

The syngas cleanup system removes particulates, ammonia, chlorides, oils, etc. from the syngas prior to sulfur removal in the acid gas removal system and combustion in the gas turbines. To ensure the proper operation of the acid gas removal system and the gas turbine, it is critical to remove the undesirables from the syngas such as the chlorides, particulates, ammonium chloride, etc.

The syngas cleanup system consists of a filter systems, a syngas scrubber, a COS hydrolysis unit, and low temperature heat recovery. Figure 5.4 shows the schematic flow diagram of the syngas cleanup system. The syngas streams from the gasification island passes through the filter system, and then merge before they enter the syngas scrubber. The following discussion describes each of the main components. A detailed equipment list can be found in Addendum B.

Figure 5.4 Schematic Flow Diagram of the Syngas Cleanup System



5.2.7.1 Syngas Scrubber

The syngas scrubber column removes trace organic compounds and water soluble inorganic compounds from the syngas. It also cools the syngas from 501°F to about 263°F. Furthermore, it provides a final trap to remove any particulates that may have passed through the filters.

The scrubber is an impingement column. Gas flows upwards through baffles in the column while the water flows downward. The washed syngas emerges at the top of the column while the sour water leaves from the bottom and goes to the sour water stripper (SWS). The scrubber makeup water is a combination of clean process water and recycled water from the SWS. By using recycled water from the SWS, the amount of fresh make-up water is minimized. Table 5.2 lists the water sources and stream data related to the wash column. The cleaned syngas leaving the wash column is saturated with water. Approximately 60% of the water entering the scrubber with the syngas leaves with the syngas. All oils, most of the chlorides, and a part of the ammonia are removed from the syngas in the wash column. Detailed stream data are given in Addendum D.

Table 5.2 Water Balance in the Syngas Scrubber Column

	Inlets			Outlets	
	Syngas	**Recycled Sour Water**	**Fresh Quench Water**	**Cleaned Syngas**	**Sour Water**
Temperature °F	501	110	80	263	263
Water flow, lb/hr	56,390*	328,500	39,160	32,532	391,518

* Contained water in vapor stream

5.2.7.2 COS Hydrolysis Unit

Most of the sulfur in the coal is converted to hydrogen sulfide (H_2S) during gasification; however, a small portion is converted to carbonyl sulfide (COS). A pretreatment system is needed to convert COS to H_2S to achieve 99% total sulfur recovery.

In the COS hydrolysis unit, COS reacts with water on the catalyst to produce CO_2 and H_2S. This reaction is slightly exothermic. To prevent catalyst degradation, it is desirable to keep water from condensing in the reactor. The syngas leaving the syngas scrubber at 263°F is saturated with water. A small heater raises the syngas temperature above its dew point to 275°F, which favors the hydrolysis reaction towards the formation of H_2S. The heater duty is about 1.25 MBtu/hr. A typical shell-and-tube heat exchanger heats the syngas by condensing 500 psig steam on the outside of the tubes.

Catalyst requirements were based on scaling the information provided by Süd-Chemie for Subtask 3.2. Detailed stream compositions around the COS hydrolysis reactor are given in Addendum D.

5.2.7.3 Low Temperature Heat Recovery System

With most of the COS being converted to H_2S, the syngas leaves the COS hydrolysis reactor at about 275°F. A three stage cooling combination is employed to cool the syngas. First a boiler feed water (BFW) preheater heats BFW from 150°F to 180°F. The heated BFW then goes to the 500 psig steam drum in the HRSG at the power block to produce saturated 500 psig steam. The syngas leaving the BFW preheater is then cooled from 236°F to 140°F in an air-fin cooler before further cooling to 110°F with cooling water in a shell-and-tube exchanger.

Downstream of the third stage water cooler, a flash drum separates the syngas from the condensate. During cooling, a substantial amount of NH_3 becomes dissolved in the process condensate. The process condensate is routed to the sour water stripper for NH_3 removal. Detailed stream compositions are given in Addendum D.

5.2.7.4 Mercury Removal

During gasification, mercury present in the coal will partition primarily to the syngas stream. The design coal has a mercury content of 0.14 ppmw based on USGS coal analysis data. The mercury concentration in the cooled syngas is about 78 µg/Nm^3, and the mass flow rate is approximately 0.03 lb/hr. The mercury removal system is designed to achieve greater than 90% removal.

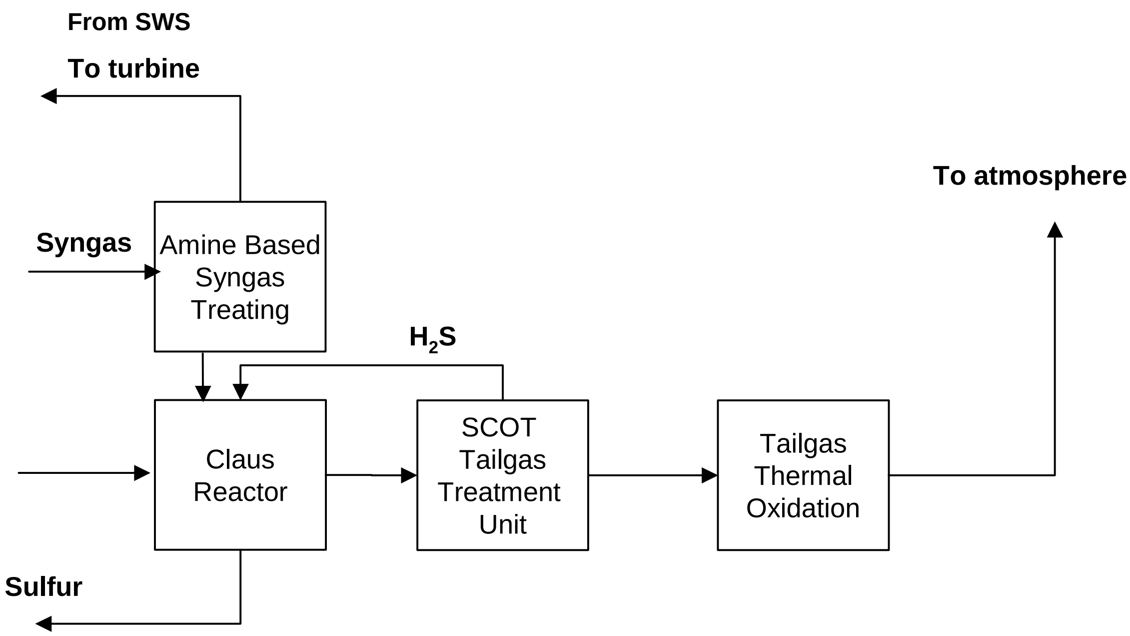
The mercury removal technology is the same as that used in Subtask 3.2, adsorption on sulfur impregnated activated carbon. Details are shown in Addendum B. The expected bed life is approximately 3 to 5 years. This is a commercially proven, reliable process that should exceed 90% mercury removal.

5.2.7.5 Acid Gas Removal and Clean-up

The syngas leaving the mercury removal drum is routed to an acid gas removal system to remove H_2S. Ortloff Engineers, Limited provided the design for this unit.

A gas treatment system features UOP's Selective AGFS (Acid Gas removal - Formulated Solvents) process, which selectively removes most of H_2S, but allows most of the CO_2 and other species to remain in the syngas stream. Figure 5.5 shows the block diagram of the recommended design. The amine based acid gas removal unit consists mainly of an absorber and a regenerator. The treated syngas then flows to the gas turbines.

Figure 5.5 Block Flow Diagram of the Acid Gas Removal System



The acid gas stream leaving the regenerator can be converted either into sulfuric acid or elemental sulfur. Based on demonstrated performance on syngas and on the required scale of production (18.7 tpd), a Claus type of sulfur recovery system was selected. H_2S is converted to elemental sulfur in a conventional multi-stage Claus reactor; the tailgas is routed to a Shell Claus Off-gas Treating (SCOT) process, where residual sulfur compounds are converted back to H_2S, and subsequently captured by an amine system. It is then routed back to the Claus reactor. Note that the sour gas (HCN, CO, CO_2, H_2S, NH_3, etc.) collected from the SWS also is treated in this system to recover any sulfur in the sour water. This results in a very high overall sulfur recovery of 98.9%. The Claus reactor produces 1,557 lb/hr of elemental sulfur that can either be sold as a source of revenue or disposed in a landfill.

The treated gas leaving the SCOT unit then is incinerated in a tailgas thermal oxidation (TTO) unit before being released to the atmosphere. Natural gas is used in the TTO to incinerate the effluent, and a waste heat recovery system is included in the TTO to generate high pressure and low pressure steam. If natural gas is not available, LPG may be substituted. These steams, along with the steam generated in the Claus reactor, are used in the reboiler of the amine stripper. The vent gas is dispersed to the atmosphere at about 550°F to avoid any condensation of SO_2.

5.2.8 Power Block - Gas Turbine, HSRG and Steam Turbine (Unit 900)

5.2.8.1 Introduction

This application gasifies lignite to produce a syngas that is used to generate clean electric power. The basis for this study is one GE 7FB combustion turbine (CT) with a power output of about 211 MW. Steam is generated from the gas turbine exhaust in a heat recovery steam generator (HRSG). The total power output from the power block is about 301 MW. The reheat steam turbine (ST) produces about 90 MW power. Heat integration with the syngas cooling and cleaning systems is accomplished by the HRSG utilizing the saturated steams produced in the syngas heat recovery system, the preheated BFW in the syngas cleaning system, and the HRSG providing the steams needed in the gasification and cleaning processes. The 500 psig steam required for the gasification process is provided through steam extraction from the steam turbine; the 50 psig steam used in the plant is supplied by a combination of stream extraction from the steam turbine and the generation in a low pressure evaporator.

5.2.8.2 Design Basis

The power block is designed around the CT. The exhaust gas exiting the CT is routed through a HRSG. The HRSG is designed such that the following specific process conditions are met:

- Stack temperature remains above the acid-dew point so that condensation and corrosion does not occur within the system
- A reheat steam turbine is to be used to improve efficiency. The steam pressure is 1,000 psig for the high pressure section and 500 psig for the low pressure section. The outlet of the low pressure section is a condenser, which operates at a vacuum pressure of 1.5 psia.
- A portion of 500 psig superheated steam is used for the gasification process.
- 50 psig superheated steam is to be extracted from the 500 psig turbine section for meeting the low pressure steam requirements in the plant.
- A portion of the 1,000 psig superheated steam is used to heat the oxygen exiting the ASU.
- Maximize the utilization of low quality heat by preheating the fuel streams.

5.2.8.3 System Description

Clean, preheated, nitrogen diluted syngas is sent to the combustion turbine at 400°F and 420 psia. Figure D.6 in Addendum D illustrates the CT/HRSG.

The use of syngas produces higher generator output than that from natural gas due to the higher mass flow rate to the turbine. This phenomena is sometimes referred to as the “syngas boost”. A previous Nexant study showed that a similar turbine generates about 210 MW. This is consistent with performance estimates provided by General Electric for the use of syngas in the combustion turbine. See Section 5.2.8.4, Special Considerations for additional discussion of CT modeling and performance.

---

1 Nexant’s 1.3 Next Plant.

The following describes the exhaust gas and water/steam flows for the HRSG train.

Flue Gas Flow – Exhaust gas exiting the CT (at about 1,129°F, 14.25 psia) flows through the parallel 1,000 psig and 500 psig steam superheaters, a 1,000 psig evaporator, a high pressure economizer, an indirect syngas preheater using BFW, a 50 psig steam evaporator, a second high pressure economizer, a low pressure economizer, and then out through the stack (at about 254°F, 13.94 psia).

Water/Steam Flow – Boiler feedwater (BFW) from offsites enters the low pressure economizer at 150°F and 75 psia. The heated water then flows to the low pressure (LP) steam generator, which operates at 50 psig. About 130 klb/hr of LP steam is generated and combined with extraction steam from the steam turbine to meet various process demands throughout the plant.

Cooling water return from the ash cooling screws are combined and enter the first of two high pressure economizers at 181°F and 1050 psia. The heated water then flows to the second economizer and the to the high pressure (HP) steam generator which operates at 1000 psig. The evaporator generates about 440 klb/hr of 1,000 psig saturated steam. Approximately 2.5% of the inlet water mass flow is blowdown from the system. The saturated steam from the evaporator is mixed with the 1,000 psig saturated steam from the tube fired boiler of the high temperature heat recovery area. The mixed saturated steam then goes to the HP superheater to produce approximately 608 klb/hr of 1,000 psig superheated steam at about 1050°F.

Medium pressure (MP) saturated steam is supplied from the gasifier section and combined with 500 psig steam from the first stage of the steam turbine. Some 500 psig steam is sent to the gasifier (65,000 lb/hr) and the rest is sent to the MP superheater where it is heated to 1050°F. The reheated 500 psig steam is sent to the second stage of the steam turbine.

13,790 lb/hr of 1,000 psig superheated steam from the HRSG is sent to the oxygen heaters in the ASU unit. High-pressure steam was selected to provide the heat due to the high temperature required at the gasifier (i.e., 590°F). The remaining 1,000 psig steam is sent to the single steam turbine for power generation. 21 klb/hr of steam are extracted from the steam turbine at 75 psig and 614°F. This extracted stream is let down to 50 psig and combined with the 50 psig steam from the LP steam generator as noted previously.

The syngas is indirectly preheated through the HRSG. To avoid the danger of a tube rupture where syngas may leak into an oxygen-rich stream (11-13%), boiler feed water (BFW) is heated in the HRSG and then cooled by preheating the syngas to 400°F outside the HRSG. The BFW is heated in a coil within the HRSG between the high pressure economizer and the low pressure steam evaporator.

5.2.8.4 Special Considerations

The firing of low- to medium-Btu coal derived syngas in a combustion turbine results in performance that is different than the combustion of natural gas or syngas from non-coal feeds. The only reliable way to estimate turbine performance is to have the manufacturer conduct a performance test on the design syngas. A GateCycle simulation of the turbine was performed, using turbine performance based on the performance reported in a previous Nexant study. A more detailed discussion of the modeling of the power block is included in Addendum A.

The steam turbine employed is a reheat two-pressure steam turbine, with the high pressure inlet at 1,000 psig and the inlet for the low pressure section at 500 psig. The steam exit pressure is about 1.5 psia, and the steam is dry at the exit. The 1,000 psig saturated steam produced in the syngas cooling section merges with the 1,000 psig saturated steam generated in the high pressure evaporator before entering the high pressure superheater. A portion (13,790 lb/hr) of the superheated 1,000 psig steam is used for heating the oxygen stream exiting the air separation unit.

The steam exits the high-pressure section at about 550 psig; it then merges with the saturated steam produced in the gasification process and the 500 psig evaporator. A portion (65,000 lb/hr) of this steam is routed to the gasifier, and the remainder is superheated in the 500 psig superheater before it is routed to the low pressure section of the steam turbine. 75 psig steam is extracted from the turbine at a rate of 21 klb/hr. This steam is mixed with the saturated 50 psig steam produced in the low pressure evaporator and it supplies the needs of low pressure steam in the plant.

Many IGCC designs employ the use of air extraction from the CT compressor as the initial stages of compression for the air to the oxygen plant. This reduces the size of compression equipment required for the plant and can lower capital and operating costs. This option was not considered in this case because it adds increased plant complexity and poses integration issues that were considered too complex for this level of study.

5.2.8.5 Results/Conclusions

Table 5.3 summarizes the net performance of the power block. The steam turbine is powered by 1,000 psig steam. Superheated 500 psig steam and 50 psig steams also are generated to satisfy the internal demands of the facility.

Table 5.3 Performance of the Power Block

CT Power Output	210.78 MW
ST Power Output	90.84 MW
Total Power Output	**301.62 MW**
500 psig Superheated Steam to the Gasifier & COS preheater (E-701)	65,000 lb/hr
1,000 psig Superheated Steam to Oxygen Heater	13,790 lb/hr
50 psig Steam for Plant Usage	151,000 lb/hr

5.2.9 Sour Water Stripper (Unit 800)

5.2.9.1 Introduction

The sour water treatment unit treats 424,171 lb/hr (~848 gallons/minute) of sour water. Figure D.4 in Addendum D includes the sour water treatment unit. Part of the treated water is recycled back to the syngas scrubber, and the remainder is discharged to the waste water treatment plant.

5.2.9.2 Design Basis

The stripping column produces a liquid effluent stream containing no more than 50 ppmw ammonia and less than 10 ppmw hydrogen sulfide.[2] The resulting stream is cooled to 110°F before either being recycled to the syngas scrubber column or discharged to the waste water treatment plant.

5.2.9.3 System Description

Sour water from the syngas water scrubber is mixed with process condensate, cooled to 200°F via heat exchange with air used for lignite drying, and flashed at 35 psia in the 3-phase separator. The 3-phase separator divides the vapors, oils and water contained in the sour water into discrete streams. The water stream is processed in the sour water stripper (SWS). The vapors are sent to the sulfur recovery unit along with the SWS overheads. The oil stream is sent to the slop system, where, for example, it can be sold to a local refinery for further processing.

Water exiting the 3-phase separator is normally pumped through a stripper feed preheater prior to the distillation column. The preheater is a shell and tube heat exchanger with the sour water on the tube side (cold side) and the warmer stripper bottoms product on the shell side (hot side). The preheated liquid is fed to a distillation column with a partial condenser and kettle reboiler. The condenser is air cooled, while the reboiler is heated by superheated 50 psig steam. The overhead vapors from the stripper column and the vapors from the upstream flash drum are mixed and sent to the acid gas removal system. The bottoms product exiting the stripper feed preheater is cooled to 140°F by an air-fin cooler, and then further cooled to 110°F with cooling water. A portion of the cooled product water stream is sent to the wastewater treatment plant to prevent the buildup of any non-volatile impurities within the system, and the remainder is recycled back to the syngas water scrubber. The specific design information and simulation results are included in Addendum D (Figure D.4 and Table D.3).

If a disturbance or upset occurs in the sour water system, the sour water can be diverted to a day tank. This offline storage tank is provided in place of a spare stripper. This tank has sufficient storage capacity to account for a one day outage of the sour water system.

---

[2] Kohl, A. and Nielsen, R; *Gas Purification – Fifth Edition*, Gulf Publishing Company, 1997

5.2.9.4 Special Considerations

The day tank is designed for atmospheric pressure to avoid pressurized vessels and to reduce the cost. The incoming sour water is cooled to condense all of the H_2S. This method is used to avoid the need for a compressor to send the corrosive H_2S vapor to the flare. The cooler reduces the vapor in the tank leaving only a small amount of non-condensables, which are sent to the flare.

Because of the corrosive nature of hydrogen sulfide, stainless steel or stainless steel cladding is required. Additional design considerations included minimizing the water content of the vapor stream sent to the sulfur plant. It is recommended that the water vapor not exceed 5 percent of the gas stream entering the acid gas stripper column.

Past experience with lower temperature fluidized-bed gasification systems have demonstrated the presence of various light oils, including water-soluble phenols, in the raw syngas.[3,4] The U-GAS® technology has been characterized as having a very low light oil content.[5] However, recent data on U-GAS® light oil generation for the specific lignite feed is lacking, and pilot scale testing should be conducted to gather design data and to determine the solubility of low-level oils and trace organics. Furthermore, research is recommended to determine the fate of such organics, primarily for the more soluble compounds that may end up in the water scrubber discharge and process condensate when processed in aerobic treatment processes.

For this study, preliminary modeling of the expected organic content in the syngas suggests that traditional aerobic wastewater treatment would be the most effective technology for their destruction.

5.2.9.5 Results/Conclusions

The sour water feed rate is about 424,171 lb/hr (~848 gallons/minute). The resulting stripper column consists of 21 stages (including the condenser). It is constructed of carbon steel clad with stainless steel and has stainless steel internals. The sour water flash drum, distillate reflux drum, and overhead condenser are constructed of carbon steel with stainless steel cladding. Stainless steel also is used for the tube side of the stripper feed preheater, but the shell is carbon steel. The sour water cooler is an air-fin design, with the surfaces that contact the sour water being stainless steel. The product recycle water has a design ammonia concentration of less than 50 ppmw, and the H_2S and COS concentrations each are 1 ppmw or less. The equipments specifications are provided in Addendum B.

---

3 Probstein, R. F. and Gold, H.; *Water in Synthetic Fuel Production*, Massachusetts Institute of Technology, 1978

4 *Advanced Techniques in Synthetic Fuels Analysis*, Proceedings of Chemical Characterization of Hazardous Substances in Synfuels, Seattle, Washington, November 2-4, 1981

5 Clarke, L.B.; *Management of By-Products from IGCC Power Generation,* IEA Coal Research, May 1991

5.2.10 Offsites/Utilities (Unit 1000)

5.2.10.1 Steam System

Three levels of steam are provided in the onsite facilities. High pressure (1,000 psig), medium pressure (500 psig), and low pressure (50 psig) steams are generated onsite. During normal operation no additional steam generation facilities are required in the outside battery limits (OSBL). A start-up boiler will be required to produce 60,000 pounds per hour at 1,000 psig, superheated to 650°F. The overall steam balance is shown in Table 5.4.

Table 5.4 Overall Steam Balance

(thousand lb/hr)

High Pressure (1,000 psig/650°F) Steam		
	Production	776.7
	Consumption	776.7
Medium Pressure (500 psig/570°F) Steam		
	Production	621.8
	Consumption	621.8
Low Pressure (50 psig/350°F) Steam		
	Production	151.0
	Consumption	151.0

5.2.10.2 Condensate Collection System

Equipment is incorporated to collect the steam condensate from the high pressure, medium pressure and low pressure steam users and return it to the steam generation system. Provisions are made for detection of contamination of condensate to eliminate the source of contamination when detected. Provisions also are made to completely dump the total condensate in case of contamination of the total return condensate. In this case, the gasification units will not be able sustain continuous normal operation, and a shutdown will be initiated. A deaerator is included to handle all condensate.

The system handles condensate at three pressures: 50, 500 and 1,000 psig. The condensate flow rates are shown in Table 5.5.

Table 5.5 Condensate Basis

(thousand lb/hr)

50 psig condensate flow rate	106.0
500 psig condensate flow rate	537.7
1,000 psig condensate flow rate	13.8
Total condensate flow rate	657.5
BFW requirement	1,446

5.2.10.3 Demineralized Water System

The estimated quantity of Demineralized Water (DMW) for the facility is approximately 867 gpm (average) and 1,200 gpm maximum. Make-up to the steam generation system constitutes the majority of this requirement. Significant additional DMW will be required when steam is injected into flare system for smokeless burning. This is a short time requirement. To meet the DMW requirement of the facility, a DMW system with a delivery capacity of 1,200 gpm is proposed.

As stated above, in the event of contamination of the condensate system, the whole quantity of the recovered condensate will be dumped. To maintain a continuous operation of the onsite generation facilities until a safe shutdown can occur, DMW needs to be supplied to the steam generation system. With this in view, the DMW tank will be sized to provide 10 hours or 720,000 gallons of hold-up when the condensate system is contaminated. This storage also will cover short term smokeless flaring. The demineralized water basis is shown in Table 5.6.

Table 5.6 Demineralized Water System

DMW Delivery Capacity	1,200 gpm
DMW Average Delivery Capacity	867 gpm
DMW Maximum Delivery Capacity	1,200 gpm
DMW tanks for 8-10 hours	720,000 gal

5.2.10.4 Cooling Water System

The cooling water system is designed to continuously circulate cooling water through various heat exchangers in the facility. The heat absorbed from the heat exchangers by the cooling water is discharged to the atmosphere at the cooling tower. Cooling tower water is circulated through the heat exchangers by the cooling water circulation pumps. The water lost from the cooling tower by evaporation, windage, and blowdown is made up by the addition of make-up water to the cooling tower basin. The chemical water treatment system for the cooling tower will be leased from standard third party suppliers. The cooling water basis is shown in Table 5.7.

Table 5.7 Cooling Water System

Supply Temperature	80°F
Return Temperature	110°F
Pump Discharge Pressure	50 psig
Cooling Water, normal flow rate	49,300 gpm
Cooling Water, maximum flow rate	62,000 gpm

The cooling water requirement for the facility is estimated to be 49,300 gpm (average) and 62,000 gpm (maximum). A cooling water system with a capacity of 62,000 gpm and a supply temperature of 80°F (return temperature of 110°F) will be provided for the facility. The chemical water treatment system for the cooling tower will be leased from standard third party suppliers.

5.2.10.5 Safety Shower / Eye Wash System

The safety shower and eye wash system will consist of a safety shower water tank, pump, and a heater/cooler to keep the water in circulation at about 70°F temperature. A booster pump is required to maintain circulation.

5.2.10.6 Raw Water / Fire Water System

The raw water system receives raw water from the river and stores it in the raw water/fire water storage tanks located in the plant. The raw water is stored in a filtered water tank large enough to hold one day's raw water requirement, and the fire water is stored in a fire water storage tank large enough to hold four hours of fire water.

Adequate water quantity is assumed to be available at this site. Average requirement of raw water for the facility is estimated to be 2.46 million gal/day.

The raw water requirement constitutes of:

- Cooling tower make-up
- Demineralized water system feed
- Miscellaneous uses

The raw water tank will be of adequate capacity to provide storage for 4 hours of firewater plus 1 day's supply of raw water requirement for the facility. The raw water / fire water basis is shown in Table 5.8.

Table 5.8 Raw Water / Fire Water Basis

Raw Water consumption	2,464,700 gal/day
Raw Water (Filtered Water Tank) capacity	2,464,700 gal
Firewater consumption in 4 hours	850,000 gal
Firewater Storage Tank capacity	850,000 gal
Cooling Tower Makeup rate	1,850 gpm
Raw Water and Fire Water requirements	3,314,700 gal
Firewater Pump rate	2,500 gpm

5.2.10.7 Drinking (Potable) Water System

Drinking water for the facility will be obtained directly from the main city water header (tapped off the public utility header) to the facility.

The function of the potable water system is to distribute potable water (supplied by the city) to various areas inside the industrial gasification site. The potable water system is adequate to provide a continuous and sufficient quantity to the plant for bathroom facilities, drinking fountains, emergency shower and eye wash stations, and various sinks (lab, maintenance, control room). The potable water provided by the city is estimated at 7.5 gpm.

Potable water shall be supplied to the following areas:

- Administration building
- Gasifier building
- Maintenance building
- Plant offices, laboratory and control room

5.2.10.8 Compressed Air System

A compressor system, dryer, and receiver will supply both the instrument air and service air for the facility. The compressed air system provides oil-free compressed air while maintaining a minimum pressure of 100 psig in the distribution headers. Instrument air requirement for the facility is estimated to be approximately 1,000 scfm. An additional provision of 600 scfm (average) and 1,200 scfm (peak) has been provided in the design for service or plant air. This includes approximately 500 scfm for the coal handling area. Three compressors (2 working and 1 standby) of 800 scfm each will be provided to meet this demand. The compressed air basis is shown in Table 5.9.

Table 5.9 Compressed Air System

Instrument air rate	1,000 scfm
Peak Service or Plant air rate	1,200 scfm
Average Service or Plant air rate	600 scfm
Total design air rate	2,400 scfm
Compressed air pressure	100 psig

5.2.10.9 Natural Gas System

Natural gas will be supplied to the facility from the main natural gas header from outside the complex. Natural gas will be used as fuel for the start-up heater in the gasifier unit, the Claus plant, the coal dryer, the flare and primary fuel for the auxiliary boiler, when required. Two knock out drums will be provided in the system.

5.2.10.10 Flare System

Specific design philosophies and instrumented control systems are usually employed in gasification plant designs to mitigate certain relief scenarios and to reduce the load on the flare. Such detailed design load calculations will be part of basic/detailed engineering for the specific facility. For purposes of this study, the scope of facilities includes a flare with a design capacity of 405 MBtu/hr (the syngas produced by one gasifier if the gas turbine is lost). This will be a steam-assisted flare with natural gas being used as pilot fuel.

The capacity of the flare will have significant impact on the layout of the facility as well as the type and cost of flare system. This is one of the key issues that need to be resolved during the basic engineering for the facility in consultation with the technology suppliers. This decision will be dependent on various factors such as permissible radiation levels at the property fence line, and the owner's design philosophy with respect to use of instrumented control systems for mitigation of relief scenarios.

The flare system consists of an elevated flare. A continuous flare system pilot flame is maintained with natural gas. A knock out pot is provided to remove any liquid entrained in the flare feed stream.

5.2.10.11 Nitrogen System

The nitrogen supply is sub-divided into two independent systems:

- A dedicated nitrogen system for onsite process applications (500 psig)
- A general purpose nitrogen system for all other applications (100 psig)

The two systems will be independently piped from the source. When the ASU is not operating, the nitrogen requirement for the gasification facilities will be provided from a liquid nitrogen tank and evaporators. The liquid nitrogen storage and evaporation system, consisting of the nitrogen unloading facility, liquid nitrogen tank, evaporators, and associated controls will be leased from the liquid nitrogen supplier.

A larger quantity of nitrogen will be required during initial start-up, which will be made available from a road tanker mounted storage vessel and evaporating system. The start-up nitrogen will be routed through the two nitrogen systems explained earlier.

Nitrogen system will be sized to meet the maximum requirement of 78 kscf/hr and a peak requirement of 101 kscf/hr during startup.

Dilution nitrogen for NOx control will be made available from the ASU.

5.2.10.12 Waste Water Collection, Treatment and Disposal System

Waste streams generated in the facility will be collected, conveyed and treated, as necessary, prior to disposal. The following have been considered for this study.

- Non-contaminated surface water

All rainwater falling on non-contaminated areas will be allowed to rundown into storm water drains, which will be connected to the area drainage system by gravity.

- Potentially contaminated waste water system

The contaminated wastewater collection system is an atmospheric sewer system where potentially contaminated surface water and process wastewater will be collected and routed for further handling. Wastewater sumps will be located at the two ends of the gasification units for collection of this wastewater. These sumps will be provided to collect any hydrocarbons that have not already evaporated. Sump pumps will transfer the collected water to the final wastewater disposal sump.

Areas around equipment where surface water can be contaminated by process spills will be curbed. The water from these curbed areas will be routed to one of the two wastewater sumps.

- Demineralizer waste water

The Demineralizer unit will generate acidic and caustic waste during the batch regeneration cycles. This wastewater will be collected in a dedicated sump, batch neutralized and pumped to the final wastewater disposal sump.

- Waste water from flare system

Wastewater from the seal drum in the flare system will be collected in a dedicated sump and pumped to the nearest of the two gasification unit sumps. From there, any collected hydrocarbons can be removed, and the wastewater can be transferred to the final wastewater disposal sump.

- Final waste water disposal sump

Hydrocarbon free waste from the above sumps, cooling water blowdown, and boiler blowdown will be routed to the final wastewater disposal sump. Any final traces of hydrocarbon will be separated, the pH adjusted, and the wastewater discharged to a river.

5.2.10.13 Electrical Distribution

The power delivery system includes the combustion turbine generators, each of which is connected through a generator breaker to its associated main power step-up transformer. The HV switchyard receives the energy from the step-up transformers at 230 kV. Internal power is distributed at 33kV from auxiliary power transformers. 33/13.8 kV transformers will service the major motor loads, such as the air compressors. Several substations will serve the balance of the project loads with 33/4.16 kV transformers supplying double-ended electrical bus.

One standby diesel engine generator will provide power for emergency loads during power outages. The generator shall be located indoors and connected to the emergency switchgear via a transfer switch. The overall plant power production and consumption are shown in Table 5.10.

Table 5.10 Electrical Basis

Gross Power Production	301.6 MW
Internal Power Consumption	50.6 MW
Power Export	251.0 MW

5.2.10.14 Miscellaneous

Interconnecting Piping

This system consists of various lines that transport liquid or gas streams within the onsite facility and to/from the OSBL areas.

The following lines are in the interconnecting piping system:

- Steam Generation System
- Condensate Collection System
- Demineralized Water System
- Cooling Water System
- Safety Shower / Eye Wash System
- Raw Water / Fire Water System
- Drinking Potable Water System
- Compressed Air System
- Natural Gas System
- Flare System
- Nitrogen System

Pipe Racks

Steel pipe racks in the ISBL and OSBL areas are included in the scope of facilities. Pipe Racks in the ISBL area and in other areas containing flammables will be fire proofed to meet local regulations.

Roads

Adequate roads to suit the plant layout are incorporated in the facility.

Site Development

The site is a reasonably flat piece of land. In the absence of a survey map for the proposed plot, a provision for site development only has been included in the estimate.

The flood level has not been established at the proposed site. It is assumed that the plot is above flood level, and no provisions have been made in the estimate in this regard.

Miscellaneous Works

Other miscellaneous works in this category include equipment foundations and the wastewater collection sumps.

Buildings

The following buildings and structures are included in the scope of facilities. All buildings are to be heated.

- Administration Building

An administration building housing all administrative personnel for the facility is included in the scope as requested by GTI. This building will be concrete and brick construction and air-conditioned.

- Gasification System Building

The gasification system (reactor, lock hoppers, cyclones, etc.) shall be enclosed within a building. (This will be about seven (7) stories high based on the pilot plant design.)

- Gate House

A gate house to locate the security personnel at the entrance of the facility is included in the scope of facility. This building will be of concrete and brick construction and air-conditioned.

- Canteen and Locker Room

A canteen and locker room is included in the scope. The building will be of concrete and brick construction and air-conditioned.

- Utility Facilities Room

A utility facilities room housing the compressed air system, utility operations room, and the firewater pump house is included in the scope. The building will be of concrete and brick construction and air-conditioned.

- Maintenance Building

A maintenance building housing a maintenance shop, maintenance offices, and a warehouse is included in the scope of facility. This building will incorporate a fire station, which will house the office for the emergency fire crew and parking for a fire tender. The building will be of concrete and brick construction. Only the offices will be air-conditioned.

- Plant Offices, Laboratory and Control Room

The control room, plant offices and laboratory are located in a single building. The building will be of brick and concrete construction and air-conditioned.

- Electrical Building

An electrical building containing the substation equipment and some of the electrical equipment catering the utility sections is included in the scope of facility. The building will be concrete and brick construction and will have a mezzanine floor for routing the cable. The building will be air-conditioned.

5.3 EMISSIONS

Gasification systems are inherently less polluting than combustion systems because the pollutants (sulfur, mercury, chlorine, and others) are removed from the syngas before it is sent to the combustion turbine. Pollutant controls in combustion systems generally are add-on processes that treat the flue gas prior to discharge to the atmosphere. Because these systems treat a large volume of gas at low pressure, they generally are expensive. Whereas, gasification systems treat a smaller amount of gas at higher pressure, and therefore are smaller and less expensive systems.

The following sections detail the emissions characteristics of the gasification facility.

5.3.1 Sulfur

Sulfur is removed from the syngas by a two-step process. First the syngas is passed over a COS hydrolysis catalyst at 275°F to convert the COS to H_2S. The H_2S is removed from the syngas by UOP's Selective AGFS process that was designed by Ortloff Engineers, Ltd. This combination has a sulfur removal rate of 99.0%. Sulfur is recovered as elemental sulfur by a Claus process with a Shell Claus Off-gas Treating (SCOT) process, where the residual sulfur compounds are converted back to H_2S and subsequently captured. The combined SO_2 release rate from the gas turbine, lignite dryer and the incinerator is 36.5 lb/hr or 0.016 lb per MBtu (HHV) of energy input. The net result of this processing scheme is an overall sulfur removal rate of 98.9%.

5.3.2 NOx and CO

The firing of combustion turbine on coal-derived syngas requires the proper design of turbine components. The specific design can influence the emission rates of NOx and CO. For this application, one GE 7FB combustion turbine is used. GE currently estimates NOx emission levels for this application are about 15 ppmvd @ 15% O_2 (0.06 lb/MBtu) when steam is used as diluent. For the current syngas heating value, using nitrogen instead of steam, according to GE literature, will increase the NOx emissions by about 100%, or 30 ppmvd @ 15% O_2 (0.12 lb/MBtu). Including balance of plant emissions (e.g., incinerator off gas and flare), total facility CO emissions are estimated at less than 0.05 lb/MBtu.

Carbon monoxide emissions result from incomplete combustion of carbon based fuels and are primarily a result of highway and off road transportation sources. CO emissions are not regulated in the New Source Performance Standards for utility boilers and combustion turbines. Because CO can be a potential issue for any combustion source, it is possible that emissions may be regulated on a site specific basis as part of the facilities operating permit. Potential sources of CO from IGCC systems include exhaust from the gas turbine, tail gas incinerator of the sulfur recovery unit, the flare system, coal drying, and possible fugitive emissions from equipment leaks.

While specific emission levels for the application under study would be site specific, for comparison carbon monoxide emission limits included in the operating permits for TECO Polk Power Station and the Wabash River Repowering Project were 0.392 lb/MW-hr to 2.2 lb/MW-hr respectively. Operating experience at the Wabash facility has resulted in CO emissions well below the permitted levels. More recent PSD permitting experience for a proposed ConocoPhillips (formerly Global Energy) IGCC plant included an emission limit of 0.19 lb/MW-hr.

5.3.3 Mercury

Mercury emissions for larger coal-fired electric generators are now starting to be regulated. For other sources, mercury emissions are regulated as a hazardous air pollutant and require maximum achievable control technologies. In anticipation of stringent mercury removal requirements, the technology selected for this study was designed to achieve 90+% mercury removal. Mercury emissions leaving the stack are estimated at 0.00039 lb/hr (0.95 lb/TBtu). Mercury emissions of this rate are equivalent to a stack gas concentration of around 1 µg/Nm^3, which approaches the detection limit of current mercury measurement technologies.

5.3.4 Water

From the work recently completed for Subtask 3.2 (see Section 5.3, Emissions), the preliminary modeling of the organic content of the syngas suggests that traditional aerobic wastewater treatment would be the most effective technology for destruction of trace organic compounds of the expected types.

5.3.5 Other Emissions

Particulate emissions are considered to be negligible. Particulates are removed from the syngas either by filters or scrubbing. Emissions from fugitive dust during the coal handling, drying and other operations will be typical of other coal handling facilities and have not been estimated.

Current emission control systems do not typically address chlorine emissions. These typically are uncontrolled from coal combustion systems. Stack gas scrubbing reduces chlorine emissions to some extent. In a coal gasification system, essentially all chlorine is removed during the gas cleaning steps.

5.4 TRADE-OFF STUDIES

Many of the trade-off studies that were examined as part of the Subtask 3.2 air and oxygen cases are applicable to this oxygen-blown lignite case. These trade-off studies have been discussed previously in Section 5.4 of the Topical Report for Subtask 3.2.[6] The improvements that were made as a result of these studies are included in this design. However, two trade-off studies were specifically performed for this case, sulfur removal by the CrystaSulf®[7] process compared to the conventional amine followed by a Claus unit, and the inclusion of a spare gasification reactor train. Beyond this, no additional trade-off studies were made for this design.

5.4.1 Sulfur Removal

The CrystaSulf process is potentially attractive because it has lower capital costs and is a simpler process than the amine/Claus system since it has fewer subsystems. When compared to a similar sized traditional amine/Claus system, CrystaSulf has lower steam and cooling requirements and consumes no natural gas. However, the CrystaSulf process has significantly greater annual consumable costs and an approximately 60% greater parasitic power requirement than a comparable amine system.

The economics of the two systems were compared over the life of the project. Although the CrystaSulf process has lower capital and utility costs, the annual consumable costs are substantial compared to the amine system. Annualized pre-tax costs (using the assumptions described for the overall plant economic analysis) for capital and operating expenses of the CrystaSulf process are more than 20% greater than the baseline amine system. However, it should be noted that the economics of the comparison are very sensitive to certain parameters, particularly on-stream factor, natural gas price, and sulfur capture. At lower on-stream factors, lower sulfur production, or higher natural gas prices, the gap between the CrystaSulf process and the amine system narrows, and at some point the CrystaSulf process becomes more economically attractive. It should be noted that the CrystaSulf system allows for much deeper sulfur recovery, achieving approximately 99.9% removal or <4 ppmv H_2S concentration in the treated syngas. The amine system is only designed to achieve 99% removal, with a treated syngas H_2S concentration of approximately 31 ppmv. Because 99% removal satisfies the sulfur removal constraints of the project, the traditional amine/Claus system was determined to be more economical and reliable for this application. (Also see Section 7.3 of this report for a discussion of other technologies.)

---

6 "Topical Report – Subtask 3.2, Preliminary Design for Eastern Coal Case," United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, November 2004.

7 CrystaSulf is a service mark of CrystaTech, Inc.

5.4.2 Spare Gasification Train

This trade off study evaluated the effect of the addition of a complete spare gasification train (units 200 through 650) on the project economics as a result of the increased annual availability. Table 5.11 compares the design rates, daily average input and output rates, and plant economics for a one and two train gasification plant feeding two GE 7FB combustion turbines. Both cases have the same design rates. It is expected that the spare gasifier would be maintained in a “warm state” (e.g., 500°F) and would take between 12 and 16 hours to bring on-line if a problem suddenly occurred with the operating gasifier.

The two train case improves the return on investment by one-tenth of a percent and shows an improved NPV even though the plant capital increases. The payout for a spare gasifier is 4.8 years. In addition reliable operation for a base load power generation facility is an important feature to an operator. The spare gasifier case increases the availability with scheduled maintenance from 80.4% to 87.3%. Considering both these pieces of information, a spare gasifier train is included in the design for this subtask.

Table 5.11 Spare Gasifier Trade-off Economics

	Design	**One Train Plant (w/o spare)**	**Two Train Plant (w/ spare)**
Feed		Daily Average Rate	
Lignite Feed, moisture-free tpd	2,558	2,058	2,234
Products			
Export Power, MW	251.1	201.9	219.1
Sulfur, lb/hr	1,557	1269	1377
Ash, lb/hr	23,116	19,089	20,720
Economics			
Plant EPC Cost, M$		385.5	410.5
ROI, %		19.3	19.4
NPV, M$		160.8	175.6

6.1 INTRODUCTION

Common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period, all are dependent on the project cash flow, and the cash flow is dependent upon the annual plant inputs and outputs. Although the design capacity is the major factor influencing the annual production, other factors that influence it include scheduled maintenance, forced outages, equipment reliability, and redundancy. These other factors must be considered in order to develop a meaningful financial analysis. An availability analysis that considers all of the above factors must be performed to predict the annual production rates. Based on these annual production rates, appropriate annual revenue streams can be developed for the financial analysis.

6.2 PLANT COST

6.2.1 Basis

A process plant can be viewed as consisting of two types of facilities. The first is the manufacturing area, containing all process equipment needed to convert the raw materials (e.g., coal) into the product (e.g., electric power and steam). The capital cost of these facilities are commonly referred to as the inside battery limits (ISBL). For this project the ISBL areas consists of Units 100-900. The second group of facilities contains the outside battery limits (OSBL) or offsites (i.e., Unit 1000). These include general utilities (e.g., instrument and utility air, nitrogen, fire water), buildings (administration, warehouse, etc.), cooling water system, electrical distribution systems, waste disposal facilities, etc. In addition to the plant capital, the owner usually has other costs associated with a project such as interest during construction, (IDC), working capital, project management, startup, etc.

For this evaluation all the investment costs are for the second quarter 2004 at a generic North Dakota site. The labor rates associated with plant construction have been adjusted for the labor rates and productivity in North Dakota.

6.2.2 Methodology

The initial basis for the Subtask 3.4 design required that the coal be dried to 10% moisture content and used two parallel gasification and gas turbine trains. The preliminary results showed a relatively low net electrical efficiency. The basis was reconsidered with items critical to increasing the efficiency and improving the design reexamined. The material and energy (M&E) balances were recalculated using GateCycle and ASPEN except for the sour water stripper, which was simply scaled from the initial design. Utilities that were not developed using the simulation tools were scaled based on throughput. Cost estimates were prorated based on throughput from the initial design using an exponent of 0.65 for all equipment.

6.2.2.1 Equipment Design

The equipment for the lignite-fueled IGCC case was designed using the material and energy (M&E) balances developed specifically for Subtask 3.4. Various groups developed the M&E balances. Raymond Professional Group (RPG) developed the coal handling and preparation area. GTI developed the gasification island. Ortloff Engineers, Ltd. (Ortloff) prepared the sulfur removal and recovery systems. Nexant and NETL developed the remainder of the ISBL facilities and the balance of the plant (BOP).

RPG and GTI provided process flow diagrams (PFDs) for their portion of the study. The BOP process flow diagrams were developed using the ASPEN and GateCycle computer simulations and previous experience with similar systems. The M&E balances and PFDs are given in Addendum D.

Using the M&E balances and PFDs established the operating and design conditions for the individual pieces of equipment. The equipment was then sized and materials were selected to provide a 20-year life. RPG provided the equipment list and sizing for Unit 100, lignite handling and drying. Air Products provided the design and cost for the ASU, Unit 150. The equipment sizing for Units 200-500 was prepared by GTI. The design for the equipment in Units 600 through 1000 (excluding the sulfur removal and recovery system) was prepared by Nexant and NETL using the ASPEN and GateCycle heat and material balances as a basis. Ortloff provided the sulfur removal and recovery system design. The equipment list is provided in Addendum B.

6.2.2.2 Cost Estimating

The total erected cost estimates were prepared in a variety of ways. The first approach was to estimate the cost of the purchased equipment either through vendor quotes or cost estimating software (e.g., Price and Delivery Quoting Service for Chemical Process Equipment, PDQ$®); use an appropriate installation factor to determine the field labor, piping, foundations, electrical, etc. costs for each individual piece of equipment; factor in the cost of instrumentation; and add 55% to the labor portion for indirect labor cost in North Dakota to determine the total erected cost for each individual piece of equipment. This method is well founded theoretically and in practice and has been in use for many years in petroleum and chemical process industries for plant cost estimating. The method relies on the observation that the total installed cost of major equipment items can be reliably represented as a multiple of the equipment cost. For a given type of equipment, the multiplier (called the installed cost factor) can vary depending on the size of the equipment item, specific process design details, site location, and other factors. Factors for the installation of various chemical and refinery equipment (e.g., pumps, pressure vessels, shell-and-tube exchangers) are readily available in the literature. This method was employed for the gas cooling, gas cleaning, and sour water stripper units.

The second approach was to determine the overall installation factor for a unit based on previous cost estimates for similar facilities. The equipment was sized, and the purchased cost was determined either through vendor quotes or cost estimating software. For the solids handling and gasification equipment, which are outside the realm of normal chemical and refinery equipment, an overall unit factor based on previous estimates for similar units was used. Overall unit factors were developed from previous estimates for other sections of the plant as needed. This method was employed for the coal feed, gasification, dust and ash removal systems, and offsites (including buildings).

A third approach was to request quotes for the installed cost of complete units. This method was employed for the coal handling and drying unit (from Raymond Professional Group, RPG), ASU (from Air Products), gas turbine (from General Electric), steam turbine (from Dresser Rand), HRSG (from Vogt Power), Mercury removal (from Calgon Carbon) and the sulfur removal and sulfur recovery units (from Ortloff Engineers, Ltd.).

6.2.3 Results

Table 6.1 shows the EPC (engineering, procurement and construction) cost for the Subtask 3.4 Lignite-Fueled IGCC Power Plant (including a second, spare gasifier). These costs are on a second quarter 2004 basis. The investment is adjusted for labor rates and productivity in North Dakota.

Table 6.1
Capital Cost Summary, Lignite-Fueled IGCC Power Plant
(US$)

Description	Total Project Cost*	Percent of Total
Coal Preparation and Handling	43,258,000	10.5
Air Separation Unit	40,318,000	9.8
Coal Feeding	6,396,000	1.6
Gasification	11,366,000	2.8
Dust Removal	5,765,000	1.4
Ash Removal	8,173,000	2.0
Gas Cooling	8,556,000	2.1
Particulate Removal	9,642.000	2.3
Gas Cleaning	7,375,000	1.8
Sour Water Stripper	5,221,000	1.3
Acid Gas Removal and Sulfur Recovery	15,927,000	3.9
Gas Turbine, HRSG and steam turbine	184,852,000	45.0
Offsites and Auxiliaries	57,026,000	13.9
Buildings	6,589,000	1.6
TOTAL	**410,464,000**	**100.0**

* All plant EPC costs mentioned in this report are second quarter 2004 +30%/-15% cost estimates which exclude contingency, taxes, licensing fees and owners costs (such as land, operating and maintenance equipment, capital spares, operator training and commercial test runs).

6.3 AVAILABILITY ANALYSIS

6.3.1 Background

Common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period, all are dependent on the project cash flow, and the cash flow is dependent upon the annual production. Although the design capacity is the major factor influencing the annual production, other factors that influence it include scheduled maintenance, forced outages, equipment reliability, and redundancy. These other factors must be considered in order to develop a meaningful financial analysis. Thus, an availability analysis that considers all of the above factors must be performed to predict the annual production rates. Based on these annual production rates, appropriate annual revenue streams can be developed for the financial analysis. A detailed discussion of the availability analysis procedures and calculations can be found in Addendum F of the Subtask 3.2 Topical Report. [1]

Availability analyses were performed for the two Subtask 3.4 Lignite IGCC Power Plant designs to account for forced and scheduled outages to determine expected annual revenue and expense cash flows. Based on these cash flows, financial analyses were performed to evaluate the comparative economics of two possible Subtask 3.4 plant configurations; i. e., with and without a spare gasification train.

The effect of sparing (back-up equipment or parallel trains of reduced capacity) can have a significant affect on the capacity factor (availability) of a plant depending upon the amount of spare equipment or parallel trains that are present. Sparing is most effective in increasing the overall plant availability when those portions of the plant with the lowest on-stream factors are replicated. Because reliability is key to the Subtask 3.4 design, sparing played an important role in the design development to provide optimum on-stream capacity while also attempting to maintain economic viability. Availability analysis for the Subtask 3.4 Lignite-Fueled IGCC Power Plant with the spare gasification train calculated an annual average on-stream factor of 87.28% including scheduled outages.

These availability analyses show the importance of designing plants and equipment that have high on-stream factors, require low maintenance (short or infrequent scheduled outages), sparing or replicating those portions which have low on-stream factors, and/or high maintenance periods (long or frequent scheduled outages).

For this analysis, most operations of Subtask 3.4, exclusive of the gasifier island and coal handling, are fundamentally similar to those of the Wabash River Repowering Project. Figure 6.1 represents the block flow of the gasification and the overall power block used for the availability analysis. The figure also illustrates the combination of parallel and series configurations. The availability analysis for Subtask 3.4 is similar to

1 "Topical Report – Task 3: Gasification Alternatives for Industrial Applications, Subtask 3.2: Preliminary Design for Eastern Coal Case," Gasification Plant Cost and Performance Optimization, United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, November 2004.

the analysis in Subtask 3.2 with the following notable exceptions: Subtask 3.4 implements a spare gasification train, candle filters are present in the each gasification train, and a single steam turbine is added to generate more electricity. Make-up of the individual blocks as well as availabilities of the component units is presented in Table 6.2. A more detailed explanation of the availability analysis is included in Addendum F of the Subtask 3.2 report.

Figure 6.1 Availability Block Diagram

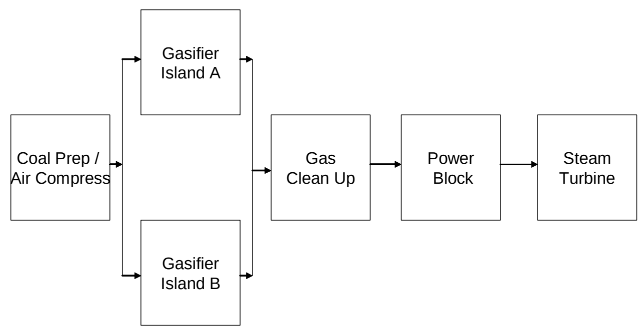


Table 6.2 Availability Estimates

Plant Section	Availability
Coal Prep/Air Compression	
Coal Drying	97%
Coal Crushing	100%
ASU	96.32%
Gasifier Block	
Gasifier Island	97.49%
High Temperature Heat Recovery	97.96%
Candle Filter	98.03%
Medium Temperature Heat Recovery	99.90%
Water Scrubber	99.87%
Gas Clean-up	
COS System	100%
Acid Gas Removal	99.72%
Sulfur Recovery	100%*
Sour Water Treatment	100%
Mercury Removal	100%
Power Block	
Combustion Turbine/Generator	98.19%
Heat Recovery Steam Generator	97.40%
Steam Turbine	99.88%

* Assumes plant operations are not interrupted by short term outages of the sulfur plant because the feed to the sulfur plant can be flared.

6.3.2 Availability Calculations

Table 6.3 presents availability calculations for individual state capabilities; (probability of operating an individual state (e.g., 2 of 2 parallel trains, 1 of 2 parallel trains operating excluding scheduled maintenance)) as well as equivalent availability, both with and without 14 days per year of scheduled outage. A scheduled outage period of 14 days per year was used for this subtask as opposed to 21 days as described in Subtask 3.2 because of the spare gasification train. It is assumed that the spare gasification train will be maintained during the time when it is not running, thus reducing the scheduled outage period by one-third.

Table 6.3 Calculated Availabilities

Coal Prep*		99.48
Syngas Operations** (1 of 2)		99.59
Power Block***		98.19
Steam Turbine		99.88
Availability		
	w/o Scheduled Maintenance	90.76
	w/ Scheduled Maintenance	87.28

Notes:
- * Represents coal drying and crushing operations
- ** Represents solid feeding system through final gas cleaning and includes sulfur recovery and sour water treatment.
- *** Includes combustion turbine, generator, and heat recovery steam generation.

Equivalent availabilities are based on operating states (e.g., number of gasifiers) in operation at a given time) and export power. The two operating states that were considered for this study are presented in Table 6.4.

The availabilities shown in Table 6.4 are a result of weighted equivalent availabilities from Table 6.3. For this study, internal power and steam demands are assumed proportional to syngas generation. Table 6.5 is a summary of the design and daily annual average plant flow rates.

Table 6.4 Operating State Statistics

Syngas Operations*	CT/ HRSG**	Steam Turbine	Net Product Output***	Equivalent Availability†
1 of 2	1 of 1	1 of 1	100%	87.21%
		0 of 1	64%	87.32%

Notes:
- * Represents coal preparation and handling through final gas cleaning and includes sulfur recovery and sour water treatment.
- ** Includes combustion turbine, generator, and heat recovery steam generation.
- *** Represents gross power and steam output minus internal power and steam demands.
- † Represents operations with at least this number of operating units, and includes a scheduled outage of 14 days/year

Table 6.5 Design and Annual Average Flow Rates

	Export Power (MW)	Moisture-free Lignite (lb/hr)	Sulfur (lb/hr)	Bottom Ash (lb/hr)	Fly Ash (lb/hr)
Design	251.04	213,360	1,557	18,399	4,717
w/o Scheduled Maintenance	227.85	193,554	1,432	16,707	4,840
w/ Scheduled Maintenance	219.11	186,130	1,377	16,066	4,654

6.4 FINANCIAL ANALYSIS

The general methodology followed for performing the financial analysis was outlined in Section 3.6. The Nexant developed IGCC Financial Model Version 3.01 was used to obtain the results described in this section. The input parameters for the Lignite-Fueled IGCC Power Plant are given in Addendum C.

The plant EPC cost used in the financial model is that shown in Table 6.1. An owner's contingency fee of 25% was added to the cost of the gasifier islands (Units 200-500), while a contingency fee of 15% was added to the EPC cost of all other plant equipment. Greater uncertainty in the cost of the gasifier justifies the higher fee. This assures that the financial results adequately reflect the additional capital that may be required during plan construction. Based on the cost of the gasification islands and the plant EPC cost, the overall contingency for the entire plant was revised to 15.77% to reflect the higher contingency value of the gasifiers. The plant feed and product rates are adjusted from that given in Section 5.1 to reflect the average availability and actual operating hours of the plant.

"Guaranteed Availability" entered into the financial model refers to plant operations excluding scheduled maintenance outages. Based on the analysis in Section 6.3, the guaranteed availability for 2 gasifier trains was calculated to be 90.76%. This number only gives insight into plant availability for times when the plant is scheduled to operate. The detailed availability analysis calculated the overall yearly availability, which provides the total availability taking into account both scheduled and unscheduled outages. Therefore, the reported availability in Section 6.3.2 of 87.28% is the "Guaranteed Availability" of 90.76% times the percentage of time the plant is scheduled to operate (8,424 hours/year, or 96.16% of the time).

6.4.1 Results

For a Lignite-Fueled IGCC Power Plant with EPC costs of 410.5 M$ and a project life of 20 years, the return on investment (ROI) is expected to be 19.4%, with a net present value (NPV) of 175.6 M$ given a 10% discount factor. Based on the design power output, the EPC estimate is equivalent to a capital cost of 1,635 $/kW. As expected, Subtasks 3.2 and 3.3, the 25 MW subbituminous industrial gasification facilities, have a higher installed cost (2,700-3,100 $/kW) because of the economy of scale disadvantage. However, studies of larger IGCC designs (450 MW) have been able to capture even greater economy of scale benefits, with installed costs of 1,300 to 1,650 $/kW[2]. The installed cost of this case is approaching the cost of the large IGCC facilities by taking advantage of a greater economy of scale. The results point to the possibility that a larger capacity design may be able to reduce installed costs further.

Table 6.6 outlines the rate of return, NPV, payback year, and required electricity selling price to obtain a 12% ROI with all other entries fixed. The ash and sulfur produced in the plant accounts for all additional revenue beyond electricity tariffs. Besides the base case, a “high” and “low” estimate is shown reflecting the current investment cost accuracy assumption of +30/-15%.

Table 6.6 Financial Cost Summary for Three Lignite-Fueled IGCC Power Plant Cases

	Base	Low -15% EPC	High +30% EPC
ROI (%)	19.4	23.6	12.8
NPV (M$) (10% Discount Rate)	175.6	232.6	60.8
Payback Year	2014	2013	2017
Electricity Selling Price for 12% ROI (cents/kWh)	4.7	4.0	5.9

The results reported in Table 6.6 do not take any credits for the environmental benefits gained by the use of IGCC technology. In order to properly compare this design versus other electricity generation technologies using lignite, a side-by-side environmental comparison also should be performed. Quantification of the environmental differences will provide a more level playing field by which alternate technologies can be evaluated. A project developer must consider alternative compliance costs to meet new emission rules versus the cost of the IGCC plant.

For the base case, Table 6.7 breaks down the total plant cost including EPC costs, all fees, start-up costs, and costs occurred from project financing. The “High” and “Low” case costs would be proportionately changed by the percentage difference in EPC costs.

2 Analysis of 4 different IGCC technologies without CO_2 capture, “Gasification Process Selection—Tradeoffs and Ironies”, EPRI, presented at the Gasification Technologies Conference 2004, October 2004.

Table 6.7 Total Plant Cost for the Lignite-Fueled IGCC Power Plant

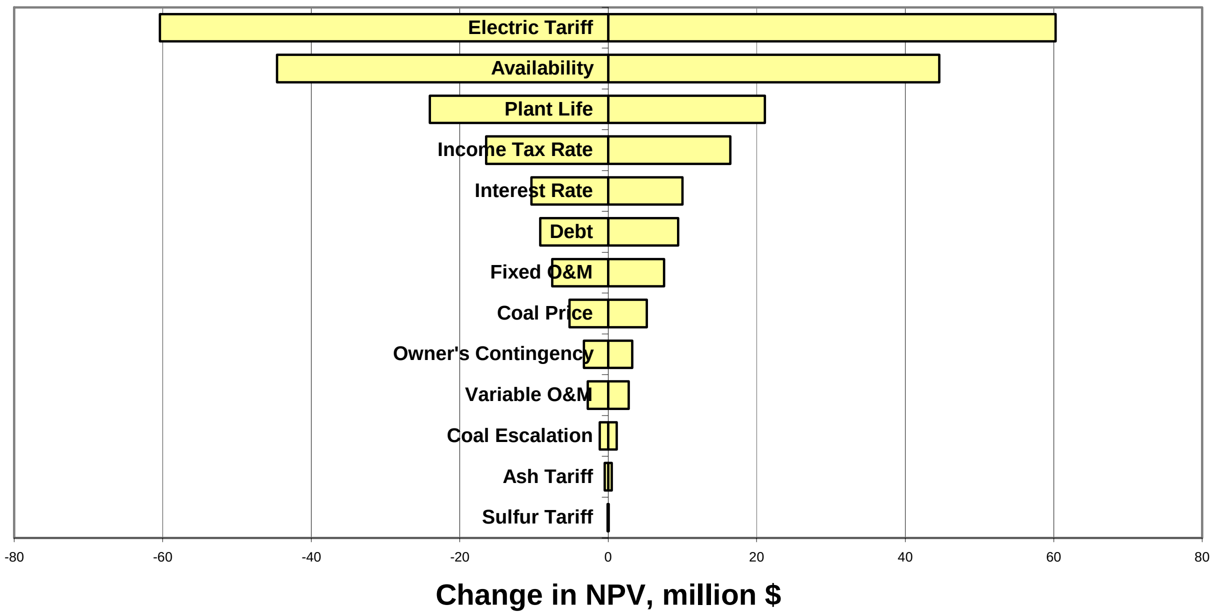
Construction/Project Cost (in Thousand Dollars)		
Capital Costs	**Category**	**Percentage**
EPC Costs	$410,464	70%
Initial Working Capital	$7,309	1%
Owner's Contingency (% of EPC Costs)	$64,730	11%
Development Fee (% of EPC Costs)	$16,419	3%
Start-up (% of EPC Costs)	$8,209	1%
Initial Debt Reserve Fund	$0	0%
Owner's Cost (in thousand dollars) COMBINED WITH DEVELOPMENT AND S/U TO = 10%	$16,419	3%
Additional Capital Cost	$0	0%
Total Capital Costs	$523,550	89%
Financing Costs		
Interest During Construction	$51,774	9%
Financing Fee	$11,391	2%
Additional Financing Cost #1	$0	0%
Additional Financing Cost #2	$0	0%
Total Financing Costs	$63,165	11%
Total Project Cost/Uses of Funds	$586,715	100%
Sources of Funds		
Equity	$199,483	34%
Debt	$387,232	66%
Total Sources of Funds	$586,715	100%

Subtask 3.4 represents a case focused on high reliability and efficiency improvements. Future analysis should be performed to consider potential plant cost savings more closely. Section 7 describes potential areas for technology and design improvements that may be able to reduce the total project cost.

6.4.2 Sensitivities

All financial parameters were varied to determine the project financial sensitivities. Figure 6.2 shows the impacts of selected variables on the NPV at a discount rate of 10%. In all of the cases, the input parameter is varied by ±10%, and the NPV changes from the base case are shown. 10% changes were used to give a common ground by which all variables were evaluated. However, the range of realistic possibilities for each variable could differ significantly. For example, 10% changes in the availability or income tax rate would capture the majority of long-term variations. This would not be the case with variables such as coal price and electricity tariff that could vary by much more than 10%. The relative significance and range of possible values were considered in determining which variables have the most impact on plant economics.

Figure 6.2 Comparisons of a +/-10% Change in Selected Inputs on Project NPV
(Discount Rate = 10%)



The electricity tariff has the greatest impact on the plant net present value; increasing it by 10% increases the net present value by more than 60 M$. In this case, "Electricity Tariff" refers to the sales price of the electricity that the plant generates. This variable was also the most significant in Subtasks 3.2 and 3.3. The significance is more pronounced in this design since, unlike Subtask 3.2, there is no steam export. Also very significant in importance is the availability (annual average on-stream time). By reducing the availability by 10%, the net present value is reduced by more than 45 M$. All other variables associated with the amount of time the plant is operating (operating hours, and plant life) also had a significant impact on plant economics.

The remainder of the input variables impacted the plant economics to a significantly lower extent. It is interesting to note that the interest rate, amount of debt financing, and the plant fixed O&M cost has a greater impact on the economics than in Subtasks 3.2 and 3.3. This is due in large part to the higher EPC cost of Subtask 3.4. Changes in these variables will impact the early cash flow to a greater extent than in the industrial gasification case. Income tax rate also has a greater impact than in Subtasks 3.2 and 3.3 due to the larger positive cash flows throughout the operational life of the project. Coal prices could change fairly significantly without changing the overall economics to a great extent. If the coal price is increased to 12 $/ton (a nearly 30% rise), the NPV is only decreased by 15.3 M$, a 0.8% change in the return on investment.

Figure 6.3 shows the relationship between the electricity tariff and ROI. The model relies most heavily on the electricity tariff for the economic outcome due to electricity being the main product. Even with the relatively low electricity prices that exist in North Dakota, the plant still demonstrates positive economics. If the electric price used for upstate New York in Subtasks 3.2 and 3.3 of nearly 8 cents/kW-hr were applied here, the plant would have a return of over 27%. Regardless of the tariff value assumed, any electricity market could obtain positive returns with this facility, all other plant inputs being equal. The fluctuating marginal prices for electricity from other feedstocks make coal based gasification a competitive option.

Figure 6.3 Effect of Electricity Tariff on Investment Return

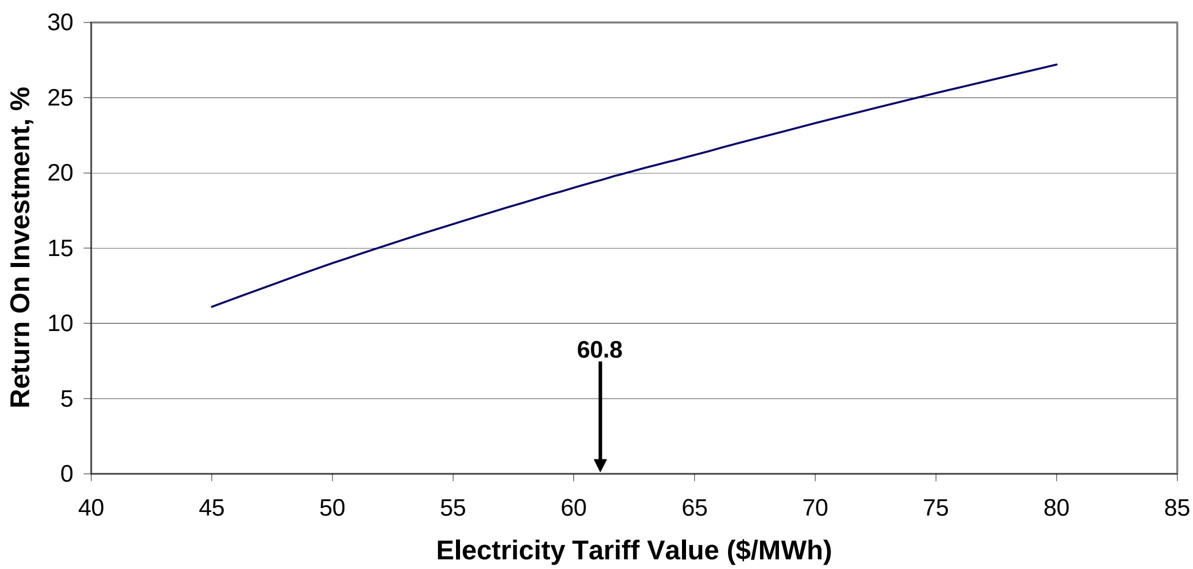
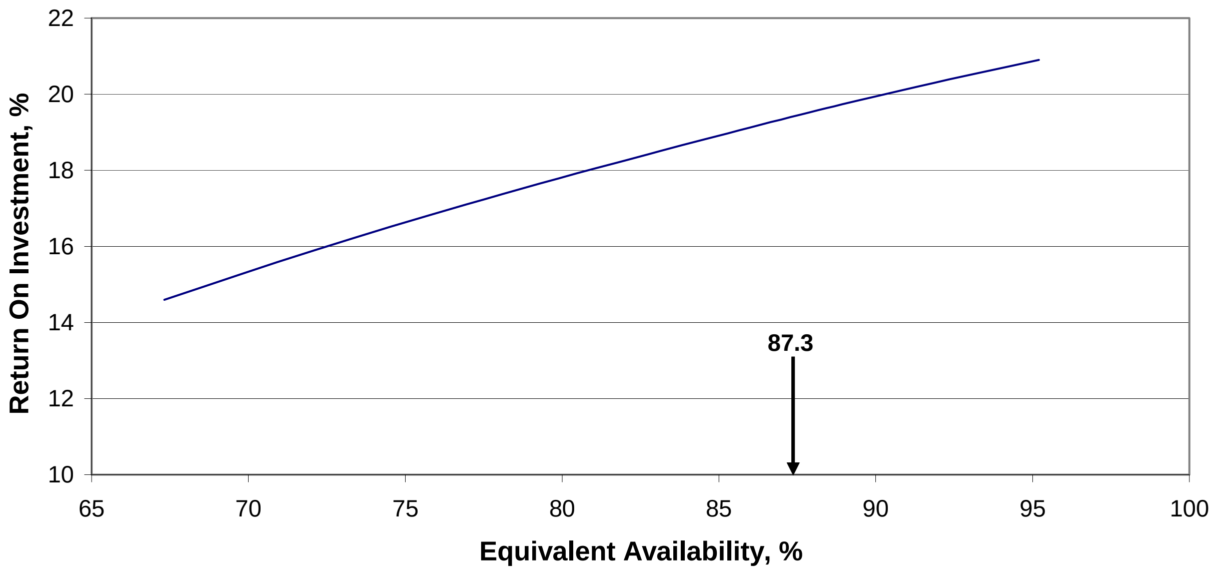


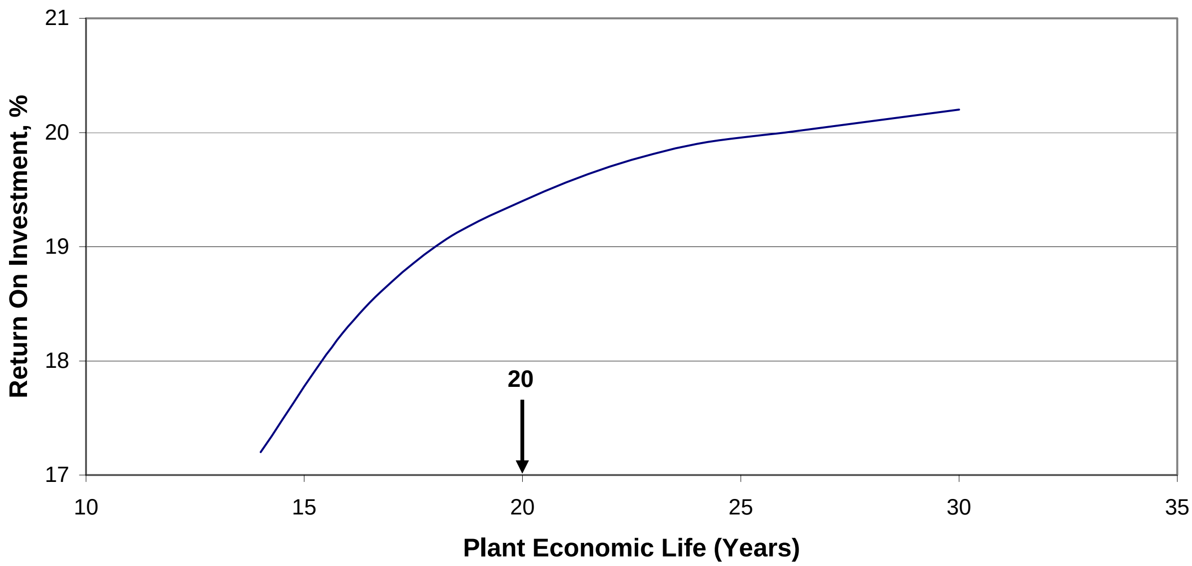
Figure 6.4 shows the relationship that varying the guaranteed availability has on the NPV assuming a 10% discount rate.

Figure 6.4 Effect of Availability on Investment Return



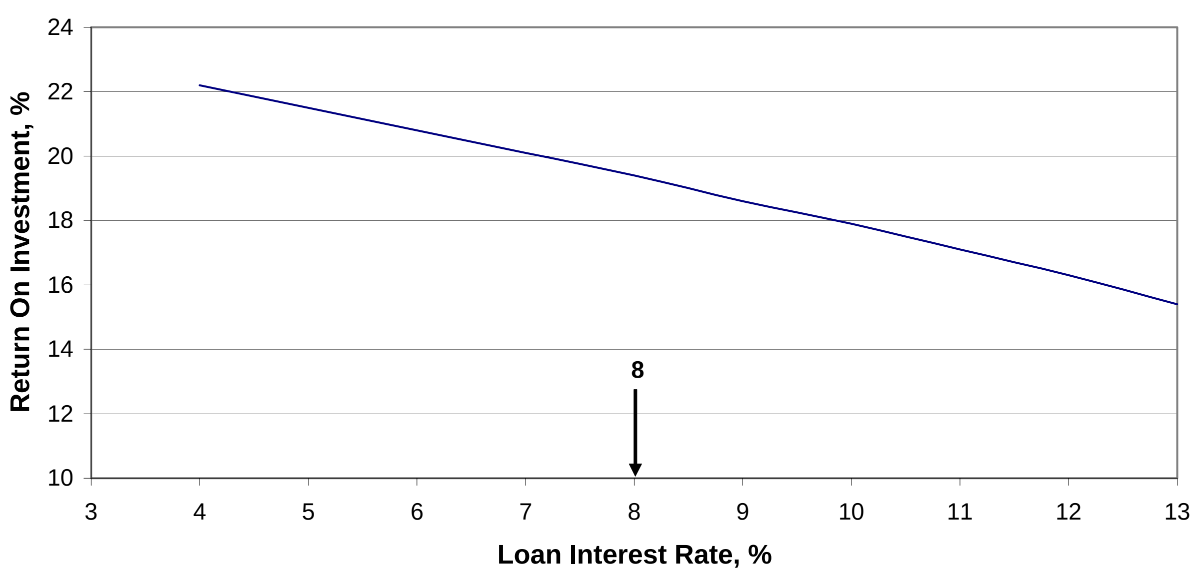
The impact that availability has on plant economics comes as little surprise. Reliable operation is very important to assure that the cost of project development and construction can be recovered. Long downtimes throughout the life of the project will significantly hurt overall project economics given a 20-year project life. The impact of availability on overall plant economics is similar to that of Subtasks 3.2 and 3.3. As mentioned earlier, both plant life and operating hours, which are related to availability since they both impact the length of plant operations, have similar impacts. Since a plant of this size should be built to have a lifetime longer than the 20 years used in the model, greater consideration of plant life should be made during the project development phases. Figure 6.5 makes this point more clearly. A certain economic life is required in order to pay off the debt incurred during project construction. Once this debt has been paid and construction costs recouped, the steady cash flow will lead to a stable rate of return.

Figure 6.5 Effect of Plant Life on Investment Return



The interest rate for debt financing plays a larger role in this case than in Subtasks 3.2 and 3.3. Monthly interest payments will be significantly higher than in the industrial gasification case. However, interest rate variations do not have a relatively greater significance than either availability or electricity price, as shown in Figure 6.6.

Figure 6.6 Effect of Interest Rate on Investment Return



As with Subtasks 3.2 and 3.3, availability and electricity tariff value should receive the most attention when considering the range of financial outcomes. Other parameters, while important to a complete picture of a facility's financial potential, do not have the impact of these two factors. Electricity tariff is especially important in this case due to the lack of other important plant outputs. The increase in capital costs in Subtask 3.4 makes the net plant investment of higher significance than in Subtasks 3.2 and 3.3. The conditions under which the plant is financed become more important with the capital cost increasing by a factor of four.

One key result of the sensitivity analysis is that positive investment returns were found for the entire range of variables that were analyzed. This demonstrates that the model and the economics are robust—even with large changes in the financial parameters required to establish a very "conservative" case, plant returns are still positive. The economic results can be stated with confidence that even if changes are made in some of the key financial parameters, the base case still provides a close estimate of plant economic performance. This range of outputs needs to be reconciled with the risk tolerance of the project developers.

The results of this analysis should not be applied to every facility considering lignite gasification. While these results are valid for the current site and timeframe, others interested in gasification applications must consider their own unique circumstances to develop a proper financial analysis. However, this sensitivity analysis provides insights into the parameters that will most likely have the greatest impact on economic feasibility.

7.1 INTRODUCTION

Because of scope, time constraints and budget limitations, several items were identified but not rigorously analyzed that could improve the process either by increasing the efficiency, increasing the availability, and/or reducing the cost. Some of these items were identified during the Value Improving Practices sessions for the optimization of Subtask 3.3, and are applicable to this subtask also. Some are not commercially demonstrated at the required capacity. Others are only applicable to the processing of lignite. This section lists those items that should be considered for improving the design of the Subtask 3.4 Lignite-Fueled IGCC Power Plant and estimates the value of the potential improvements.

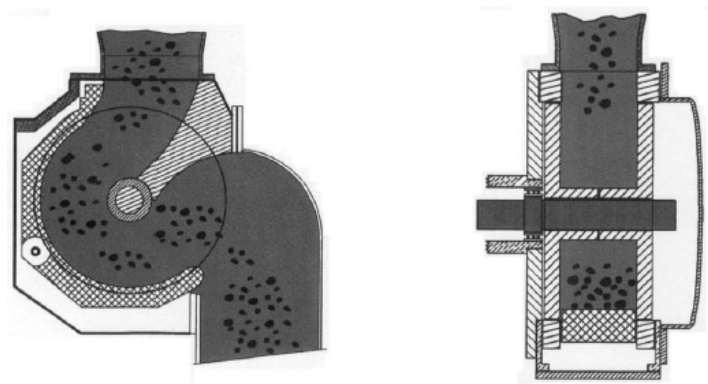
7.2 STAMET SOLIDS FEEDING SYSTEM

Stamet, Inc., of North Hollywood, CA, successfully completed the first phase of their cooperative agreement with NETL by pumping dry pulverized coal from atmospheric pressure to 300 psig at a rate of 150 pounds per hour.[1] Stamet's solid feeding system utilizes a radical technology known as Posimetric® solids feeding. The machine relies on a simple continuously rotating element that also provides precise flow control without valves or pressure vessels. No nitrogen or other gas is needed to pressure the system or maintain the operating pressure. The machine delivers fuel directly into the pressurized gasifier in a continuous controlled and uniform way that is more reliable and efficient than currently available dry-feed lock-hopper systems. The concept for the Posimetric® system was originally developed as a means to feed crushed oil shale into retorts.

The Stamet Posimetric® Pump has only one moving part; discs on a shaft forming a spool, which rotate within the housing as shown schematically in Figure 7.1. An abutment, extending between the discs to the hub, separates the inlet from the outlet. Material entering the pump becomes locked or bridged between the discs and is carried around by their rotation. This locking principle means the pump experiences virtually no wear. The abutment prevents material being carried around for an entire rotation and also makes the pump self-cleaning.

1 Contract No. DE-FC26-02NT41439, "Continuous Pressure Injection of Solid Fuels into Advanced Combustion System Pressures"

Figure 7.1 Diagram of Stamet's Posimetric® Solids Pump[2]



The Stamet solids feeding system, the so-called rock pump, was rejected for this application because of several reasons.

- At the current stage of development, the Stamet pump is not ready for a commercial application of this size. Multiple pumps would be required since this design is significantly larger than the current sizes available. To supply the required 108 tpd of solids to the gasifiers would require eight of the currently largest available pumps working in parallel.
- The Stamet pump has not been demonstrated at the pressure levels required to feed a gasifier that will be operating at 440 psia.
- The expected cost of the required pumps and associated equipment appeared to be higher then a comparable lock hopper/screw feeder system (see Table 6.1).
- Since the emphasis of Subtask 3.4 design was on developing a reliable base case facility using proven technology, the commercially unproven Stamet pump was rejected.

However, as improvements continue to be made and their demonstrated capacity increases, Stamet pumps should be reconsidered for this application.

7.3 WARM SULFUR REMOVAL SYSTEM

Commercially available sulfur removal technologies such as the MDEA, Selexol, and Rectisol processes require syngas temperatures below 140ºF. Other low-temperature processes currently available or under development include the LO-CAT® chelated iron solution, CrystaSulf®, and Morphysorb® processes.

The LO-CAT® process is a patented, wet scrubbing, liquid redox system that uses a chelated iron solution to convert H_2S to innocuous, elemental sulfur. It does not use any toxic chemicals and does not produce any hazardous waste byproducts. The iron

2 www.stametinc.com/html/technology.html

catalyst is readily available and continuously regenerated in the process. The iron catalyst is held in solution by organic chelating agents that wrap around the iron in a claw-like fashion preventing precipitation of either iron sulfide or iron hydroxide. The LO-CAT® process is based on reduction-oxidation (Redox) chemistry. Two different Redox reactions take place – one in the absorber section, which converts the H_2S to elemental sulfur, and one in the oxidizer section, which regenerates the catalyst.

As proposed for this application, the H_2S is removed from the syngas in an absorber using an amine solution. The COS is hydrolyzed to H_2S and CO_2 upstream in the COS hydrolysis reactor. The H_2S is stripped from the amine, thereby regenerating it. The H_2S is then sent to the LO-CAT® unit where it is converted to sulfur. Although the LO-CAT® process is commercially proven, it was rejected for this application because high chemical costs made the process more expensive compared to the conventional MDEA/Claus plant system.

The CrystaSulf® process is a new, non-aqueous sulfur recovery technology that can be used for direct treatment of gas streams. The CrystaSulf® solvent components are high-boiling organics in which sulfur has a high solubility. The solvents catalyze the reaction of H_2S with SO_2 to form elemental sulfur that remains dissolved in the solvent until removed by crystallization. This process was rejected because there are no commercial operating units at this time. However, three units are under license, but not yet operational.

The Morphysorb® process is an alternative absorption process for scrubbing H_2S from a gas stream. It leaves a residual H_2S content of about 10 ppm compared to MDEA that leaves a residual H_2S content of about 30 ppm. It requires a pretreatment step to convert COS to H_2S just like MDEA.

The MDEA system was chosen for this study because it is a commercially proven, off-the-shelf technology that can meet the sulfur removal levels necessary for this base case study that is less expensive than either Rectisol or Selexol. The other technologies mentioned have different process and performance characteristics and deserve consideration in an optimized case study. While the amine system is likely the most economical choice for the requirements of this study (see Addendum F), in cases where single digit H_2S concentrations are required (e.g., if SCR would be required for NOx reduction), CrystaSulf® or other alternative amine based systems may be a better choice.

Warm temperature (>240ºF) clean-up options for sulfur removal also are currently under investigation. The benefit of sulfur removal at elevated temperatures is greater efficiency because of the reduced need to cool the syngas upstream of the gas turbine. Several technologies exist including the SulfaTreat process and the UC Sulfur Recovery Process. However, these technologies are not yet commercially available for syngas operation, and several performance hurdles need to be overcome (e.g., the need for a polishing step to get >90% sulfur removal in the SulfaTreat case) before wide application. Additionally, the lower temperature process requirements of other gas cleanup steps (i.e., chloride and mercury removal) can reduce or eliminate the benefits of a warm temperature sulfur removal process.

7.4 WARM MERCURY REMOVAL SYSTEM

The traditional commercially proven method of mercury removal is by adsorption on sulfur-impregnated activated carbon. Because the syngas has to be cooled to 100 to 110°F for the amine acid gas removal system, there is no incentive to use a commercially unproven warm mercury removal system. Currently, higher temperature (400 - 550°F) mercury removal systems are under investigation, but they have not been demonstrated beyond pilot scale. Any benefits that would result from the use of gas clean-up technologies at higher temperatures (including sulfur removal as described above) would be reduced or eliminated if additional cooling is required in a downstream processing step since the syngas would have to be reheated before going to the gas turbine. The greatest benefits of a higher temperature mercury removal system likely will be achieved only when warm temperature sulfur removal options also are available. However, incremental benefits may be recognized once alternative mercury removal technologies progress beyond the pilot scale. Such technologies again should be evaluated after substantial performance and economic data are available.

7.5 IMPROVED PARTICULATE REMOVAL SYSTEMS

The current design has the particulates and dust being removed from the syngas by three stages of cyclones followed by metallic candle filters. The third stage cyclone operates at near gasifier temperature, about 1600°F, and the metallic filters operate at about 650°F. The solids collected from both systems are discarded. Use of metallic candle filters on syngas at these conditions has been commercially proven at the Wabash River Repowering project.

The primary purpose of adding the filters to the system was to reduce the water usage in the scrubber column, to reduce the size and complexity of the wastewater treatment system and to increase the useable heat recovery. Furthermore, removing all the solids above organic hydrocarbon condensing temperatures should eliminate plugging problems in the scrubber.

As more experimental studies are conducted using syngas from the non-slagging U-GAS® gasifier, there are several modifications to this system that should be considered.

- Elimination of the third-stage cyclones. With demonstrated filter performance at the gasifier temperature, the third-stage cyclones could be replaced with filters.
- Increasing the operating temperature of the filters. If the filters can be operated at the gasifier temperature, the primary syngas cooler will be simpler, less costly, and less prone to problems because the syngas will be solids-free.

- Replacing the metallic filters with cheaper ceramic filters. The ceramic filters that were originally used at the Wabash River Repowering project did not have the required durability. Newer ceramic filters should be investigated to see if they have improved durability.

7.6 OPTIMIZE THE MAKEUP WATER PURIFICATION SYSTEM

The current design purifies the makeup river water by a generic filtration and ion exchange system. Since a specific site and makeup water composition is not known, the properties of the makeup water are unknown. Alternate makeup water purification schemes, such as reverse osmosis should be considered.

7.7 IMPROVED HEAT INTEGRATION

There are various opportunities for improving the heat integration in the system, but these were not implemented due to the excessive complexity they would add to the system. Some of the options identified but not rigorously analyzed are listed below.

- The heated boiler feed water at 350°F leaving the ash removal screws is used to preheat the syngas or oxygen going to the combustion turbine. This will improve the plant efficiency. The cooler boiler feed water still can be reheated to the same temperature (450°F) in the HRSG because there is extra heat in the flue gas that can be utilized. The net result is that less syngas would be required to fire the GE 7FB turbine, less lignite would be used, and the plant size could be slightly reduced.
- The heated BFW from the HRSG also could be used for this purpose. This would reduce the size of the syngas heater but increase the size of the economizer in the HRSG.

7.8 COMBINED ASH HANDLING SYSTEMS

The bottoms ash and fly ash handling systems could be combined as was done for Subtask 3.3, Alternate Air-Blown Design Case. This not only simplifies the design, but also allows for economy of scale when designing the ash handling system. A description of the new scheme is provided below.

The ash removal system consists of a series of equipment to cool both the bottom ash and fly ash and transport the combined ash via a lockhopper from gasifier pressure to storage at atmospheric pressure. A pressurized cooling screw cools the ash from the gasifier temperature to about 500°F to protect the lockhopper valves and to allow use of carbon steel equipment downstream. The screw is rotated when the valve to the lockhopper is open and stopped when the lockhopper valve is closed. The screw is also rotated when the valve from the fly ash surge hopper is opened and stopped when both the lockhopper valve is closed and the surge hopper valve is closed. A refractory lined surge hopper collects the ash when the screw is not operating (lockhopper closed). When the lockhopper is full (confirmed by nuclear level detectors) the upper valve is closed, the vessel pressure is lowered to atmospheric, and the discharge valve is opened. Ash is then transported via a pneumatic system to a day tank from which it can be disposed or sold.

The dust (fly ash) removal system consists of a surge hopper below the tertiary cyclone and a refractory-lined carbon steel pipe that combines the fly ash with bottom ash discharge line.

7.9 ESTIMATED SAVINGS BY IMPLEMENTING THESE IMPROVEMENTS

The estimated savings by implementing these improvements is estimated in Table 7.1.

Table 7.1 Capital Savings from Potential Improvements

IMPROVEMENT	ESTIMATED SAVINGS, k$
STAMET SOLIDS FEEDING SYSTEM	750
COMBINED ASH REMOVAL SYSTEMS	7,000
WARM SULFUR REMOVAL SYSTEM	3,600
WARM MERCURY REMOVAL SYSTEM	0
IMPROVED PARTICULATE REMOVAL SYSTEMS	400
OPTIMIZE THE MAKEUP WATER PURIFICATION SYSTEM	3,900
IMPROVED HEAT INTEGRATION	5,500
TOTAL	**21,150**

8.1 SUMMARY

A conceptual design was developed for a lignite-fueled IGCC power plant using Gas Technology Institute's U-GAS® fluidized bed gasification technology. The plant is located at a generic North Dakota site. Table 8.1 summarizes the major input and output streams along with some key operating parameters.

Table 8.1 Overall Plant Summary

	Lignite-Fueled IGCC Power Plant
Design Inputs	
Lignite Feed, moisture-free tpd	2,558
Lignite Feed, moisture-free lb/hr	213,160
Fuel (Natural Gas), million Btu/hr	8.93
Makeup Water, gpm	1,920
Design Outputs	
Export Power, MW	251.0
Sulfur, lb/hr	1,557
Ash, lb/hr	23,729
EPC Cost, M$*	410.5
Plant EPC Cost, $/kW	1,635
Plant Energy Input, k$/million Btu/hr	174.7
Plant Energy Output, k$/million Btu/hr	478.2
Cold Gas Efficiency, % (HHV basis)**	84.0
Overall Thermal Efficiency, % (HHV basis)***	36.5

* EPC cost is on second quarter 2004 dollars at the North Dakota location. Contingency, taxes, fees, and owners costs are excluded

** Cold Gas Efficiency is defined as the energy in the syngas leaving the gasifier relative to the energy of the feed coal (HHV basis)

*** Overall Thermal Efficiency is defined as the export electrical energy from the turbine relative to the energy of the feed coal (HHV basis)

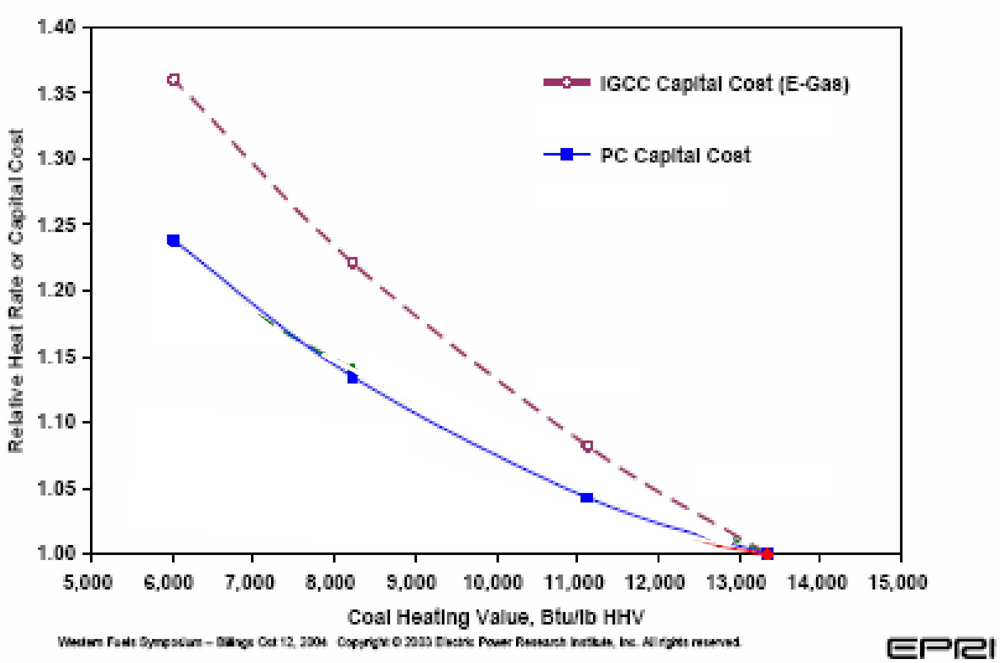
At design conditions, the plant consumes 2,558 tpd of moisture-free lignite feed and produces 251 MW of export electric power. It also produces 277 tpd of by-product sulfur. Because the as-received lignite contains 32.24% moisture, the lignite crushing and drying area processes 3,775 tpd of wet lignite. In addition the plant requires 2,763 gpm of makeup river water and 8.93 MBtu/hr of natural gas. This small amount of natural gas is used in the Claus plant, and since this usage is small, it is being considered as an operating cost and not as a plant feed.

At design conditions, the plant exports 251 MW of power. It has a thermal efficiency of 36.5%. The thermal efficiency is a direct function of the high moisture content of the lignite feed. In the gasifier, the residual moisture is vaporized and heated to 1600°F, thereby consuming a substantial amount of energy.

A financial analysis has shown that the addition of a spare gasification train (from coal feeding up to and including the particulate filters) improves the financial performance of the facility so that it will have a return on investment (ROI) of 19.4% and a payout period of 4.8 years. This ROI is about 0.1 percentage point higher than the ROI of a similar plant without the spare gasification train even though the spare gasification train adds 25 M$ to the EPC cost. The addition of a spare gasification train increases the overall plant availability by almost 7 percentage points (with scheduled maintenance). This, when coupled with the slightly improved economics, led the team to choose this plant configuration over a design without a spare train.

The estimated EPC cost of the grass roots facility is 410.5 M$ (second quarter 2004 dollars), or about 1,635 $/kW of design export power. Although IGCC is competitive with pulverized coal (PC) plants for bituminous coals (e.g. Illinois No. 6 or Pittsburgh No. 8), the capital cost of a PC plant is 300-400 $/kW lower than IGCC for low rank coals[1] (e.g., Texas lignite or Wyoming PRB). This relative costs can be seen in Figure 8.1.

Figure 8.1 Effect of Coal Quality on PC and IGCC Plant Capital Cost[1]



1 Holt, Neville, Issues and Research Needs for Low-Rank Coal Gasification, Western Fuels Symposium, Billings, MT October 12-14, 2004

8.2 CONCLUSIONS

This study has shown that:

- There are commercially available processes and technologies available for the design of a lignite-fueled IGCC power plant based on the U-GAS® gasification technology that should provide reliable, long-term operation.

- A ROI of 19.4% is achievable at the current market price of electricity in North Dakota. Future optimization of this plant design should identify several enhancements that will further improve the economics of IGCC power plants (also see below for a list of potential enhancements and improvements).

- Results of a sensitivity analysis showed that capital investment, availability and electrical tariff are the most sensitive financial parameters.

- During development of this design, a spare gasification train was justified because it increases the overall availability and ROI even though it increases the capital investment.

- As a result of this study, a list of potential enhancements has been identified (see Section 7) that should provide additional cost savings as some of the improvements are researched, developed and implemented.

    - The Stamet solids feeding system
    - Warm sulfur and mercury removal systems
    - Improved particulate removal systems
    - Optimization of the makeup water purification system
    - Combined ash removal systems
    - Improved heat integration

- As a result of this study, a list of R&D needs have been identified including:

    - Further study of lignite drying techniques
    - Investigate the effect that the lignite moisture content has on the U-GAS® gasifier operation
    - Update the database for gasification reactivity of the desired coal
    - Further study of the ash characteristics associated with the char

- Based on the simulations prepared for this study the design should meet emission targets established by the DOE in their roadmap for 2010 (see Section 5.3).

Another objective was to train several NETL employees in the methods of process design and system analysis. These individuals worked closely with the Nexant and Gas Technology Institute personnel in developing the design.

8.3 RECOMMENDATIONS

Technology development will be the key to the long-term commercialization of gasification technologies and integration of this environmentally superior solid fuel technology into the existing mix of power plants. The following areas are recommended for further development through additional systems analysis or R&D efforts:

- Additional optimization work is required for lignite coals. This may entail changes to the plant configuration to include improved heat integration and sulfur recovery methods.

- Demonstration of the warm gas clean-up technologies so that cooling of the syngas (i.e., below 300°F) can be eliminated, and the plant can be made more efficient from a thermal standpoint.

- Develop a R&D program that will address the critical issues such as:
  - Developing a technique for drying lignite feedstock that doesn't lead to spontaneous combustion and at the same time increases the overall efficiency.
  - Commercialize the Stamet feed pump at high pressure and high capacity to reduce the cost of the gasifier feed system and provide a more reliable feed system.
  - Prove the availability of the gasification system and various sub-systems.
  - Determine the combustion turbine performance (both power output and emissions) on the design syngas in order to prepare for commercialization.

- Although it is known that reducing the moisture content of the lignite feed to the gasifier is more efficient than evaporating the moisture in the gasifier, it has not been established that 20% is the optimum moisture content of the gasifier feed. This needs to be more thoroughly investigated.

- The physical characteristics and properties of lignite must be studied further in order to better predict the gasification systems. These include:
  - Determination of the gasification reactivity of the desired feedstock.
  - Determine the ash characteristics associated with the char
  - Characterize the particulate properties

  - o Characterize the hydrocarbon content of the syngas to confirm the design of the sour water stripper and effluent water treatment facilities

- Determination of cyclone performance at higher temperatures.

  - o During a visit to a gasification facility in China it was noted that at temperatures above 1000°F, the cyclone efficiency drops off sharply. This was confirmed by Emtrol, a world leader in cyclone design.

A.1 ASPEN

A.1.1 Gasification Island

A.1.1.1 Basis

This model was developed to reproduce the exit gas and solids compositions from GTI's design of the gasification island (Units 300, 400, and 500).

A.1.1.2 Introduction

GTI's gasifier, U-GAS®, is a fluidized bed gasifier that can operate over a wide range of temperatures. For North Dakota lignite, GTI recommends operation at about 1600°F. The gasifier also can operate over a wide range of pressures ranging from 1 to 70 atmospheres. The pressure selected is sufficient to provide the syngas to the gas turbine without compression. The gasifier has two cyclone separators that recycle the unburned carbon particles and fly ash back to the gasifier. The oxidant, steam and fuel enter the gasifier from the bottom.

Aspen Plus® version 11.1.1 was used to model the GTI gasifier. The basic model was developed for modeling the gasifier and dryer during Subtask 3.2. In this task, the model was fine tuned to reflect the lignite gasification process. GTI has provided design parameters along with syngas compositions and ash content. The coal drying process was designed by the Raymond Professional Group. A generic Aspen model was adopted to simulate the coal drying process. A description of the Aspen model and significant results are provided in this section.

The gasifier temperature was set at 1600°F, and the outlet pressure was set at 440 psia. Nine individual reactions were specified and the chemical equilibrium of each reaction was restricted by varying the temperature approach of each individual reaction. This technique is reasonable since the gasifier is not completely homogenous.

A.1.1.3 Setup

Coal Prep – Drying

The component attributes of the coal were taken from the U.S. government's Quality Guidelines for Energy System Studies corresponding to a seam near Beulah, North Dakota. The "as-determined" proximate analysis for the coal sample was used for the PROXANAL attribute in ASPEN. The "dry basis" ultimate analysis for the coal sample was used for the ULTANAL attributes. These coal properties are shown in Table A.1. Trace components in the coal such as mercury, etc., were neglected because this information was not available. Therefore, no comparisons could be made.

Table A.1 Typical Beulah North Dakota Lignite Properties

Ultimate Analysis	**Dry Basis**
Carbon	65.85
Hydrogen	4.36
Nitrogen	1.04
Chlorine	0.04
Sulfur	0.80
Oxygen	18.19
Ash	9.72
Total	**100**
Proximate Analysis	**As Determined**
Total Moisture	32.24
Volatile Matter	30.99
Fixed Carbon	30.18
Ash	6.59
Total	**100**

An RSTOIC reactor block was used to dry the coal. In the RSTOIC block, a portion of the coal reacts to form water. A FORTRAN calculator block was used to calculate the amount of water that had to be removed so that the dried coal would have 20% moisture content. The model also is setup to incorporate nitrogen if used in the drying process. Other gases can be used as well, such as flue gas drying as proposed by the Raymond Professional Group. Since very few details were provided related to the drying fluid in the dryer, a more sophisticated coal drying model was not needed. A FLASH2 block is used to separate the dried coal from the moisture and nitrogen. A schematic diagram of the ASPEN simulation is given in Figure A.1.

The Gasifier

The gasifier model contains 5 blocks. A non-stoichiometric reactor based on a known yield distribution (RYIELD), a component separator (SEP), a Gibbs equilibrium reactor (RGIBBS), a substream splitter (SSPLIT), and a stream mixer (MIX).

- RYIELD - This block, named DECOMP, decomposes the coal into its constituent elements and ash. The heat of reaction associated with the decomposition of the coal is passed to the RGIBBS block.

- SEP – This block, named UNREACT, selectively separates specific amounts of each component into various streams. The block is used to bypass the unburned coal around the RGIBBS reactor block. The “BLEED” stream, one of the outlet streams for this block, contains N_2, O_2, and H_2 in vapor form; they are part of the particulates in the bottom ash. It is important to separate them before being routed to the RGIBBS reactor for reaction, and they are counted towards the bottom ash.

- RGIBBS – This block, named GASFIRXR, models the gasification of the coal. The RGIBBS block models chemical equilibrium by minimizing the Gibbs free energy.
- SSPLIT – This block, named ASHSEP, splits the bottom ash and flyash from the syngas.
- MIX – This block, named BOTMIX, mixes the unburned carbon and sulfur that bypassed the RGIBBS block with the bottom ash.

Split Fractions - Modeling the unburned coal and ash

To properly model the unburned carbon in a gasifier, calculated amounts of the components that make-up the char (carbon, hydrogen, oxygen, nitrogen, and sulfur; commonly abbreviated as CHONS) and the ash in the coal have to bypass the RGIBBS reactor. These component split fractions are calculated based on GTI's data obtained from proprietary simulation tools. The two cyclone separators that remove fly ash from the syngas and return it to the gasifier are considered part of the gasifier and were not modeled separately. Since ASPEN converts the solid coal stream to a "MIXED" stream, the char must be handled as gaseous components. Therefore a bleed stream is used to remove the hydrogen, oxygen, and nitrogen elements so that they will not be included in the gaseous syngas composition. The sulfur, carbon, and ash components are included in the streams after the gasifier RGIBBS reactor. Based on the GTI data, the split fractions of C, H, O, N, S, and ash are calculated. The char components and the ash bypass the gasifier reactor block by implementing a component separator block. The separated char and ash split fractions are then sent to the bottoms and raw syngas streams in order to accurately represent the amount of char and ash in these streams. Also, it is noted that the composition of the char in the bottom ash is different than that of the char in the raw syngas stream. Next, the char and ash in the raw syngas stream are sent to another stream splitter that modeled the external cyclone separator. The split fraction is based on the performance of the cyclones (which is 50% removal of the ash/char particles).

Tuning the Gasifier

Tables A.2 lists the reactions and the temperature approaches from the actual gasifier temperature for the model. These temperatures were calculated by trial and error with the aim to minimize the differences in syngas heat content and component mass compositions between the ASPEN evaluated syngas compositions and the GTI data.

Table A.2 Temperature Approach for RGIBBS Reactions

Reactions	ΔT
1. $C + 2H_2 = CH_4$	-6.00
2. $CO + H_2O = CO_2 + H_2$	5
3. $H_2 + S = H_2S$	31
4. $C + O_2 = CO_2$	0.0
5. $C + 0.5O_2 = CO$	3.5
6. $0.5N_2 + 1.5H_2 = NH_3$	-700
7. $CO + S = COS$	60
8. $NH_3 + CO = HCN + H_2O$	100
9. $Cl_2 + H_2 = 2HCl$	10

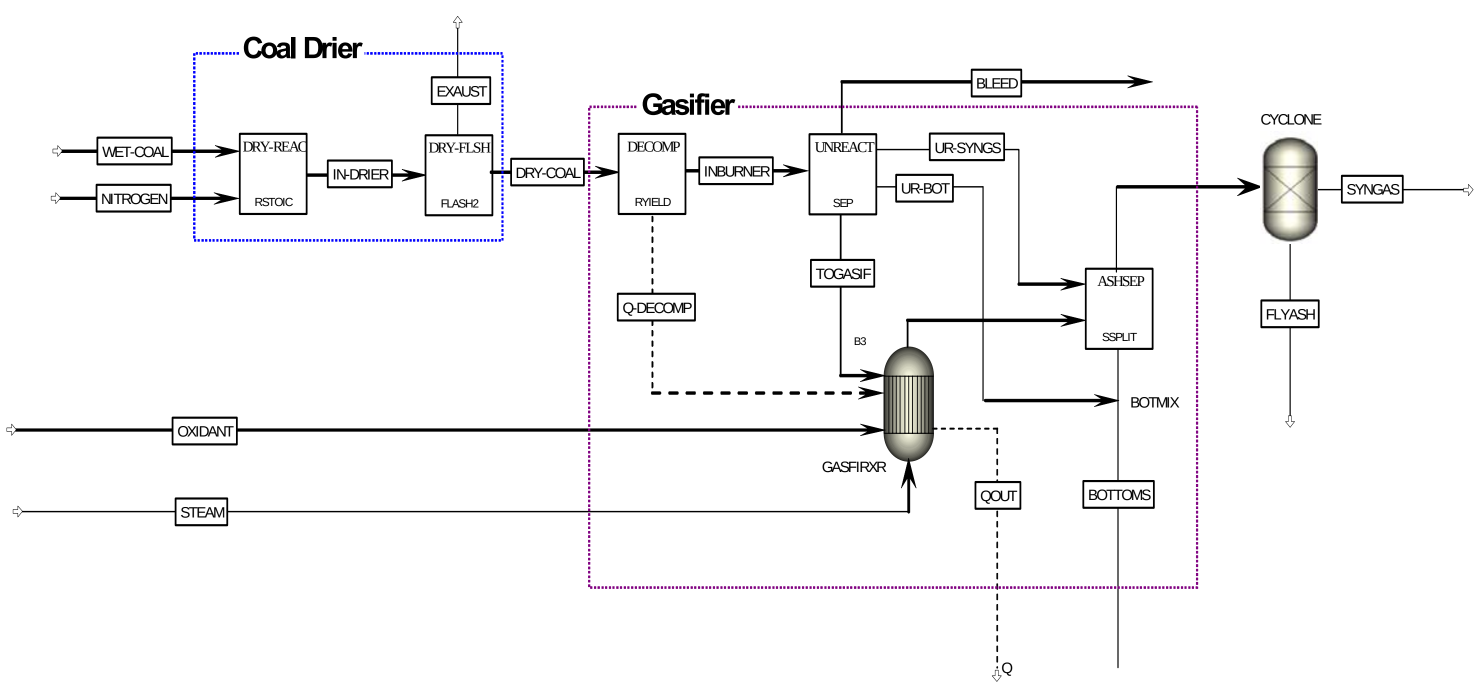
A.1.1.4 Results

With the evaluated temperature differences, the model was able to match GTI's results with a high level of accuracy. Table A.3 compares the results from the ASPEN model with GTI's results. Note that the GTI results show a carbon imbalance of 0.4%. The components are well predicted by this model, with the greatest difference in the methane prediction. The resultant heating value of the syngas differs from the GTI result by less than 0.25%.

Table A.3 U-GAS® Gasifier Simulation Results

Stream Composition	GTI (lb/hr)	Model (lb/hr)	Difference (lb/hr)	% Error
CO	84,505.1	84,506.8	1.65	0.00%
CO_2	68,403.3	68,398.14	-5.16	-0.008%
H_2	4,911.0	4,886.4	-24.58	-0.50%
H_2O	28,173.0	28,175.4	2.44	0.01%
CH_4	10,244.9	10,341.3	96.37	0.94%
H_2S	800.8	800.8	0.05	0.01%
COS	35.3	35.6	0.30	0.84%
NH_3	307.6	307.5	-0.08	0.0%
HCN	19.5	19.5	0.00	0.0%
N_2	3,178.1	3,178.2	0.06	0.0%
HCl	40.0	40.03	0.03	0.08%

Figure A.1 ASPEN Block Diagram



A.1.2 Syngas Cleanup System

ASPEN Plus provides a number of physical property methods for calculation of stream thermodynamic properties under various conditions; different property methods will yield different results, and sometimes these results can have significant repercussions on the entire design. For the current system, caution needs to be exercised in designing the syngas water scrubber and the flash drum downstream of the low temperature heat recovery system, since some of the gases are dissolved in the sour water and process condensate, which will be treated in the sour water stripper. It is important to obtain a realistic estimate of the sour gas composition such that the downstream equipment (i.e., the sour water stripper and the acid removal system) can be conservatively designed.

For the syngas, which contains a large quantity of hydrocarbons, ASPEN Plus recommends the use of the Peng Robinson – Boston Mathias (PR-BM) property method set. However, for applications involving electrolytes, such as an acid gas removal system, the ElectrolyteNRTL property method set is suggested. A portion of the NH_3, H_2S, and CO_2 in the syngas are dissolved in the sour water and process condensate. To correctly account for the acid gases in the sour water and process condensate the ASPEN Plus simulation developed for the current design incorporates results obtained from both the ElectrolyteNRTL and PR-BM methods.

ASPEN simulations were prepared for both the initial 10% and final 20% lignite moisture cases.

A.1.3 Sour Water Stripper

The sour water stripper simulation was only prepared for a case drying the lignite to 10% moisture content instead of the 20% moisture content used in the final design. The 20% moisture lignite material balance was prorated from the 10% moisture lignite material balance.

A.1.3.1 Sour Water Treatment System

The sour water treatment system removes ammonia, hydrogen sulfide, other volatile impurities, and solids from the sour water such that the cleaned water is of sufficient purity for process recycle or discharge to the waste water treatment system.

A.1.3.2 Sour Water Streams

The largest sour water feed stream is from the water scrubber down stream of the intermediate temperature heat recovery boiler. A portion of the process condensate also is mixed with the scrubber water and treated in the sour water treatment system. In addition to the dissolved impurities to be removed by the stripper (CO_2, NH_3, H_2S), the sour water also contains some fine particles (<1.3 microns) that are not removed by the candle filter particulate removal system. Some condensed oils (benzene and toluene derivatives) also may be in the sour water.

A.1.3.3 Sour Water Stripper

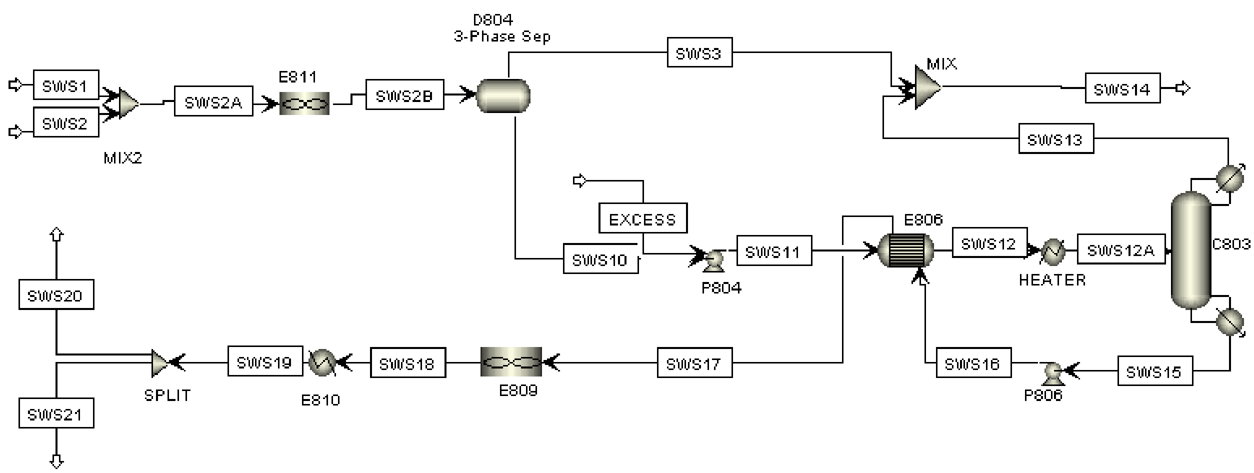
The sour water treatment unit processes the effluent from the syngas water scrubber and process condensate from the flash drum upstream of the amine system. This unit consists of a 3-phase separator, sour water stripping column, and associated heat exchangers and pumps. Vapors from the 3-phase separator and stripping column are sent to the sulfur plant. Stripped water from the bottom of the column is recycled to the water scrubber with a blowdown stream sent to the wastewater treatment plant.

The distillation column was designed based on past experience and information obtained from Kohl.[1] The stripping column generates a liquid effluent stream containing no more than 50 ppmw ammonia and less than 10 ppmw hydrogen sulfide.

A.1.3.4 Sour Water Stripper Modeling

The sour water stripper was modeled using ASPEN Plus Version 11.1. Figure A.2 shows the ASPEN process flowsheet of the sour water treatment unit.

Figure A.2 ASPEN Process Flowsheet of the Sour Water Treatment Unit



Input streams were obtained from modeling the gas clean-up system, which is described in Section A.1.2 of the this addendum. Physical properties were modeled using the ASPEN ELECNRTL (NRTL Electrolyte) property method. Proper selection of the correct property method is critical to the modeling of this system.

The 3-phase separator (Figure A.2, D804) was specified at 35 psia and zero duty to provide a vapor stream at the same pressure as the vapor stream of the sour water stripper. Preheater (E806) is a countercurrent heat exchanger that recovers heat from the column bottoms stream. Because ASPEN blocks are calculated sequentially, an additional heat exchanger (HEATER) was used for the initial modeling run (zero

[1] Kohl, A. and Nielson, R; *Gas Purification – Fifth Edition*, Gulf Publishing Company, 1997

pressure drop and temperature specified near column operation) to speed up model convergence. After the column (see below) converged, the duty of this additional heat exchanger (HEATER) was set to zero and the model was re-run.

The stripper column (C803) was modeled using a RADFRAC block. The column was specified similar to the design described in the literature with 20 actual stages (10 theoretical stages) plus a partial condenser and kettle reboiler. Sieve trays were specified for the design. The kettle reboiler duty was specified to meet the desired discharge water composition (<50 ppmw NH_3) which also satisfied the H_2S specification (<10 ppmw). The reflux ratio was manually adjusted to keep the water concentration between 25 wt% and 30 wt% of the total vapor stream going to the sulfur plant.

Cooling of the hot stream downstream of the pre-heater (E806) was modeled using two coolers in series based on the assumption that 140°F is the economic break point between an air-finned cooler (E809) and a water cooler (E810). The cooled water stream at 110°F is split between recycle water to the water scrubber and discharge to the waste water treatment plant.

A.2 GATECYCLE

A.2.1 Power Block

The power block consists of the one GE 7FB combustion turbine set (CT and generator), one heat recovery steam generator (HRSG), and a single steam turbine. The power block was modeled using the GateCycle computer simulation program for Windows Version 5.52.0.r. The syngas composition was generated separately using an ASPEN Plus model of the syngas cooling and cleanup areas. It is described in other sections of this report.

A.2.2 Combustion Turbine Modeling

One GE 7FB combustion turbine was selected for this facility. The GateCycle library model of the GE PG7241 turbine [(FA) (DLN natural gas 1998)] was used to represent the turbine. The turbine model was adjusted to match the results of a previous study by Nexant.[2] The calibration involved adjusting a number of parameters affecting turbine performance curves; these parameters include turbine heat rate, flue gas temperature, and power output. The syngas composition was generated using ASPEN Plus and was the basis for generating the fuel input for GateCycle.

It is important to note that some degree of uncertainty exists when modeling coal-derived syngas (or any low Btu syngas) with the stock turbines provided in the GateCycle software turbine library. The actual performance needs to be validated by GE engineers. NOx control was achieved by nitrogen dilution of the fuel mixture.

[2] Nexant 1.3 plant.

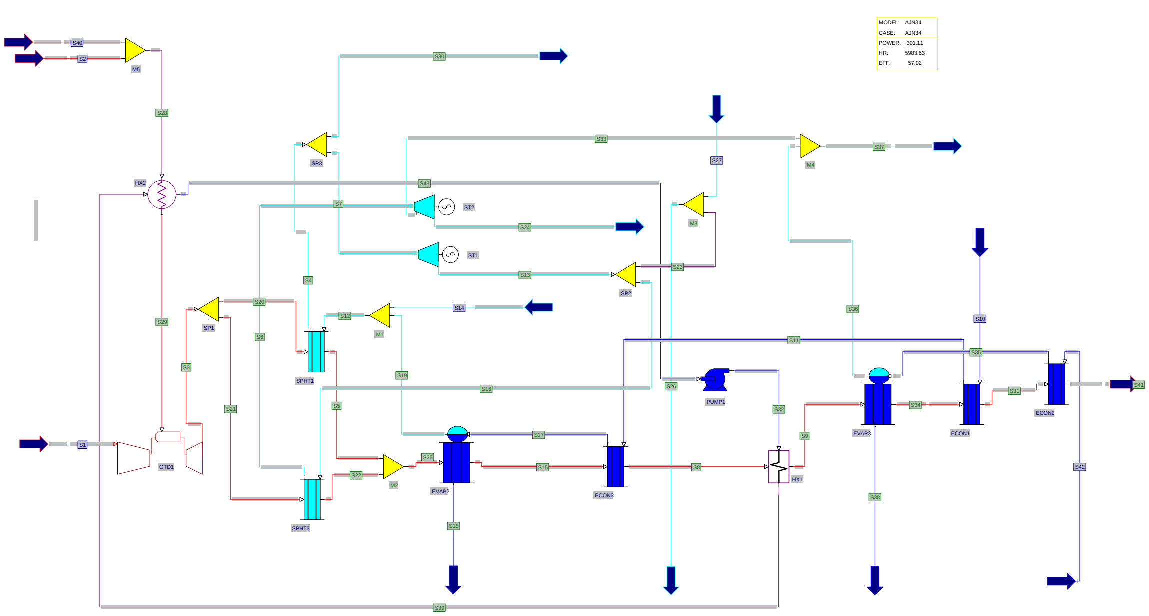
A.2.3 HRSG Modeling

The HRSG was modeled such that the following specific process conditions were met:

- Stack temperature remains above the acid-dew point so that condensation and corrosion do not occur within the system.
- A reheat steam turbine was used for better performance. 1,000 psig superheated steam powers the high pressure section; the inlet of the condensing section is at 500 psig and the exit is at 1.5 psia. Both sections and the electricity generator share a common shaft.
- A portion of 500 psig superheated steam is used for the gasification process.
- 50 psig superheated steam is extracted from the 500 psig turbine section for the low pressure steam requirements in the plant.
- A portion of the1,000 psig superheated steam is used to heat the oxygen exiting the ASU.
- The mixture of syngas and the diluent nitrogen is preheated in the HRSG prior to entering the gas turbine combustor to improve efficiency.

The modeling was accomplished by inserting the appropriate HRSG components downstream of the turbine exhaust. Figure A.3 is a screen capture of the GateCycle process flow diagram of the power block.

Figure A.3 GateCycle Process Flow Diagram of the Power Block



A description of the HRSG is provided in the body of the report (para. 5.2.8.3).

Appropriate pinch temperatures were applied to various pieces of equipment; 20°F for all evaporators and the economizer. The amount of 50 psig steam extracted was dictated by the steam requirements of the plant. The efficiencies of the steam turbines and the pump are 90% and 85%, respectively.

A.2.4 Steam Turbine

A reheat steam turbine was selected for this application to improve performance. The inlet to the high-pressure section is at 1000 psig, and the inlet to the low-pressure section is at 500 psig. The condenser side of the low-pressure section is at 1.5 psia, and steam exits the high pressure section at 550 psig and it is in the form of superheated steam.

B.1 COAL PREPARATION

The equipment included in the coal handling and preparation area (Unit 100) includes:

- Rail Dump Hopper and Auxiliaries
  - Stephens-Adamson railcar mover-positioner with 50 hp Hydraulic Pump.
  - 32' Kinergy railcar discharger with two 10 hp motors
  - Kinergy railcar shaker, 15 hp Motor
  - Two 84" by 14' Belt Feeders with 20 hp Drives
  - 48" x 48" Diverter Valve
  - Griffin Dust Collector Model 588 CG requiring 35 cfm air for bag cleaning
  - Dust Collector Fan, 60,000 cfm, 200 hp Drive
  - 9" diameter x 30' long Dust Return Screw Conveyor with 3 hp Drive
  - Belt Conveyor 48" x 152' C-C, 125 hp drive with 75' covers, walkway and magnetic head pulley
  - Transfer Tower
  - Motor Control Center/Control Panel

- Thawing Equipment
  - The railcar thawing equipment consists of infrared heaters that require 200 kW of electric power to heat the contents of the car to 500°F. The radiant heaters are mounted on the sidewalls of the building and on the middle of the rail track.

- Seven Day Active Pile and Reclaimer
  - Belt Conveyor 48" x 351' C-C x 120' Lift, 550 fpm belt speed, 250 hp, with full length
  - Walkway and covers, 1,500 tph design capacity.
  - 12' Diameter x 110' high concrete stacking tube.
  - Three 15' Diameter Kinergy SPD 15-HD pile dischargers with two 10 hp Motors each, stainless steel liners
  - Three Kinergy 24" x 5' vibrating feeders with 2 hp drive and dust covers
  - Two (2) Belt conveyors 24" x 329' C-C x 50' lift, 350 fpm belt speed, 20 hp with 100' covers and Walkway, 161 tph capacity.
  - Diverter Valve
  - Two (2) Screw feeder/distributor with drive

- Thirty Day Inactive Pile and Reclaimer
  - Belt conveyor 48" x 322' C-C x 110' lift, 550 fpm belt speed, 200 hp drive, with full length walkway and covers, 1,500 tph.

- Belt conveyor 48” x 660’ C-C x 110’ Horizontal, 550 fpm belt speed, 100 HP drive, with traveling tripper, full length walkway, 1,500 tph.
- Ten 15’ Diameter Kinergy SPD-15-HD pile dischargers with two 10 hp motors each, stainless steel liners.
- Ten Kinergy 48” x 8’ vibrating feeders with 3 hp drive and dust covers.
- Belt conveyor 48” x 533’ C-C x 25’ lift, 550 fpm belt speed, 100 hp drive, with 30’ length walkway and covers, 1,500 tph.
- Belt conveyor 48” x 302’ C-C x 12’ lift, 550 fpm belt speed, 60 hp drive, with full length walkway and covers, 1,500 tph.

- Crusher and Auxiliaries (two identical packages)
  - Heyl & Patterson Crusher Type to crush 2” x 0” lignite coal at 40% moisture to ¼” x 0 with a minimum fines generation below 100 mesh, design capacity at 162 tph, 200 hp motor.
  - Explosion-proof motor, slide base, coupling, guard
  - Feed Hopper to Crusher
  - Vibratory Screen
  - Heyl & Patterson vibrating feeder to retrieve product from crusher w/driver
  - Discharge Surge Bin (common to two crushers)

- Dryer and Screen (four identical packages)
  - Kinergy Fluid Bed Dryer System with the following auxiliaries:
    - One Air Blower
    - Four Heating Coils employing cooling water return, sour water from the venturi scrubber, stripped water from the SWS and 50 psig steam as the heating medias
    - One Cyclone Separator
    - One Exhaust Baghouse with air lock
    - One Exhaust (ID) Fan
    - One Supply Screw Feeder
  - Product Screw Conveyor with drive
  - One product collecting screw feeder
  - One lot dust collecting duct and stack
- Primary Storage Silo, Filling and Discharge System
  - Bucket elevator, 24” x 54”, 160’ tall, and 60 hp drive.
  - 30” Diameter x 30’ long screw conveyor with 20 hp drive
  - 44’ Diameter x 150’ high concrete silo, 3,888 ton capacity, with 60 degree conical section with stainless steel liner, 15’ diameter outlet at 10’ elevation.
  - 15’ Diameter Kinergy vibrating bin discharger with 10 hp drive
  - 30’ Diameter x 30’ long screw feeder with 20 hp variable frequency drive

  - o Bucket elevator 24” x 54”, 130’ tall with 50 hp drive
  - o JV-36Q bin vent with 2 hp fan.
  - o Nitrogen blanketing system

- Coke Silo, Filling and Discharging System Package

  This package includes the following:
  - o One lot truck dump discharger, elevator and recycle with
    - Three 16” Diameter x 18’ screw conveyors, each with 5 hp drive
    - 16” Diameter x 15’ screw conveyor with 3 hp drive
    - 16” Diameter x 40’ screw conveyor with 5 hp drive
    - 16” Diameter x 45’ screw conveyor with 7.5 hp drive
    - One lot chutes, supports, gates
    - SB-12724 Bucket elevator, 95’ height, 10 hp drive
  - o 20’ Diameter x 80’ high concrete coke storage silo, 350 ton capacity, 60 degree conical section with stainless steel liner, 8’ diameter outlet at 10’ elevation
  - o 8’ Diameter Kinergy vibrating bin discharger with 1.5 hp drive
  - o 30” Diameter x 30’ long screw feeder with 20 hp variable frequency drive
  - o Bucket elevator 24” x 54 “, 130’ tall with 50 hp drive

- Facility Railroad Tracks and Locomotive Engine

  - o One diesel locomotive switch engine, manufactured by Caterpillar, 2,000 hp
  - o 11,000 feet of railroad tracks.

The Coal Handling System Supplier will provide input/output signals to the plant main control system (DCS) provided by the Owner.

B.2 AREAS 150 THROUGH 1000

Area 150 Oxygen Supply

Identification	No.	Description	Comments	Unit Size
S-161	1	Air Separation Plant	Quote from Air Products	
	2	N2 Heat Exchanger	Cool the N_2 used for blanket the coal silo. N_2 Flowrate =	10 kW.
E-151 A/B	3	O2 Heat Exchanger	using 1000 Psig steam to heat up the O_2 to the gasifier from 90 F to 590 F. O_2 flowrate = 1272 tpd; Steam flowrate: 12000 lb/hr.	180 tubes with a length of 25 ft. ID = 2", Pitch = 3". Shell Dia = 3.8 ft.

Area 200 Coal Feeding

Identification	No.	Description	Comments	Unit Size
T-201	4	Weigh Hopper		
D-202	4	Lock Hopper		
D-203	4	Surge Hopper		
S-201	4	Rotary Feeder		
S-203	4	Screw Feeder		

Area 300 Gasification

Identification	No.	Description	Comments	Unit Size
R-301	2	Gasifier		
		Refractory		
		Internals		
H-301	2	Startup Heater		

Area 400 Dust Removal

Identification	No.	Description	Comments	Unit Size
CY-401	2	Primary Cyclone		
CY-402	2	Secondary Cyclone		
CY-403	2	Tertiary Cyclone		
		Refractory		
		Connecting Refractory Pipe		
D-401	2	Cyclone Surge Hopper		
		Refractory		
S-401	2	Cyclone Transport Screw		
D-402	2	Cyclone Lock Hopper		
T-403	2	Cyclone Pneumatic Transport Hopper		
S-403	2	Dust Feeder		
T-404	2	Dust Storage Silo		

Area 500 Ash Removal

Identification	No.	Description	Comments	Unit Size
D-501	2	Ash Surge Hopper Refractory		
S-501	2	Ash Transport Screw		
D-502	2	Ash Lock Hopper		
T-503	2	Ash Pneumatic Transport Hopper		
S-503	2	Ash Feeder		
T-504	2	Ash Storage Silo		

Area 600 Gas Cooling

Identification	No.	Description	Comments	Unit Size	
P-601 NA/SA	4	High Pressure BFW Pump	572 gpm, 965 psi dp, 181.3 bhp		
D-601 NA/SA	2	High Pressure Steam Drum	CS Horizontal, P_{des} =1020 psig T_{des} = 547F	Dia=9.5ft	L=28.5ft
E-601 NA/SA	2	High Pressure Steam Boiler	SHELL: DP= 1135 psig; DT= 600 F; TUBE: DP= 480 psig;DT= 1650 F; Inconel Tubes	6750 sq ft	
P-602	2	HP Steam Boiler Start-up Pump	flowrate: 250 gpm; head = 100 ft, 40hp		
D-602	2	Mid Pressure Steam Drum	CS Horizontal, P_{des} =567 psig T_{des} = 520F	Dia=4.7ft	L=14.0.ft
E-602	2	Mid Pressure Steam Boiler	SHELL: DP= 550 psig; DT= 350 F; TUBE: DP= 450 psig;DT= 700 F; CS Tubes	9365 sq ft	
P-603	4	500 psig BFW pump	238 gpm, 450 psi dp, 24.8 bhp		
P-604	2	MP Steam Boiler Start-up Pump	flowrate: 250 gpm; head = 100 ft, 5hp		

Area 650 Particulate Removal

Identification	No.	Description	Comments	Unit Size	
F-651 NA/SA	2	Candle Filter Assembly	Vendor quote from Pall Corporation		
P-651	2	Nitrogen Compressor	2-stage, no cooling. Inlet conditions = 500 psia, 219^{o}F, 8.7 acfm Outlet conditions = 1010 psia, 430^{o}F, 5.8 acfm	21.1 BHP	
D-651	2	Nitrogen Drum	CS vertical P_{des} =1120 psig T_{des} = 675F	Dia=4.5ft	L=18ft
D-652	2	Ash Surge Hopper	Equivant to Area 500- same ash removal rate		
D-653	2	Ash Lock Hopper	Equivant to Area 500- same ash removal rate		
T-651	2	Ash Pneumatic Transport Hopper	Equivant to Area 500- same ash removal rate		
S-651	2	Ash Feeder	Equivant to Area 500- same ash removal rate		
T-652	2	Ash Storage Silo	Equivant to Area 500- same ash removal rate		

Area 700 Gas Cleaning

Identification	No.	Description	Comments	Unit Size
C-701	1	Syngas Scrubber Column	Inerts = 410SS, Vessel = CS 8 Trays, P_{des} = 457psig, T_{des}=551F	D=8.3 ft h =33 ft
E-701	1	COS Hydrolysis Reactor Preheater	TEMA Type AEU, 410SS tubes	area =70 ft2
P-701	2	Fresh quench water pump	91 gpm, 400 psi dp, 24 bhp	
E-702	1	COS Reactor Effluent Cooler/BFW Heater	TEMA Type AEU, 410SS tubes	area = 6290 ft2
E-703	1	Effluent air cooler	SS Construction, Fan HP = 160	Bare tube area = 7600 ft2
E-704	1	COS Reactor Effluent Water Cooler	TEMA Type AEU, 410SS tubes	area =2895 ft2
P-702	1	HRSG BFW Pump	590 gpm, 1050 psi dp, 189 bhp	
P-703	2	Recycle SWS Water Pump	774 gpm, 400 psi dp, 175.6 bhp	
R-701	1	COS Hydrolysis Reactor		15.0' ID by 18.0' TT
D-701	1	Effluent condensor drum	CS Horizontal, P_{des} =380 psig T_{des} = 160F	Dia=1.5ft L=4.0ft
S-701	Lot	Sud Chemie C53-2-01 1/8" catalyst	1,295 cu ft of catalyst	
R-711	1	Mercury Adsorption Vessel		15.4 ft ID by 28 ft TT
S-711	Lot	Sulfur Impregnated Activated Carbon		163,815 lbs

Area 800 Acid Gas Removal and Sulfur Recovery

Identification	No.	Description	Comments	Unit Size
C-801	1	Amine Absorber Column		
C-802	1	Amine Regenerator Column		
C-803	1	Sour Water Stripper Column	Internals = 410SS, Vessel = CS 24 Trays, Pdes = 50psig, Tdes=320F	D=8.27 ft Tray Spacing=2 ft, T-T=66 FT
D-801	1	Rich Amine Flash Drum		
D-802	1	Lean Amine Carbon Filter		
D-803	1	Amine Regenerator Reflux Drum		
D-804	1	Sour Water 3 Phase Separator	Pdes = 50psig, Tdes=250F, horizontal	D=15.5ft L=31ft
D-805	1	Sour Water 3-Phase Separator Drum	Pdes = 50psig, Tdes=317F, horizontal	D=7.63ft L=10.2ft
E-801	1	Lean Amine/Rich Amine Exchanger		
E-802	1	Lean Amine Cooler		
E-803	1	Amine Regenerator Steam Reboiler		
E-804	1	Amine Regenerator Reflux Condenser		
E-805	1	Sour Water Sub-Cooler - Air Fin	Stand by for Day Tank Pdes = 20psig, Tdes=225F	Area =1,210 ft2 Fan HP = 24
E-806	1	SWS Feed Pre-Heater	Tubes: Pdes = 78psig, Tdes=300F Shell : Pdes = 50psig, Tdes=300F, 410ss TUBES	area = 5,400 ft2
E-807	1	SWS Condensor - Air Fin		Area = 5,070 ft2 Fan HP =101
E-808	1	SWS Kettle Reboiler	Tubes Pdes = 50psig, Tdes=317F Shell : Pdes = 75psig, Tdes=400F, 410ss TUBES	area = 3,434 ft2
E-809	1	Recycle Water Cooler - Air Fin		Area = 5,250 ft2 Fan HP =105
E-810	1	Recycle Water Cooler - Water Cooled	Tubes Pdes = 72psig, Tdes=190F Tubes Pdes = 75psig, Tdes=160F, CS TUBES	area = 1,346 ft2
E-811	1	Sour Water Pre-Cooler - Air Fin	SS Construction	area = 2,010 ft2 Fan HP =40
E-812	1	Rich Amine Filter		
E-813	1	Lean Amine Carbon Bed Inlet Filter		
E-814	1	Lean Amine Carbon Bed Outlet Filter		
E-815	2	Filter Press		
E-816	2	Amine Circulation Pump and Motor		
E-817	2	Regenerator Reflux Pump and Motor		
E-818	1	Amine Transfer Pump and Motor		
E-812	1	Day Tank Reheater	Stand by for Day Tank Pdes = 40psig, Tdes=250F, CS TUBES	area = 302 ft2
P-804A/B	2	SWS Feed Pump		815 gpm, 38psi dp BHP = 19
P-805A/B	2	SWS Reflux Pump		117 gpm, 25 psi dp BHP = 1.8
P-806A/B	2	SWS Bottom Pump		836 gpm, 30 psi dp BHP = 15.4
P-807A/B	2	Pre Day Tank Pump	Stand by for Day Tank	843 gpm, 20 psi dp BHP = 10.3
T-801	1	Fresh Amine Storage Tank		
P-808A/B	2	Post Day Tank Pump	Stand by for Day Tank	815 gpm, 60 psi dp BHP = 30
T-803	1	Sour Water Storage (Day Tank)	CS Vertical, P_{des} =0 psig T_{des} = 225F	1,200,000 gallons (80 ft dia by 32 ft T-T)
S-801	1	Initial Fill of Amine Solution		
S-802	1	Initial Fill of Activated Carbon SRU (installed)		

Area 900 Gas Turbine and HRSG

Identification	No.	Description	Comments	Unit Size
GT-901	1	Syngas Turbine	Gas tubine generator - GE 7FA including lube oil console, static frequency converter, intake air filter, compressor, turbine expander, generator exciter, Mark V control system, and generator control panel	210.78
F-901	1	Final Syngas Filter	see GT-901	
	1	HRSG		1,127,905
901	1	Steam turbine	Steam turbine generator including lube oil console, hydraulic oil EHC system, steam turbine, generator, static exciter, Mark V control system, generator control panel and associated skids	90.84
E-901 A/B/C/D	1	STG Surface Condenser (includes Gland Steam Condenser and Condenser SJAE Skid)	Tubes: Pdes = -12.3psig & 50 PSIG, Tdes=182F Shell : Pdes = 100psig, Tdes=160 F	area = 26,000 ft^2 (total for 4 shells)
E-902	1	Syngas preheater	Tubes Pdes = 600psig, Tdes=525F Shell Pdes = 300psig, Tdes=450F, CS tubes	area = 6,780 ft^2
P-901 A/B	2	BFW pumparound		1200 gpm, 30 psi dp BHP = 25

Area 1000 Offsites and Auxiliaries

Identification	No.	Description	Comments	Unit Size
		Steam generation system	start-up boiler, 1000 psig, 650 F	60,000 lb/hr
		Condensate collection system		660 gpm
		Demineralized water system		1,200 gpm
		Cooling water system		62,000 gpm
		Safety shower/eye wash system		
		Raw water/fire water system	raw water conxumption / firewater rate	2,646,700 gpd / 2,500 gpm
		Drinking (potable) water system		
		Compressed air system		1,300 scfm
		Natural gas supply system		
		Flare system		405 MM Btu/hr
		Nitrogen system		101,000 scfh
		Waste water collection, treatment and disposal system		
		Electrical distribution system		
		Interconnecting piping		
		Telecommunications systems		
		Buildings		
		General Facilities		
		Miscellaneous		

The input values into the financial model are listed below:

- Fees: Per NETL guidelines and past team experience, a 10% fee rate was used to include project development, start-up costs, licensing, permitting, spares, training, construction management, commissioning, transportation, and owner's costs. This fee was entered in the model by placing 4% as the development fee, 2% as start-up costs, and the remainder in "owner's costs". Owner's costs are a dollar value calculated to equal 10% of the plant EPC cost when combined with the development fee and start-up costs.
- Interest during construction, financing costs, and working capital requirements all have been entered separately.
- Unit engineering and installation already are included in the plant EPC costs. These factors were used when scaling-up the as-built unit costs.
- Royalties and land costs have not been included. It is assumed that the land is already owned by the developer. Royalties will vary considerably based on the technology vendor.
- Fixed and Variable Operations and Maintenance Costs: The entries for fixed and variable O&M costs reflect work done on previous gasification studies and NETL guidelines, with a few modifications for the plant location. Typical default values for O&M combine to equal 4.2% of the plant EPC cost. However, this number reflects a United States Gulf Coast (USGC) plant site. Labor rates have been adjusted upwards to reflect the high labor costs in North Dakota.
- Operating Hours: Operating hours are defined as the total hours available for plant operation after scheduled outage time has been deducted. The amount of planned plant downtime for scheduled maintenance, 14 days, is based on the GTI estimate for Subtask 3.2 (i.e., 21 days) adjusted by Nexant to reflect the inclusion of a spare gasifier. The inclusion of a second, spare, gasifier train decreased the amount of planned downtime from 21 days, which was used in Subtask 3.2, to 14 days. It is assumed that other scheduled maintenance can be done during this time. While this number will vary throughout the life of the plant because of plant turnarounds and major maintenance, this value is expected to be the annual average planned plant outages for the life of the facility. The trade-off study evaluating plant operation with no spare assumed 21 days of downtime, similar to what was performed in Subtask 3.2.
- Start-Up Scenarios: The financial model allows the user to enter a different availability for the first two years of plant operation than that for the rest of the plant life. From previous design and operations experience, the total first year availability was estimated to be 69%. The second year of operation was assumed to be the same as that for the remaining years.

Table C.1 Financial Model Entries—Plant Inputs

Project Name	Lignite IGCC w/ Spare
Project Location	North Dakota
Primary Output/Plant Application **(Options: Power, Multiple Outputs)**	Multiple Outputs
Primary Fuel Type (**Options: Gas, Coal, Petroleum Coke, Other/Waste)**	Coal
Secondary Fuel Type (**Options: None, Gas, Coal, Petroleum Coke, Other/Waste)**	None
Plant Output and Operating Data : Note - All ton units are US Short Tons (2000 lbs)	
Syngas Capacity (Mcf/Day)	0
Gross Electric Power Capacity (MW)	301
Net Electric Power Capacity (MW)	251.045
Steam Capacity (Tons/Hr)	0.0
Hydrogen Capacity (Mcf/Day)	0
Carbon Monoxide Capacity (Mcf/Day)	0
Elemental Sulfur Capacity (Tons/Day)	18.7
Slag Ash Capacity (Tons/Day)	277.4
Fuel (Tons/Day)	0
Chemicals (Tons/Day)	0
Environmental Credit (Tons/Day)	0
Operating Hours per Year	8424
Guaranteed Availability (percentage)	90.76%
Enter One of the Following Items(For Each Primary/Secondary Fuel) Depending on Project Type:	
Primary Fuel Heat Rate (Btu/kWh) based on HHV	0
Secondary Fuel Heat Rate (Btu/kWh) based on HHV	0
Primary Fuel Annual Fuel Consumption (in Mcf *OR* Thousand Tons)	1203
Secondary Fuel Annual Fuel Consumption (in Mcf *OR* Thousand Tons)	
Initial Capital and Financing Costs (enter 'Additional Costs' in thousand dollars)	
EPC (in thousand dollars)	410,464
Owner's Contingency (% of EPC Costs)	15.77%
Development Fee (% of EPC Costs)	4%
Start-up (% of EPC Costs)	2%
Owner's Cost (in thousand dollars) **COMBINED WITH DEVELOPMENT AND S/U TO = 10%**	16,419
Operating Costs and Expenses	
Variable O&M (% of EPC Cost)	1.5%
Fixed O&M Cost (% of EPC Cost)	3.5%
Additional Comments	North Dakota U-Gas IGCC power generation facility

Table C.2 Financial Model Entries—Scenario Inputs

Capital Structure	
Percentage Debt	66%
Percentage Equity	34%
Project Debt Terms	
Loan 1: Senior Debt	
% of Total Project Debt **(total for Loans 1,2, and 3 must = 100%)**	100%
Interest Rate	8%
Financing Fee	3%
Repayment Term (in Years)	15
Grace Period on Principal Repayment	1
First Year of Principal Repayment	2010
Loan Covenant Assumptions	
Interest Rate for Debt Reserve Fund (DRF)	4%
Debt Reserve Fund Used on Senior Debt **(Options: Yes or No)**	No
Percentage of Total Debt Service used as DRF	20%
Depreciation (Straight-Line)	
Construction (Years) : *Note - DB Method Must be 15 or 20 years*	15
Financing (Years) : *Note - DB Method Must be 15 or 20 years*	15
Working Capital	
Days Receivable	30
Days Payable	30
Annual Operating Cash (Thousand $)	$50
Initial Working Capital (% of first year revenues)	7%
ECONOMIC ASSUMPTIONS	
Cash Flow Analysis Period	
Plant Economic Life/Concession Length (in Years)	20
Discount Rate	10%
Escalation Factors	
Project Output/Tariff	
Electricity: Capacity Payment	3.0%
Electricity: Energy Payment	3.0%
Steam	3.0%
Elemental Sulfur	3.0%
Slag Ash	3.0%
Fuel/Feedstock	
Gas	4.0%
Coal	2.0%
Petroleum Coke	2.0%
Other/Waste	2.0%
Operating Expenses and Construction Items	
Variable O&M	3.0%
Fixed O&M	3.0%
Other Non-fuel Expenses	3.0%
EPC Costs	3.0%
Tax Assumptions	
Tax Holiday (in Years)	0
Income Tax Rate	40%
Subsidized Tax Rate (used as investment incentive)	0%
Length of Subsidized Tax Period (in Years)	0

FUEL/FEEDSTOCK ASSUMPTIONS

Fuel Prices : For the Base Year, then escalated by fuel factors in B71-B74 above	
Gas ($/Mcf)	6.00
Coal ($/US Short Ton)	9.29
Petroleum Coke ($/US Short Ton)	0.00
Other/Waste ($/US Short Ton)	0.00
Alternatively, use Forecasted Prices (From Fuel Forecasts Sheet)? (Yes/No)	No

TARIFF ASSUMPTIONS

INITIAL TARIFF LEVEL (In Dollars in the first year of construction)	
Electricity Payment ($/MWh)	60.82
Elemental Sulfur ($/US Short Ton)	26.52
Slag Ash ($/US Short Ton)	10.00

CONSTRUCTION ASSUMPTIONS (Base Year: 2005)	
Construction Schedule	
Construction Start Date	7/1/2005*
Construction Period (in months)	42
Plant Start-up Date *(must start on January 1)*	1/1/2009*
EPC Cost Escalation in Effect? (Yes/No)	No

* Note: These dates are used in the financial model only to provide a common ground to compare the economics versus Subtasks 3.2 and 3.3. The actual expected start-up date is 1/1/2015.

Percentage of Cost for Construction Periods	**Four Year Period**			
	Year 1	**Year 2**	**Year 3**	**Year 4**
Capital Costs: Unescalated Allocations	10.0%	30.0%	30.0%	30.0%
EPC Costs: Escalated Allocations	23.9%	24.9%	25.4%	25.9%
EPC Costs	0.0%	0.0%	0.0%	100.0%
Initial Working Capital	0.0%	0.0%	0.0%	100.0%
Owner's Contingency (% of EPC Costs)	35.0%	35.0%	30.0%	0.0%
Development Fee (% of EPC Costs)	0.0%	30.0%	70.0%	0.0%
Start-up (% of EPC Costs)	0.0%	30.0%	70.0%	0.0%
Initial Debt Reserve Fund	0.0%	30.0%	70.0%	0.0%
Owner's Cost (in thousand dollars) COMBINED WITH DEVELOPMENT AND S/U TO = 10%	0.0%	30.0%	70.0%	0.0%
Interest During Construction	0.0%	30.0%	70.0%	0.0%
Financing Fee	0.0%	30.0%	70.0%	0.0%
Additional Financing Cost #1	0.0%	30.0%	70.0%	0.0%
Plant Ramp-up Option (Yes or No)	Yes			

Start-Up Operations Assumptions (% of Full Capacity)	
Year 1, First Quarter	50%
Year 1, Second Quarter	65%
Year 1, Third Quarter	75%
Year 1, Fourth Quarter	85%
Year 1 Average Capacity %	69%
Year 2, First Quarter	91%
Year 2, Second Quarter	91%
Year 2, Third Quarter	91%
Year 2, Fourth Quarter	91%
Year 2 Average Capacity %	91%

Figure D.1 Simplified Flow Diagram - Coal Handling System

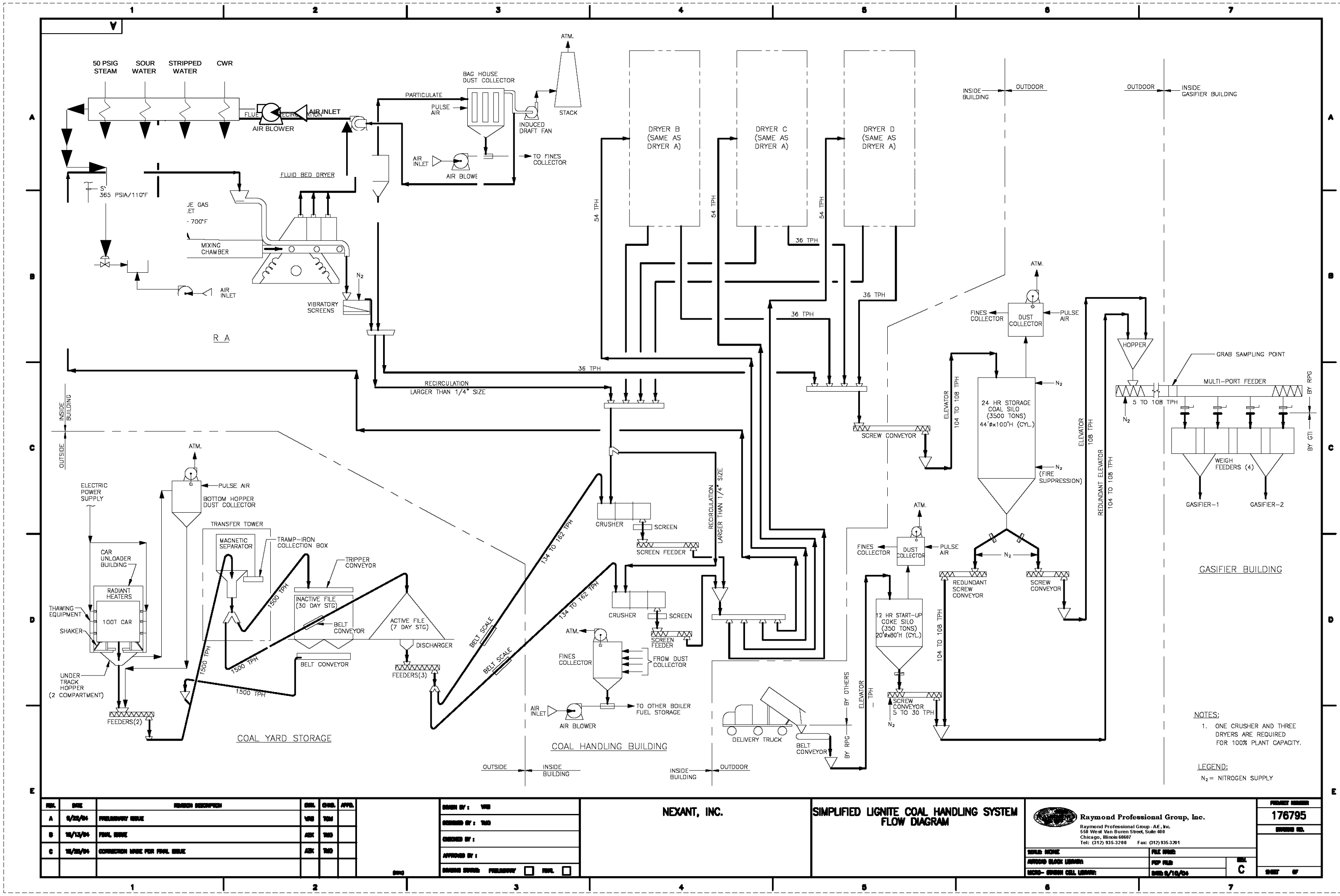


Figure D.2 Air Separation Unit

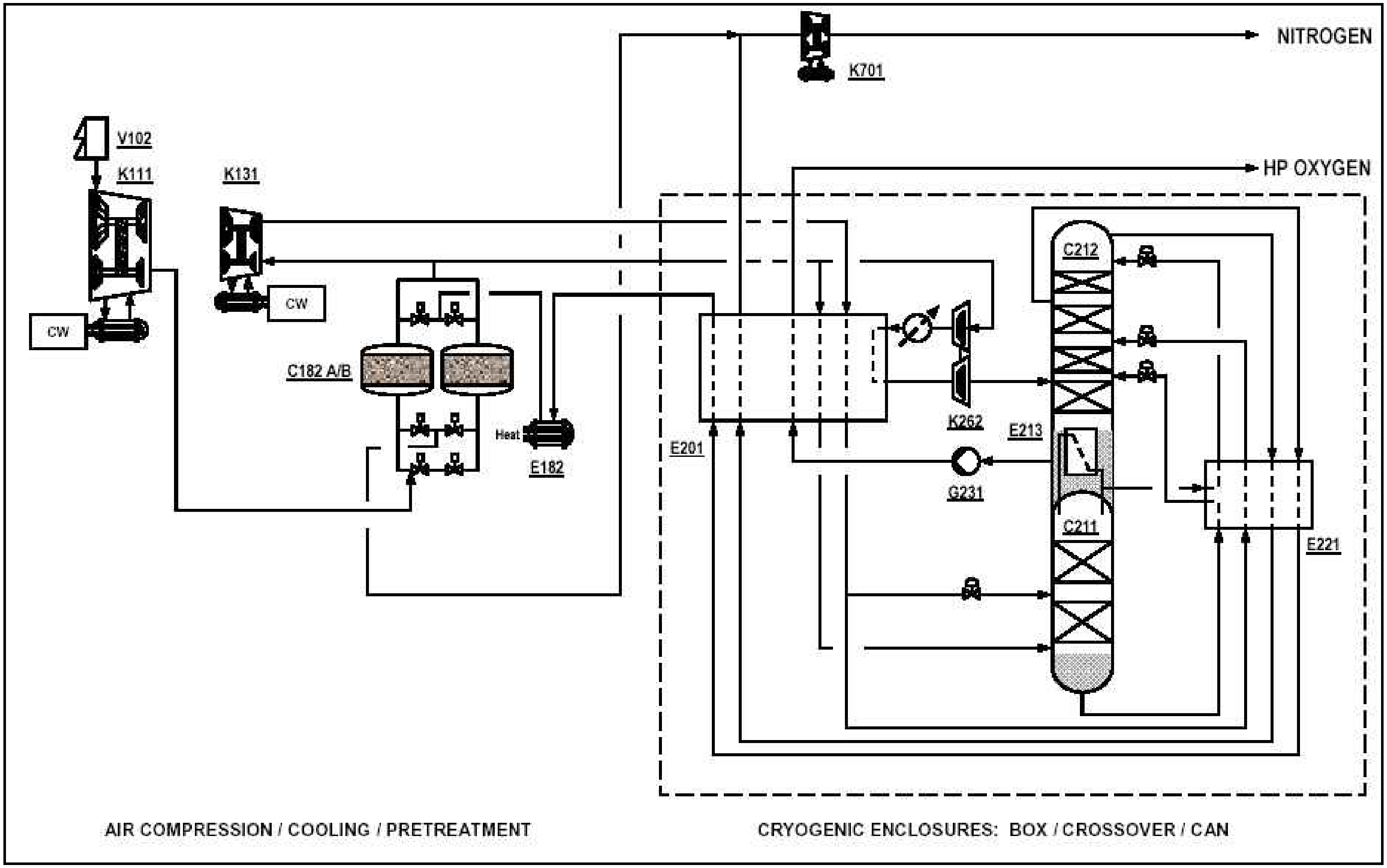
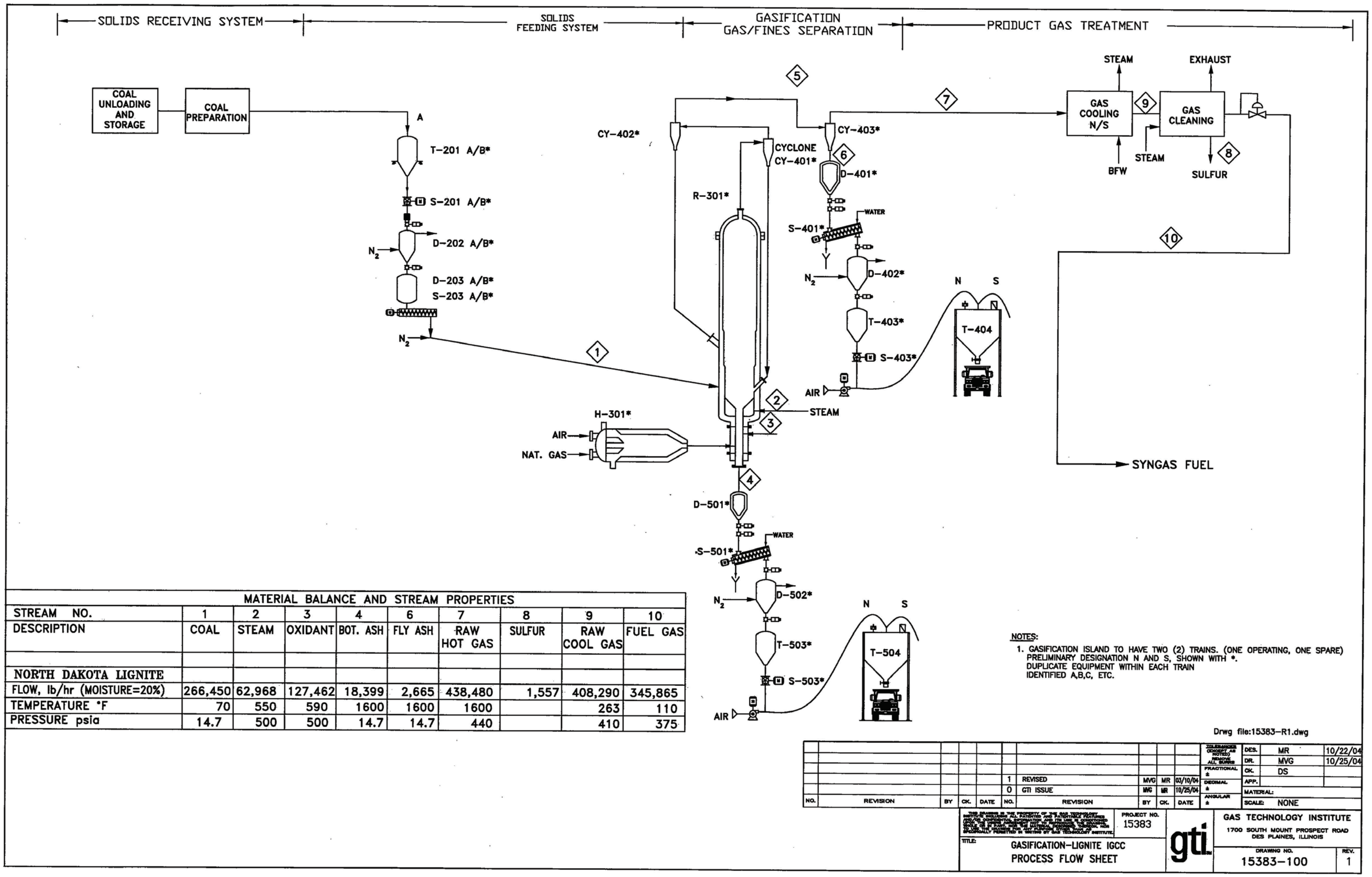


Figure D.3 Gasification – IGCC Process Flow Sheet



MATERIAL BALANCE AND STREAM PROPERTIES

STREAM NO.	1	2	3	4	6	7	8	9	10
DESCRIPTION	COAL	STEAM	OXIDANT	BOT. ASH	FLY ASH	RAW HOT GAS	SULFUR	RAW COOL GAS	FUEL GAS
NORTH DAKOTA LIGNITE									
FLOW, lb/hr (MOISTURE=20%)	266,450	62,968	127,462	18,399	2,665	438,480	1,557	408,290	345,865
TEMPERATURE °F	70	550	590	1600	1600	1600		263	110
PRESSURE psia	14.7	500	500	14.7	14.7	440		410	375

Figure D.4 Heat Recovery and Gas Clean Up Process Flow Sheet

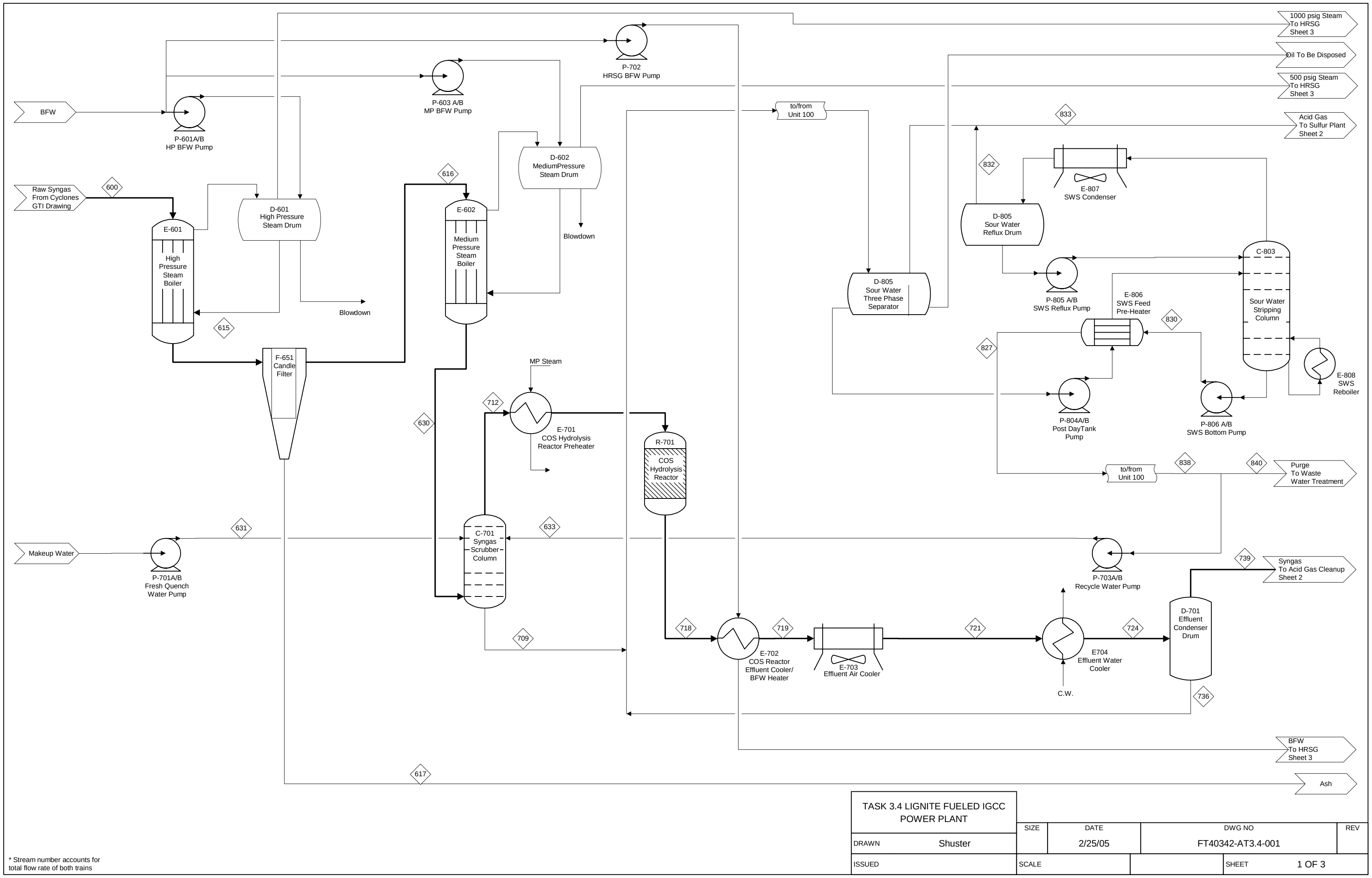
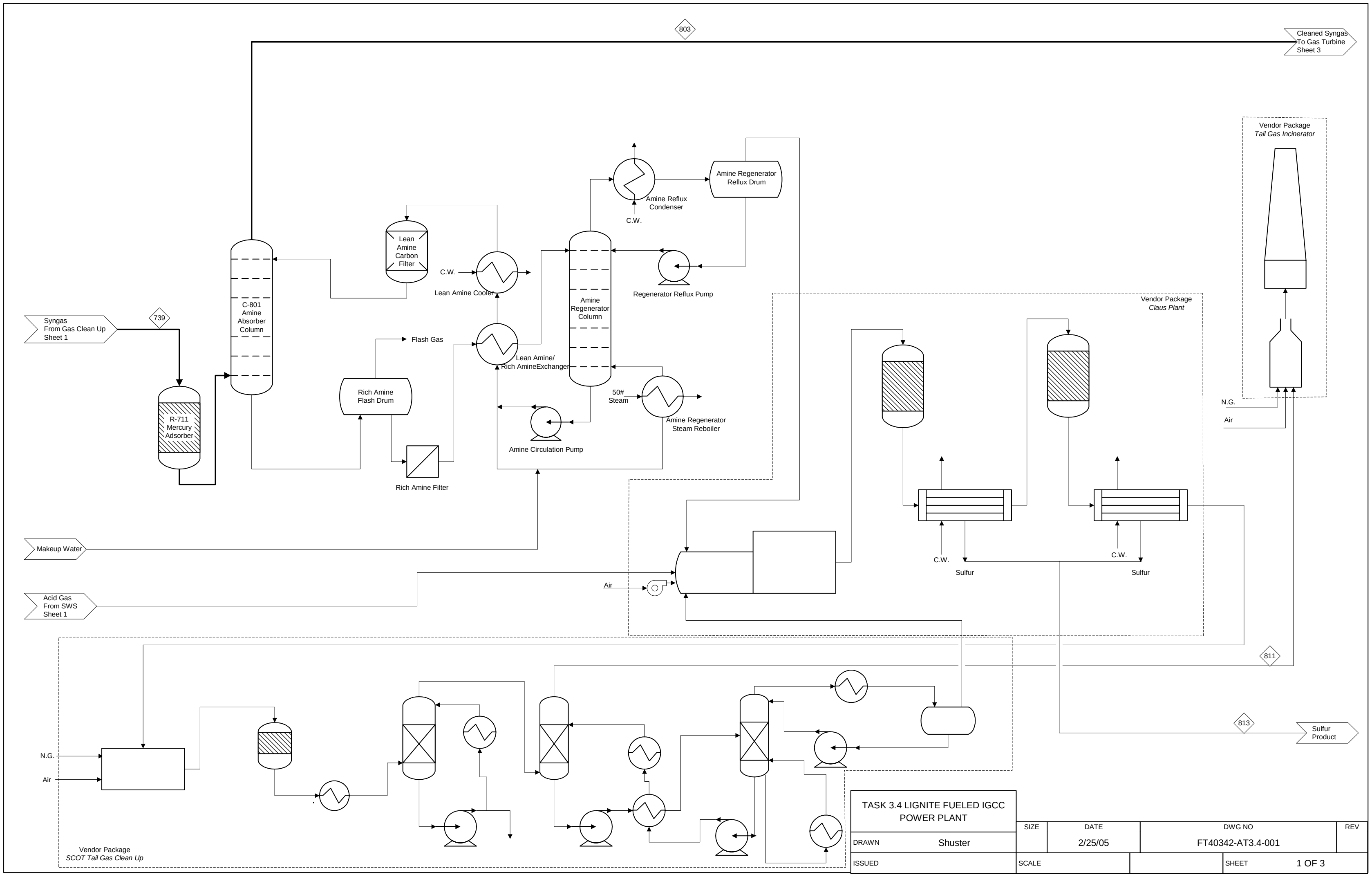


Figure D.5 Mercury and Acid Gas Removal Process Flow Sheet



TASK 3.4 LIGNITE FUELED IGCC POWER PLANT	SIZE	DATE	DWG NO	REV
DRAWN Shuster		2/25/05	FT40342-AT3.4-001	
ISSUED	SCALE		SHEET 1 OF 3	

Figure D.6 Gas Turbine & Gas Recovery Steam Generation Process Flow Sheet

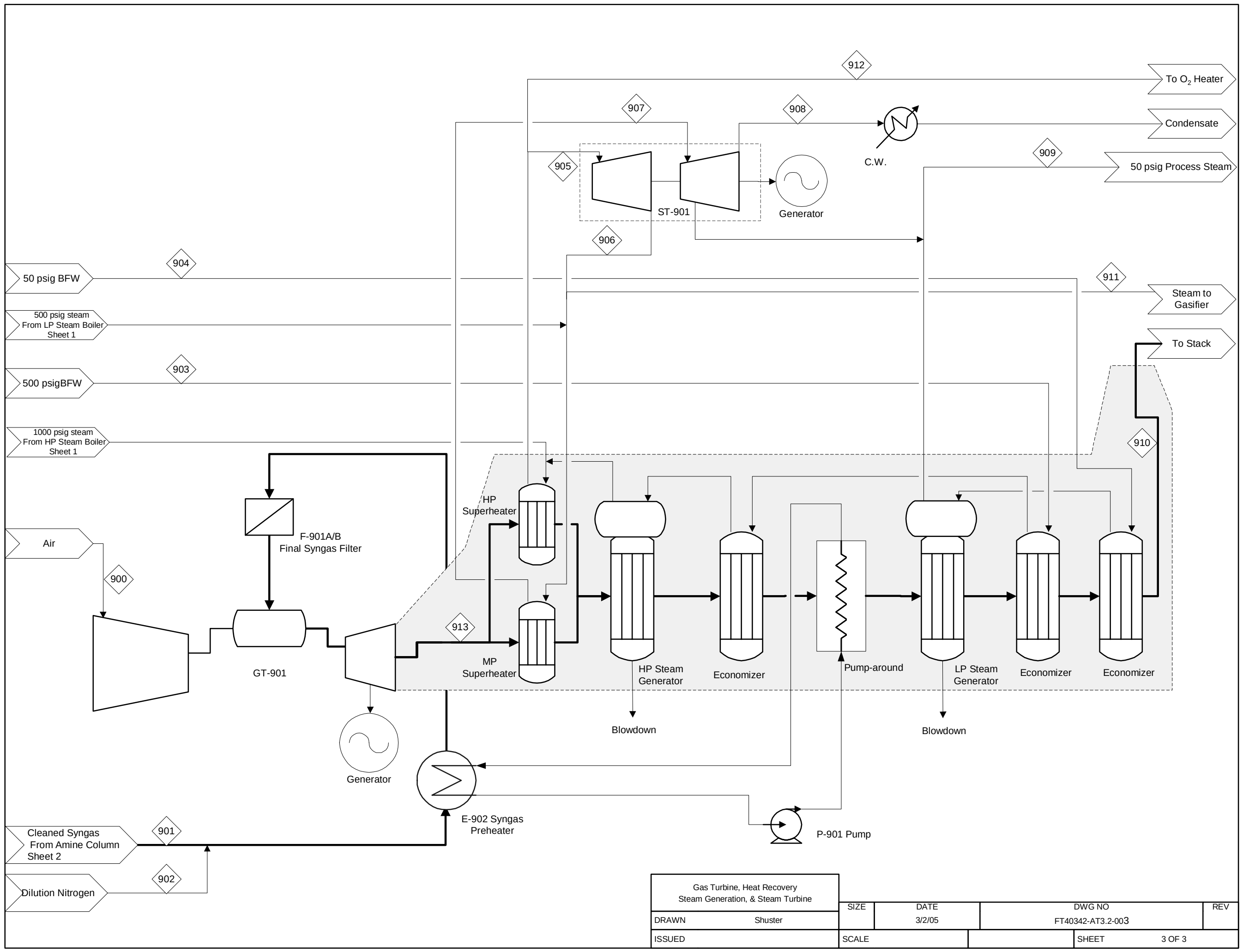


Table D.1 Gasifier Island Material and Energy Balance

(one of two gasifiers)

Stream No.	1	2	3	4	5	6
Stream Description	**Coal**	**Steam**	**Oxidant**	**Bottom Ash**	**Raw Syngas**	**Fly Ash**
Stream Composition, lb/h						
CO					186,032	
CO2					148,450	
H2					9,971	
H2O	53,290	62,968	0		56,391	
CH4					22,445	
H2S					1,754	
COS*					77	
NH3					674	
HCN					43	
N2			5,615		7,227	
O2			121,847			
HCl					88	
Coal/residue [1]	192,441			1,785	1,224	612
Mineral Matter/Ash	20,719			16,614	4,105	2,052
Total, lb/h	**266,450**	**62,968**	**127,462**	**18,399**	**438,480**	**2,665**
Temperature, F	70	550	590	1600	1600	1600
Pressure, psia	14.7	500	500	14.7	439.7	14.7

[1] mixture of mostly carbon plus unconverted oxygen, hydrogen, nitrogen, and sulfur

Table D.2a Gas Cooling & Cleaning Material and Energy Balance

Stream ID	600	615	616	617	630	631	633	636
Description	Syngas to High Temp HR	Syngas to Candle Filter	Syngas to Mid Temp HR	Ash Discharge	Syngas to Water Scrubber	Makeup to Water Scrubber	From SWS Recycle Pump	Syngas + Scrubber Water
Temperature F	1,500	650	650	650	501	80	110	263
Pressure psi	440	430	425	425	415	415	435	415
Vapor Frac	1.00	1.00	1.00		1.00	0.00	0.00	0.46
Mole Flow lbmol/hr	19,844	19,844	19,844	0	19,844	2,174	18,235	40,253
Mass Flow lb/hr	435,816	435,816	433,151	2,665	433,151	39,160	328,500	350,166
Volume Flow cuft/hr	955,170	551,730	558,172	0	493,018	740	6,279	350,166
Enthalpy MBtu/hr	-1,011	-1,167	-1,167		-1,192	-269	-2,242	-3,703
Density lb/cuft	0.45	0.79	0.78		0.88	52.92	52.31	2.29
Mass Flow lb/hr								
CO	186,032	186,032	186,032	0	186,032	0	0	186,032
CO2	148,450	148,450	148,450	0	148,450	0	0	148,450
H2	9,971	9,971	9,971	0	9,971	0	0	9,971
H2O	56,391	56,391	56,391	0	56,391	39,160	328,500	424,051
CH4	22,445	22,445	22,445	0	22,445	0	0	22,445
H2S	1,754	1,754	1,754	0	1,754	0	0	1,754
COS	77	77	77	0	77	0	0	77
H3N	674	674	674	0	674	0	0	674
CHN	43	43	43	0	43	0	0	43
N2	7,227	7,227	7,227	0	7,227	0	0	7,227
O2	0	0	0	0	0	0	0	0
HCL	88	88	88	0	88	0	0	88
ASH	2,665	2,665	0	2,665	0	0	0	0
Mole Flow lbmol/hr								
CO	6,642	6,642	6,642	0	6,642	0	0	6,642
CO2	3,373	3,373	3,373	0	3,373	0	0	3,373
H2	4,946	4,946	4,946	0	4,946	0	0	4,946
H2O	3,130	3,130	3,130	0	3,130	2,174	18,235	23,538
CH4	1,399	1,399	1,399	0	1,399	0	0	1,399
H2S	51	51	51	0	51	0	0	51
COS	1	1	1	0	1	0	0	1
H3N	40	40	40	0	40	0	0	40
CHN	2	2	2	0	2	0	0	2
N2	258	258	258	0	258	0	0	258
O2	0	0	0	0	0	0	0	0
HCL	2	2	2	0	2	0	0	2
ASH			0		0	0	0	0

Table D.2b Gas Cooling & Cleaning Material and Energy Balance

Stream ID	709	712	718	721	724	736	739	800	803
Description	Sour Water to SWS	Syngas from Water Scrubber	Syngas from COS Hydrolysis	Syngas from Air Fin Cooler	Syngas from Low Temp HR	Process Condensate	Syngas from Process Condenser	To Sulfur Recovery Unit	Cleaned Syngas
Temperature F	263	263	275	140	110	111	110	110	110
Pressure psi	410	410	390	380	375	375	375	375	375
Vapor Frac	0.00	1.00	1.00	0.91	0.91	0.00	1.00	1.00	1.00
Mole Flow lbmol/hr	21,775	18,477	18,477	408,290	18,477	1,757	16,720	714	16,006
Mass Flow lb/hr	392,521	408,290	408,290	408,290	408,290	31,650	376,641	30,775	345,865
Volume Flow cuft/hr	8,075	346,065	370,563	-1,143	269,358	606	268,670	10,200	257,939
Enthalpy MBtu/hr	-2,609	-1,094	-1,092	1	-1,149	-215	-934	-113	-822
Density lb/cuft	48.61	1.18	1.10	1.45	1.52	52.19	1.40	3.02	1.34
Mass Flow lb/hr									
CO	11	186,021	186,021	186,021	186,021	0	186,021	0	186,021
CO2	220	148,231	148,287	148,287	148,287	6	148,280	29,097	119,184
H2	2	9,969	9,969	9,969	9,969	0	9,969	0	9,969
H2O	391,518	32,532	32,509	32,509	32,509	31,431	1,078	0	1,078
CH4	7	22,437	22,437	22,437	22,437	0	22,437	110	22,327
H2S	200	1,554	1,597	1,597	1,597	13	1,584	1,568	16
COS	0	77	1	1	1	0	1	1	0
H3N	474	200	200	200	200	199	1	0	1
CHN	0	43	43	43	43	0	43	0	43
N2	0	7,227	7,227	7,227	7,227	0	7,227	0	7,227
O2	0	0	0	0	0	0	0	0	0
HCL	88	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0
Mole Flow lbmol/hr									
CO	0	6,641	6,641	6,641	6,641	0	6,641	0	6,641
CO2	5	3,368	3,369	3,369	3,369	0	3,369	661	2,708
H2	1	4,945	4,945	4,945	4,945	0	4,945	0	4,945
H2O	21,733	1,806	1,805	1,805	1,805	1,745	60	0	60
CH4	0	1,399	1,399	1,399	1,399	0	1,399	7	1,392
H2S	6	46	47	47	47	0	46	46	0
COS	0	1	0	0	0	0	0	0	0
H3N	28	12	12	12	12	12	0	0	0
CHN	0	2	2	2	2	0	2	0	2
N2	0	258	258	258	258	0	258	0	258
O2	0	0	0	0	0	0	0	0	0
HCL	2	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0

Table D.3 Sour Water Stripper Material and Energy Balance

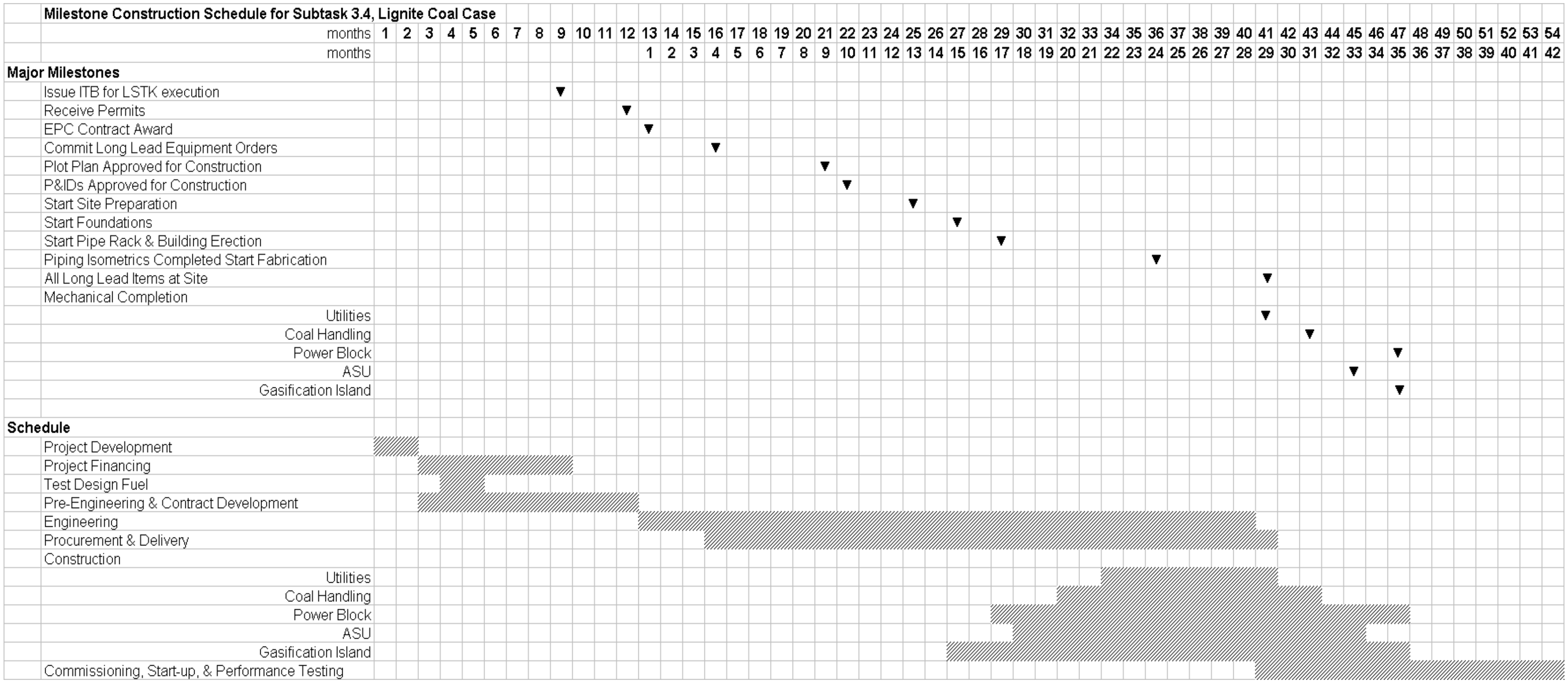
	709	736	820	821	822	829	830	831	832	833	834	835	836	837	838	840	841
Description	Sour Water from Water Wash	Process Condensate	Mixed input	Cooled input	Overhead from flash	3 Phase Sep to pump	Pump to stripper preheater	To sour water stripper	Overhead from stripper column	Mixed vapor stream to sulfur plant	Stripped water from stripper column	Stripped water to stripper preheater	Stripped water to air fin cooler	Stripped water to water cooler	Stripped water from cooling train	Purge stream to water treatment	Liquid Reflux
Temperature F	267	111	255	200	200	200	200	252	201	201	267	267	215	140	110	110	201
Pressure psi	410	375	375	374	35	35	38	38	35	38	40	70	70	65	60	60	35
Vapor Frac	0.0	0.0	0.0	0.0	1.0	0.0	0.0	0.0	1.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Mole Flow lbmol/hr	21,597	1,939	23,536	23,536	1	23,536	23,536	23,536	69	69	23,467	23,467	23,467	23,468	23,468	5,451	3,017
Mass Flow lb/hr	389,236	34,935	424,171	424,171	7	424,164	424,164	424,164	1,399	1,406	422,765	422,765	422,765	422,774	422,774	98,202	55,967
Volume Flow cuft/hr	6,688	567	7,251	10,030	166	7,069	7,069	7,869	13,725	12,790	7,251	7,251	7,076	6,887	6,833	1,587	67,256
Enthalpy MBtu/hr	-2,578.42	-236.12	-2,814.54	-2,838.21	-0.03	-2,838.18	-2,838.17	-2,815.84	-3.75	-3.78	-2,804.66	-2,804.61	-2,826.94	-2,858.86	-2,871.53	-667.00	-329.94
Density lb/cuft	58.20	61.64	58.49	42.29	0.04	60.01	60.01	53.90	0.10	0.11	58.30	58.31	59.74	61.39	61.87	61.87	55.46
Mass Flow lb/hr																	
CO	11.1	0.1	11.2	11.2	0.3	10.9	10.9	10.9	10.9	11.2	0.0	0.0	0.0	0.0	0.0	0.0	0.5
CO2	219.1	6.8	225.9	225.9	0.2	225.8	225.8	225.8	225.8	225.9	0.0	0.0	0.0	0.0	0.0	0.0	2,377.0
H2	2.0	0.0	2.0	2.0	1.0	1.0	1.0	1.0	1.0	2.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	388,408.2	34,702.0	423,110.2	423,110.2	4.9	423,105.3	423,105.3	423,105.3	356.3	361.2	422,749.0	422,749.0	422,749.0	422,758.7	422,758.7	98,198.4	46,538.5
CH4	7.3	0.1	7.4	7.4	0.0	7.4	7.4	7.4	7.4	7.4	0.0	0.0	0.0	0.0	0.0	0.0	3.7
H2S	51.5	10.0	61.5	61.5	0.0	61.5	61.5	61.5	61.2	61.2	0.3	0.3	0.3	0.3	0.3	0.1	1,014.7
COS	0.4	0.0	0.4	0.4	0.0	0.4	0.4	0.4	0.4	0.4	0.0	0.0	0.0	0.0	0.0	0.0	1.3
H3N	449.2	215.4	664.6	664.6	0.1	664.5	664.5	664.5	649.4	649.5	15.1	15.1	15.1	15.1	15.1	3.5	5,898.1
CHN	0.2	0.0	0.3	0.3	0.0	0.3	0.3	0.3	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.0	2.1
N2	0.4	0.0	0.4	0.4	0.3	0.2	0.2	0.2	0.2	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HCL	86.6	0.0	86.6	86.6	0.1	86.5	86.5	86.5	86.5	86.6	0.0	0.0	0.0	0.0	0.0	0.0	131.2
Mole Flow lbmol/hr																	
CO	0.4	0.0	0.4	0.4	0.0	0.4	0.4	0.4	0.4	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	5.0	0.2	5.1	5.1	0.0	5.1	5.1	5.1	5.1	5.1	0.0	0.0	0.0	0.0	0.0	0.0	54.0
H2	1.0	0.0	1.0	1.0	0.5	0.5	0.5	0.5	0.5	1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	21,559.9	1,926.3	23,486.2	23,486.2	0.3	23,485.9	23,485.9	23,485.9	19.8	20.1	23,466.1	23,466.1	23,466.1	23,466.7	23,466.7	5,450.8	2,583.3
CH4	0.5	0.0	0.5	0.5	0.0	0.5	0.5	0.5	0.5	0.5	0.0	0.0	0.0	0.0	0.0	0.0	0.2
H2S	1.5	0.3	1.8	1.8	0.0	1.8	1.8	1.8	1.8	1.8	0.0	0.0	0.0	0.0	0.0	0.0	29.8
COS	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H3N	26.4	12.6	39.0	39.0	0.0	39.0	39.0	39.0	38.1	38.1	0.9	0.9	0.9	0.9	0.9	0.2	346.3
CHN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1
N2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HCL	2.4	0.0	2.4	2.4	0.0	2.4	2.4	2.4	2.4	2.4	0.0	0.0	0.0	0.0	0.0	0.0	3.6

Table D.4 GT/HRSG/ST Material and Energy Balance

Stream ID	900	901	902	903	904	905	906	907	908	909	910	911	912
Description	Air	Cleaned Syngas	Diluent Nitrogen	1000 Psig Preheated BFW	50 Psig BFW	Superheated Steam to HP Turbine	Steam Exiting HP Turbine	Superheated Steam to LP Turbine	Condensate from LP Turbine	50 Psig Superheated Steam to Plant	Flue Gas To Stack	Steam to Gasifier	Steam to O2 Heater
Temperature F	41	120	291	181	150	1,050	876	1,050	116	345	254	686	1,050
Pressure psi	13.7	433	470	1,050	75	1,025	550	535	1.5	75	13.9	525	1,025
Vapor Frac	1	1	1	0	0	1	1	1	0	1	1	1	1
Mass Flow lb/hr	3,291,500	345,851	300,000	451,182	133,298	594,524	594,524	5,567,804	535,780	150,966	3,937,351	65,000	13,790
Volume Flow cuft/hr	44,454,100	227,078	185,557	8,881	2,591	497,492	824,887	9,131,133	10,268	938,878	74,010,900	78,733	11,539
Enthalpy MBtu/hr	-99	-821	16	-3044	-904	-3174	-3220	-29634	-3654	-854	-2525	-359	-74
Density lb/cuft	0.074	1.52	1.62	50.80	51.44	1.20	0.721	0.610	52.18	0.161	0.053	0.826	1.20
Mass Flow lb/hr													
AR	42,200	0	0	0	0	0	0	0	0	0	42,200	0	0
O2	759,347	0	0	0	0	0	0	0	0	0	484,892	0	0
N2	2,477,270	7,272	300,000	0	0	0	0	0	0	0	2,784,543	0	0
CO2	1,594	119,178	0	0	0	0	0	0	0	0	474,280	0	0
CO	0	186,012	0	0	0	0	0	0	0	0	0	0	0
H2	0	9,969	0	0	0	0	0	0	0	0	0	0	0
CH4	0	22,326	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	30	0	0
H2S	0	16	0	0	0	0	0	0	0	0	0	0	0
H2O	11,088	1,078	0	451,182	133,298	594,524	594,524	5,567,804	535,780	150,966	151,406	65,000	13,790
Mole Flow lbmol/hr													
AR	1,056	0	0	0	0	0	0	0	0	0	1,056	0	0
O2	23,730	0	0	0	0	0	0	0	0	0	15,153	0	0
N2	88,431	260	10,709	0	0	0	0	0	0	0	99,400	0	0
CO2	36	2,708	0	0	0	0	0	0	0	0	10,777	0	0
CO	0	6,641	0	0	0	0	0	0	0	0	0	0	0
H2	0	4,945	0	0	0	0	0	0	0	0	0	0	0
CH4	0	1,392	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0.464	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	615	60	0	25,044	7,399	33,001	33,001	309,060	29,740	8,380	8,404	3,608	765

Figure E.1 shows the Project Construction Schedule. Project completion, as defined by completed performance testing, will occur 42 months after the award of the EPC contract.

Figure E.1 Project Construction Schedule



F.1 INTRODUCTION

This Design Basis defines the process units and process support units to set the basic plant configuration for Subtask 3.4. This section includes the design basis and criteria for the subsequent engineering study and capital cost estimate. Subtask 3.4 is a base case design for a power plant using North Dakota lignite. Subtask 3.4 is defined as follows:

- Develop a design for a base case integrated coal gasification combined cycle (IGCC) power plant in North Dakota using GTI's U-GAS® gasification technology based on best engineering judgment and a few selected trade-off studies.

F.2 SUBTASK 3.4, PRELIMINARY DESIGN FOR THE LIGNITE CASE

F.2.1 Plant Description

The U-GAS® plant located in North Dakota will consist of the following basic process blocks and subsystems:

- Unit 100: Lignite Preparation - Handling, Sizing and Drying
- Unit 150: Air Separation Unit
- Unit 200 Solids Feeding System
- Unit 300 Gasification
- Unit 400 Fines Separation
- Unit 500: Ash Handling
- Unit 600: High Temperature Heat Recovery
- Unit 650: Particulate Removal
- Unit 700: Water Scrubber, COS Hydrolysis Reactor, Low Temperature Heat Recovery and Mercury Removal
- Unit 800: Sulfur removal and recovery, SWS
- Unit 900: Power Block including one operating General Electric 7FB combustion turbine (CT) with heat recovery steam generator (HSRG). Additional power will be generated by a single steam turbine.
- Unit 1000: Utilities (e.g., instrument and plant air, cooling water systems, firewater system) and other offsites (e.g., flare, DCS, plant roads, buildings, chemical storage)

A block flow diagram of the plant is shown in Figure F.1 in Section F.14 of this addendum.

F.2.2 Site Selection

The North Dakota location is an unspecified, generic site. Critical site issues include:

- Sufficient open space for all equipment
- Distance for power interconnect
- Access for coal storage and handling

The plant shall be a stand-alone facility. Normal infrastructure (electric power, potable water, sanitary sewer, etc.) shall be available at the plant boundary. River water shall be available. The site is assumed to be level and cleared. A generic plot plan will be prepared.

F.2.3 Feedstocks

The key to coal selection is to identify a cost effective candidate fuel for use in North Dakota. Lignite coal from North Dakota best fits these criteria. We will use an existing lignite analysis as representative of the coals from this region. Seeking fuel bids and mine analysis at this time is not practical for the study. The coal analysis, which comes from NETL's Quality Guidelines for Energy System Studies, is given in Section F.4 of this addendum.

Coal delivery to the site is by rail. Drying facilities are designed to handle as-received lignite that contains up to 40% moisture and deliver the feed to the gasifier with a moisture content of about 20%. Normal operations shall assume the as-received lignite feed shall have a moisture content of 32.24%.

Excess fines produced during the coal handling process either will be shipped to a nearby PC plant or sent to a landfill. These fines are too small to be used in the gasifier.

F.2.4 Plant Capacity

The plant capacity will use a single GE 7FB turbine, generating about 211 MW of electric power. Additional power will be generated by a single steam turbine.

F.2.5 Configuration

The plant will have a single operating 100% gasifier vessel with about 2,560 tpd (dry) of coal capacity. The operating pressure of the gasifier will be 440 psia.

F.2.6 Gasification Unit

This plant will contain a single 100% gasifier train. Some natural gas will be used for startup.

Downstream of the gasifier each reactor will have a 3-stage cyclone system followed by a high temperature heat removal system generating 1,000 psig steam, metallic filters, a low temperature heat removal system generating 500 psig steam, and a water scrubber. This will be followed by the final cleanup which will include the removal of sulfur, mercury, and other contaminants.

F.2.7 Air Separation Unit

The gasifier will be oxygen-blown. The air separation unit will produce 95% pure oxygen. There will be no nitrogen, oxygen, or argon export from the facility. The nitrogen produced will be used in the plant.

F.2.8 Power Block

A single combustion turbine (GE 7FB) are specified with a nominal syngas power rating of 211 MW. The turbine will have an associated heat recovery steam generator (HRSG). Additional power will be generated by a single steam turbine.

F.3 SITE CONDITIONS

Location	Williston, North Dakota
Elevation, feet	1,900
Air Temperature, ^{o}F	
Maximum	110
Average Monthly Maximum	83
Annual Average	40.5
Minimum	-50
Relative Humidity, %	
Maximum	84
Minimum	36
Normal	74
Maximum Hourly Rain Fall, in	2.5
Seismic Zone	0
Wind Speed, mph	
Mean Speed	10
Maximum	70

F.4 FEEDSTOCKS

The following table contains the properties of North Dakota lignite from the Beulah-Zap seam that was taken in Mercer County, ND. These properties were taken from NETL's Quality Guidelines for Energy System Studies.[1]

1 National Energy Technology Center, Office of Systems and Policy Support, "Quality Guidelines for Energy System Studies," November 24, 2003.

	Proximate Analysis, wt%	
	As-Received	Dry
Moisture	32.24	0
Ash	6.59	9.72
Volatile Matter	30.45	44.94
Sulfur	0.54	0.80
Fixed Carbon	30.18	44.54
HHV, kJ/kg	17,338	25,588
HHV, Btu/lb	7,454	11,001
LHV, kJ/kg	15,894	24,625
LHV, Btu/lb	6,833	10,587

	Ultimate Analysis, wt%	
	As-Received	Dry
Moisture	32.24	0
Carbon	44.62	65.85
Hydrogen	2.95	4.36
Nitrogen	0.70	1.04
Chlorine	0.03	0.04
Sulfur	0.54	0.80
Ash	6.59	9.72
Oxygen	12.32	18.19

	Sulfur Analysis, wt%	
	As-Received	Dry
Pyritic	---	0.14
Sulfate	---	0.03
Organic	---	0.63

Ash Analysis	Wt%
SiO2	18.4
Al2O3	10.22
Fe2O3	8
TiO2	0.48
CaO	24.72
MgO	7.48
Na2O	7.76
K2O	0.94
SO3	17.55
P2O5	0.48
BaO	0.84
MnO	0.14
SrO	1.12
Total	98.13

Oxidizing Ash Fusion Temperatures, F	
IT	2329
ST	2393
HT	2425
FT	2460

Reducing Ash Fusion Temperatures, F	
IT	2246
ST	2310
HT	2349
FT	2394

F.5 ELECTRIC POWER

Export Power, MW	Maximize
Voltage, kV	230

F.6 EXPORT STEAM PRODUCTION

There shall be no export steam production

F.7 WATER MAKE-UP

Source	River Water
Supply Pressure, psig	14.7
Supply Temperature, ^{o}F	60

F.8 NATURAL GAS

HHV, Btu/lb	1050
LHV, Btu/lb	960

F.9 BY-PRODUCTS

Ash, tpd	231
Sulfur, tpd	18

F.10 ENVIRONMENTAL GOALS (BASED ON THE DOE TARGET EMISSION AND PERFORMANCE GOALS ESTABLISHED IN THEIR ROADMAP FOR 2010)

SOx	> 99% removal
NOx	< 0.05 lb/M Btu
Particulates	< 0.005 lb/M Btu
Mercury	> 90% removal
Target Efficiency	45-50%

F.11 FINANCIAL

Process Contingency (gasifier block only)	25%
Project Contingency (ex. Gasifier block)	15%
Accuracy	+30/-15%
Capacity Factor	Determined by Section 6 Analysis
Fees (engineering, start-up, owner's costs)	10%
Consumables	Set by EIA and USGS data
O&M	5%
Project, book and tax life	20 years
Tax rate	40%
Debt-to-equity ratio	2:1
Cost of capital	8%

F.12 ANNUAL ESCALATION

Annual escalation will be 3%, with the exception of coal and natural gas. Coal shall have an annual escalation rate of 2%, and natural gas shall have an annual escalation rate of 4%.

F.13 BLOCK FLOW DIAGRAM

Figure F.1 Block Flow Diagram
Part 1 – Syngas Generation and Processing

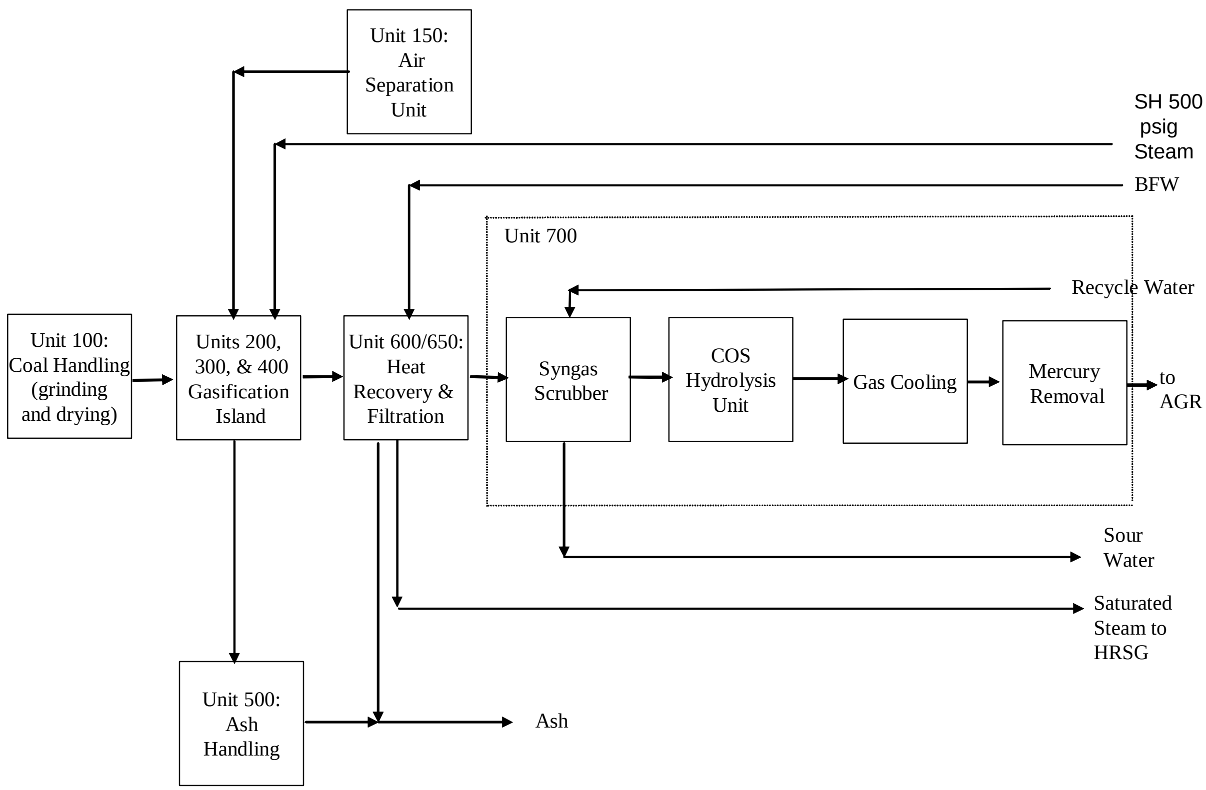


Figure F.1 Block Flow Diagram
Part 2 – Sulfur Removal, Sulfur Recovery and Power Block