Appendix B

Task 3 Gasification Alternatives for Industrial Applications

DOE Contract No
DE-AC26-99FT40342
Subtask 3.3 - Alternate Design for the Eastern Coal Case

Prepared For:

United States Department of Energy / National Energy Technology Laboratory

In association with:

May 2005

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DOE Contract
No. DE-AC26-99FT40342

April 2005

Prepared For:

United States Department of Energy / National Energy Technology Laboratory

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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 used the E-GAS™ gasification technology (now owned by ConocoPhillips). NETL has expanded this effort to evaluate *Gasification Alternatives for Industrial Applications* (here industrial scale is considered to be less than 100 MW). For this effort the Gas Technology Institute (GTI) fluidized bed U-GAS® gasifier was selected for the gasification portion of the plant. 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 Alternatives for Industrial Applications*. This task has two basic 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. 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 gasifier as a stand-alone lignite-fueled IGCC power plant in North Dakota. Subtask 3.4 developed a base case design for that scenario.

This report describes the work performed on the third subtask, Subtask 3.3. Subtask 3.3 developed an alternate design for an air-blown Eastern Coal Case located in upstate New York by considering additional ideas for improving performance and/or reducing investment and operating costs generated during the Value Improving Practices (VIP) sessions. The determination of the exact coal-processing rate was part of this study. This rate was chosen so as to fully load two GE 10 gas turbines, and it is a function of the coal that is processed and the system design.

Table 1.1 summarizes the major input and output streams from the plant along with some key operating parameters and compares them to the Subtask 3.2 base case.[1] Compared to the base case, this alternate case has about 2% less export power, 3.5% more export steam, about 9% less capital investment, a higher overall net CHP efficacy, and a higher return on investment.

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

Table 1.1 Overall Plant Summary

	Alternate (Subtask 3.3)	Base (Subtask 3.2)	Difference
Design Inputs			
Coal Feed, moisture-free tpd	345.7	345.7	0
Coal Feed, moisture-free lb/hr	28,810	28,810	0
Fuel (Natural Gas), MMBtu/hr	-	5.1	-5.1
Makeup Water Input from the Industrial Facility			
Boiler Feed Water, gpm	418	495	-77
Quench Water, gpm	0	30	-30
Cooling Tower Makeup Water, gpm	58	53	+5
Design Outputs			
Export Power, MW	21.3	21.7	-0.4
Export Steam (400 psig, 550°F), Mlb/hr	105.34	101.72	+3.62
Sulfur, lb/hr	899	899	0
Ash, lb/hr	2,719	2,097	+622
Condensate (to industrial facility), Mlb/hr	54.4	60.9	-6.5
EPC Cost, MM$*	82.1	90.0	-7.9
Plant EPC Cost, $/kW**	2,755	3,090	-335
Plant Energy Input, k$/MMBtu/hr	209.7	229.9	-20.2
Plant Energy Output, k$/MMBtu/hr	421.6	469.2	-47.6
Equivalent Availability, %	84.7	85.7	-1.0
Return on Investment, %	8.4	5.9	+2.5
Cold Gas Efficiency, % (HHV basis)	79.3	79.3	0
Net CHP Efficacy, % (HHV basis)	49.7	49.0	+0.7

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

** Based on converting the steam export to power using an average turbine efficiency

For an air-blown facility with EPC costs of 82.1 MM$ and a project life of 20 years, the return on investment (ROI) is expected to be 8.4%, with a net present value (NPV) of -5.3 MM$ (based on a 10% discount factor). Table 1.2 shows the rate of return, NPV, payback year, and required electricity and steam selling prices to obtain a 12% ROI with other items fixed. There are two major products from this facility, electricity and steam, and both must be considered when determining the suitability of this project.

When compared to the Subtask 3.2 air-blown base case, Table 1.2 illustrates the improvements in financial performance of the plant that resulted from the process design improvements incorporated in Subtask 3.3. All parameters in Table 1.2 demonstrate a significant improvement compared to the base case as shown by an increased ROI and a shortened payback period.

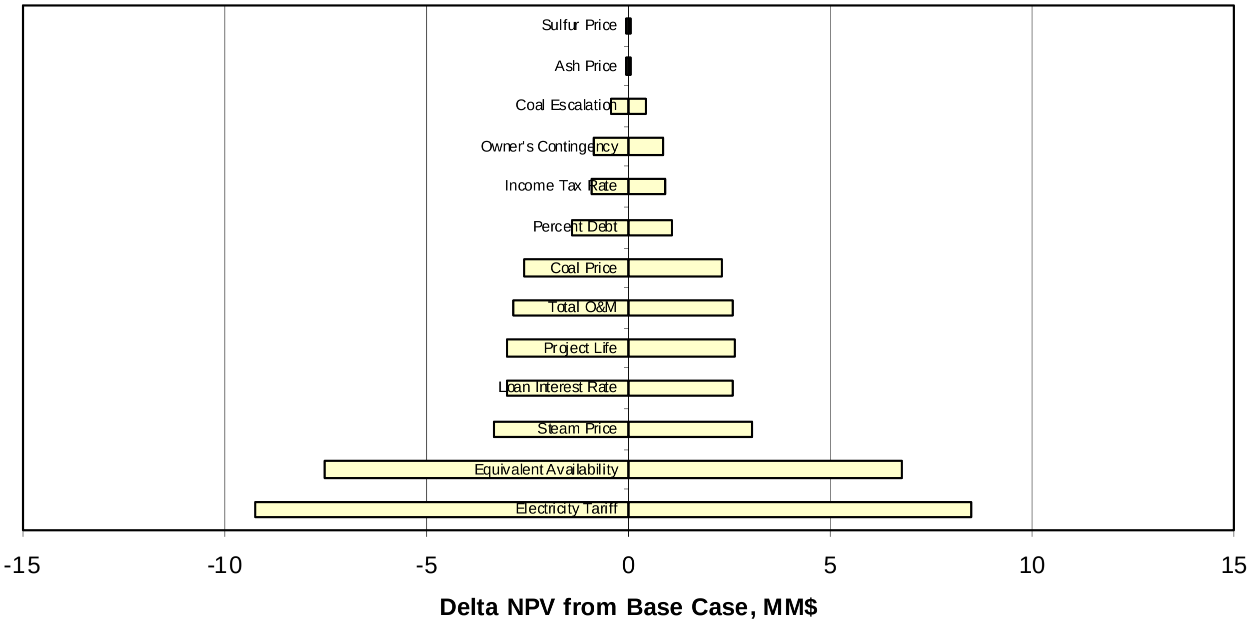
Table 1.2 Financial Cost Summary Comparisons, Subtask 3.3 Alternate Air-Blown Case vs. Subtask 3.2 Air-Blown Base Case

	Subtask 3.3	Subtask 3.2
ROI, %	8.4	5.9
NPV, MM$ (10% Discount Rate)	-5.3	-14.6
Number of years to Payback	14	17
Electricity Selling Price for 12% ROI, cents/kWh	8.5	9.0
Steam Selling Price for 12% ROI, $/ton	14.1	17.6

A number of financial parameters that were likely to influence overall economic performance were varied to determine the project financial sensitivities. Model input changes deemed to be reasonable based on previous sensitivity analysis, commodity input ranges, and team estimates were entered into the model. The impact that these changes had on the NPV and ROI were recorded, along with the percent change to the parameters that were modified. The financial impacts were normalized by calculating the overall impact relative to the size of the modification. The variables and their impact on the financial outputs were then ranked to determine the parameters with the highest sensitivity.

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 varied by ±10%, with the changes of NPV from the base case shown. 10% changes were used to give a common ground by which all variables were evaluated. It is worthwhile to note, however, that the range of realistic possibilities for each variable differs significantly. For example, 10% changes in the availability or income tax rate should capture the majority of long-term variations. This would not be the case with variables such as coal price and electricity tariff which could vary by much more than 10%. The relative significance and range of possible values were considered in determining which items 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 nearly 8 MM$ while decreasing it by 10% results in a decreased net present value by nearly 9 MM$. In this case, "Electricity Tariff" is used to refer to the sales value for the electricity that the plant generates. This variable was also the most significant in Subtask 3.2. The guaranteed availability also is very significant. Although a theoretical increase of the guaranteed availability would result in an unrealistic 100% guaranteed availability, operating at or near 100% would result in a net present value increase of more than 6 MM$. By reducing the availability by 10%, the net present value is reduced by more than 7 MM$. All other variables associated with the amount of time the plant is operating (e.g., operating hours and plant life) also had a significant impact on the plant economics.

As was the case for Subtask 3.2, the remainder of the input variables impacted the plant economics to a lower extent. The steam price, plant life, and interest rate were next in importance, with all other items showing a less significant impact. While the remaining items had a less significant impact relative to those described above, many could push the project to a near zero or negative net present value within the ±10% range evaluated here.

The model relies more heavily on the electricity tariff for the economic outcome due to electricity being 72% of the total revenue stream for the base case facility. Although steam also is a primary product for this facility, the contribution to total revenue is only 27%, making this variable less sensitive to fluctuations in price. Figure 1.2 shows the relationship between the ROI and electricity tariff. The reference power price of 80 $/MWh is indicated by an arrow on the abscissa.

Figure 1.2 Effect of Electricity Tariff on Return on Investment

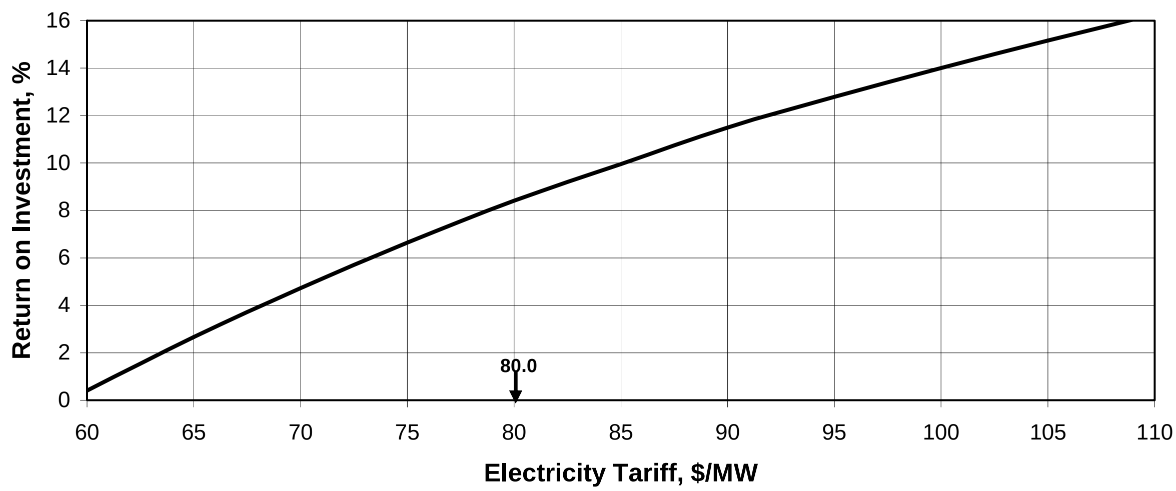
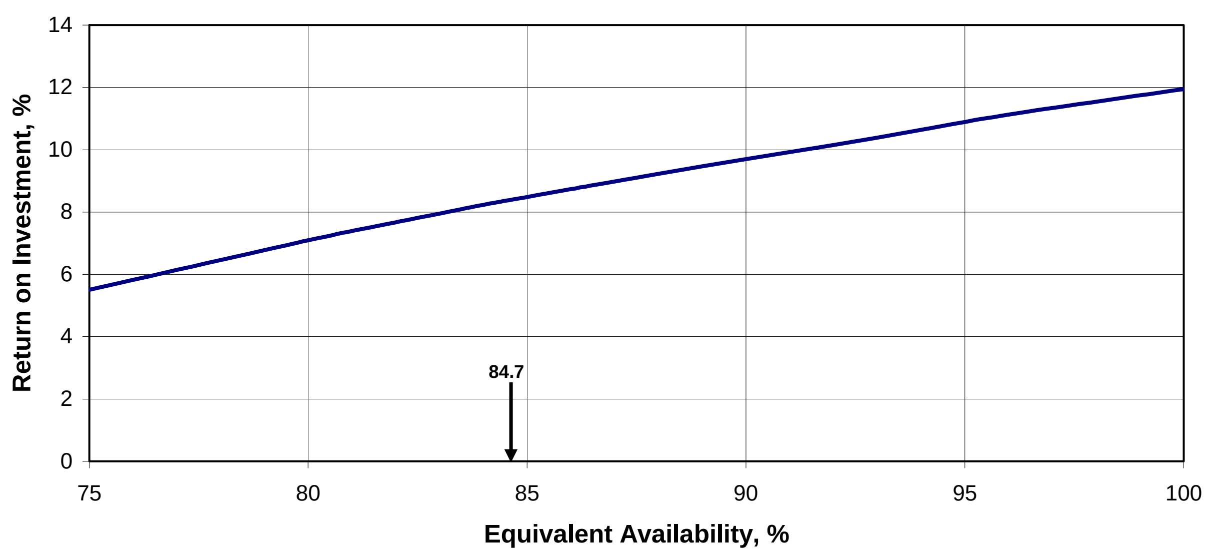


Figure 1.3 shows the relationship that varying the guaranteed availability has on the ROI. At the projected availability of 84.7%, the alternate case has an ROI of 8.4%.

Figure 1.3 Effect of Availability on Return on Investment



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 the overall plant economics is similar to that of Subtask 3.2. 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.

As with Subtask 3.2, 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, will not have the impact of these two items.

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 for 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 gasification. While the inputs 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, the above sensitivity analysis can provide insight into the outcome for plants with somewhat different base assumptions.

This study has shown that:

- Improvements that were made to this Subtask 3.3 Alternate Case increased the return on investment by 3 percentage points (5.9 to 8.4%) over that of the Subtask 3.2 Base Case.

- Commercially available processes (e.g., LO-CAT®) and technologies (e.g., Stamet solids pump) are being developed for the design of a coal fueled IGCC power plant based on the U-GAS® gasification technology that should provide reliable, long-term operation.

- A ROI of 8.4% is achievable at the current market price of electricity in upstate New York. Future plant designs may be able to identify several additional enhancements that could further improve the economics of IGCC power plants (see below for a list of other potential enhancements and improvements that were outside the scope of this study).

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

- 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.
    - Improved sulfur removal methods including warm sulfur removal
    - Warm mercury removal systems
    - Improved particulate removal systems resulting in reduced capital costs and higher efficiency
- As a result of this study, a list of R&D needs have been identified including:
    - Studying improved coal drying techniques
    - Investigating the effect that the coal moisture content has on the U-GAS® gasifier operation
    - Updating the database for gasification reactivity of the desired coal
    - Characterizing the particulate properties
    - Characterizing the hydrocarbon content of the syngas to confirm the sour water stripper design and effluent water treatment facilities
    - Investigating cyclone performance at high temperatures (greater than about 1000°F)
    - Determining the combustion turbine performance capabilities for the desired engine(s) (both output and emissions)
    - Determining the characteristics of the ash 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 (re. Section 5.4).

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 and/or R&D efforts:

- Additional optimization work is required for coals including further optimization of the plant configuration such as heat integration and sulfur recovery. One example is integration of the gas turbine and oxygen plant (although not considered herein), which could reduce compression costs. This change may significantly reduce the cost and improve the efficiency of the gasification plant. A commercial demonstration of this type of integration would be valuable to all gasification systems.

- 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 made more thermally efficient.

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

    - Improving the availability of the gasification system and various sub-systems
    - Demonstrating combustion turbine performance (both power output and emissions) on syngas in order to prepare for widespread commercialization of gasification

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

- The physical characteristics and properties of coal must be studied further in order to better predict gasification system performance. These include:

    - Determination of the gasification reactivity of the desired feedstock
    - Determination of the ash characteristics associated with the char
    - Characterization of the particulate properties
    - Characterization of 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 (above 1000°F).

    - 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 domestic company that is 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 used 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 U-GAS® gasifier was selected 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. The objective was to examine the application of a GTI fluidized bed gasifier at an industrial application in upstate New York using a Southeastern Ohio coal. Subtask 3.2 developed a base case design for this case. Subtask 3.4 developed a base case design for a stand-alone lignite-fueled IGCC power plant that produces about 251 MW of export power. (Subtask 3.1 covers management activities.)

This report describes the work performed on the third subtask, Subtask 3.3. Subtask 3.3 developed an alternate design for an air-blown Eastern Coal Case by considering additional ideas for improving performance and/or reducing investment and operating costs that were generated during the Value Improving Practices (VIP) sessions. The alternate design was developed by a series of trade-off studies that considered improvements which would be commercially available within 10 years (i.e., by 2015) such as:

- Economy of scale results in lower capital investment in the gasifier island.
- Alternative mercury removal technologies were examined. For Subtask 3.2 mercury was removed from the syngas by adsorption on activated carbon before the acid gas removal facilities. Newer technologies are being developed.
- Enhanced heat integration between the various components of the gasification block, syngas cleanup, gas turbine, and HRSG was improved.
- Alternative sulfur removal technologies were examined. The common method of sulfur removal from a syngas stream is hydrolysis of carbonyl sulfide (COS) to H_2S and CO_2 followed by adsorption of the H_2S in an amine solution. The H_2S is desorbed and processed in a Claus plant to make elemental sulfur, which is sold. One alternate method that was considered is a liquid redox method, which is an iron-based process. This technique generally is best for small plants. LO-CAT®, SulFerox™ and Sulfint are commercial processes that possibly could be used for this application. The CrystaSulf® and Morphysorb® technologies also were considered.

- Another alternative that was considered is whether to install a spare gasifier to increase the syngas availability. An availability analysis procedure similar to the one used in the previous Gasification Plant Cost and Optimization Study was used do this.[1] This required a financial analysis using a discounted cash flow financial analysis program.[2]

- The Stamet Posimetric solids pump was examined as an alternative coal feeding system. A screw feeder was selected for Subtask 3.2, and for Subtask 3.3, a trade-off study between the two coal feeding systems was performed.

Figure 2.1 is a simplified block flow diagram of the facility. The ISBL plant contains an air compression unit for the GTI U-GAS® gasifier. The syngas leaving the gasifier is cooled in the high temperature heat recovery (HTHR) area. After any remaining particulates are removed by metallic candle filters, the syngas then goes to the low temperature heat recovery (LTHR) area. The cooled syngas cleanup facilities consist of a water scrubber, mercury removal system, and sulfur recovery. The cleaned syngas then goes to the power block which consists of two GE 10 combustion turbines, each with a dedicated heat recovery steam generator (HRSG).

This design is based on the premise of providing combined heat and power (CHP) to an existing industrial or large commercial facility. The plant serves as a supplement or replacement to existing utility systems at the facility and is not intended to be a stand-alone plant design. However, it is complete from the coal grinding through the heat recovery steam generator. Since it is part of an existing complex, the financial analysis assumes that:

- Coal receiving and long term storage facilities are available.

- Boiler feed water is available for a reasonable cost from the industrial complex.

- Wastewater treating facilities also are available for a reasonable fee.

- Nominal 400 psig/550°F pressure steam from the HRSG is transferred to the existing facility for a reasonable price.

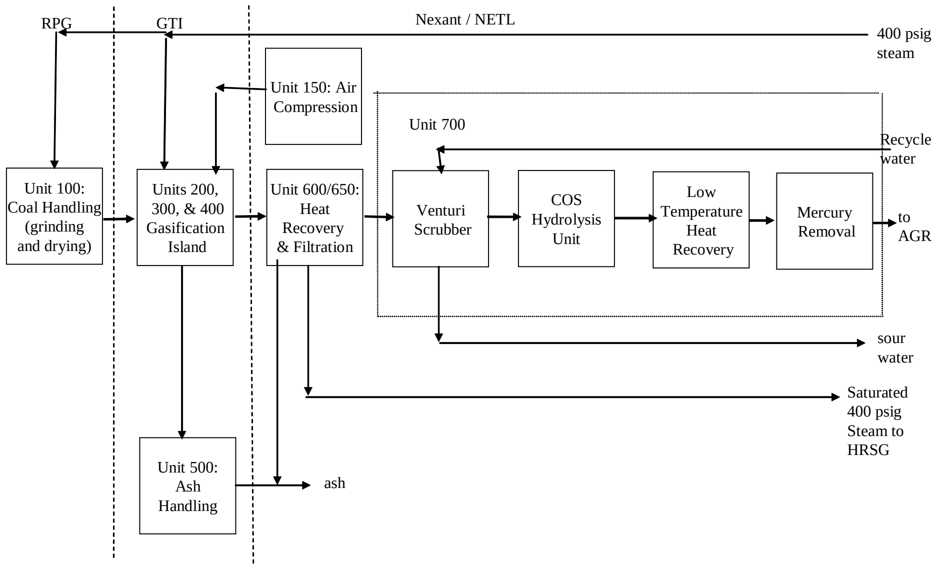
- Import steam for startup of the gasifier and other equipment is available from the other boilers at the industrial complex.

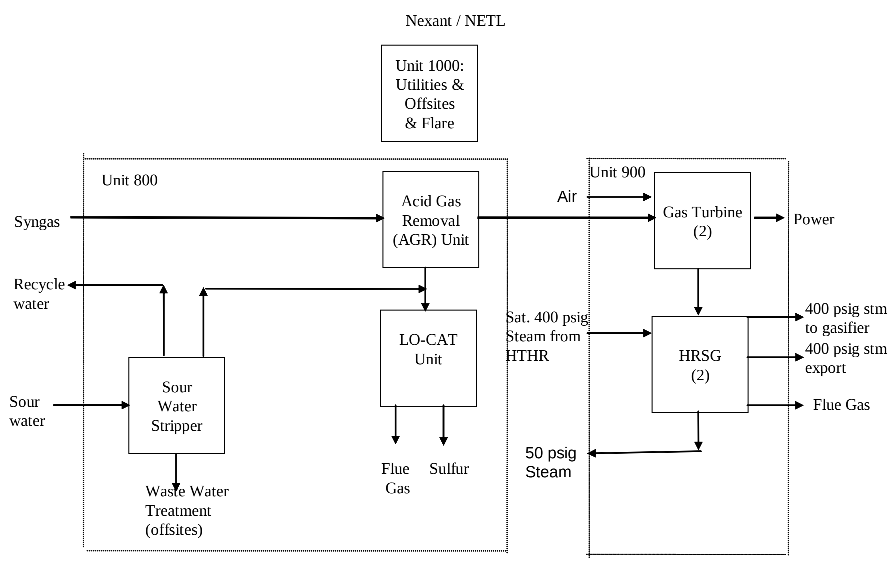
The determination of the exact coal-processing rate was part of this study. This rate was chosen so as to fully load two GE 10 gas turbines, and it is a function of the coal that is processed and the system design.

1 DOE Contract DE-AC26-99FT40342, "Gasification Plant Cost and Performance Optimization" 1999-2003.

2 Nexant, Inc., "Financial Model User's Guide – IGCC Economic and Capital Budgeting Evaluation," Report for the U. S. Department of Energy, Contract DE-AMO1-98FE64778, May 2000.

Figure 2.1 Simplified Block Flow Diagram





3.1 STUDY OBJECTIVES

The primary objective of this study was 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. The Subtask 3.2 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 Subtask 3.4 topical report examined an oxygen-blown GTI fluidized bed U-GAS® gasifier coupled with a GE 7FB (or similar sized) gas turbine (CT), heat recovery steam generator (HRSG), and a steam turbine (ST) to produce power in a stand-alone power plant. The plant was fueled by North Dakota lignite and will be located at a generic North Dakota site.

This Subtask 3.3 topical report examines an alternate design for the air-blown case developed for Subtask 3.2. The objective of Subtask 3.2 was to develop a conceptual base case design that would provide reliable, long-term operation. This was because this concept for an industrial-scale facility had not been demonstrated in a sustained long-term performance at the commercial level. Thus, the base case design was developed at the expense of thermal efficiency with limited heat integration to promote operability and reliability on the premise that the utility systems at an industrial plant generate no revenue. The objective of this current study is to improve the financial performance of the facility by reducing the capital investment and operating costs of the gasification facility.

One of the objectives of this task is to enhance NETL's capabilities to perform system analysis. In order to accomplish this 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, to 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 approach commensurate to the level of results needed

- Used spreadsheets, ASPEN, GateCycle, 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 improved financial performance
- 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, the E-GAS™ gasification technology was purchased by ConocoPhillips. 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 was added to the project 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 upstate New York that co-produces both

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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

power and steam.[6] Both air and oxygen-blown gasification systems were considered. Subtask 3.3 developed an alternate plant design based on the air-blown plant design from Subtask 3.2. This case was chosen over the oxygen-blown case due to superior economics and performance. Subtask 3.4 developed a design of a nominal 251 MW power plant fueled by North Dakota lignite.[7]

This document is the Topical Report for Subtask 3.3, an Alternate Design for the Eastern Coal Case.

3.3 METHODOLOGY

Task 3, *Gasification Alternatives for Industrial Applications*, shifts the focus of the study in Tasks 1 and 2 from large plants to smaller ones in Task 3. The objective of Subtask 3.2 focused on 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. Subtask 3.2 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 3.3 provides an alternate improved design of the base case developed for Subtask 3.2.

The U-GAS® gasification technology system developed by the Gas Technology Institute was the basis for this project 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 has been studied to compile relevant information for this project. A history of the U-GAS® process is provided in Section 3.2.3 of the Subtask 3.2 Topical Report.[6]

Figure 3.1 is a schematic diagram of the steps involved in developing the design, cost and economics for a specific case. More information can be found in Addendum E (the design basis work plans). Using Subtask 3.2 as the basis, a series of trade-off studies together with the results from a Value Improving Practices (VIP) Program were analyzed and the results incorporated into the design. Based on this alternate design case, process simulations were developed for the syngas cooling, syngas cleanup, and sour water stripper portions of the plant. The resulting heat and material balances provided the input to the GateCycle simulation program for a detailed simulation of the power block. This report and the addendums contain sufficient information for verification of the carbon, slag, sulfur, and heat balances.

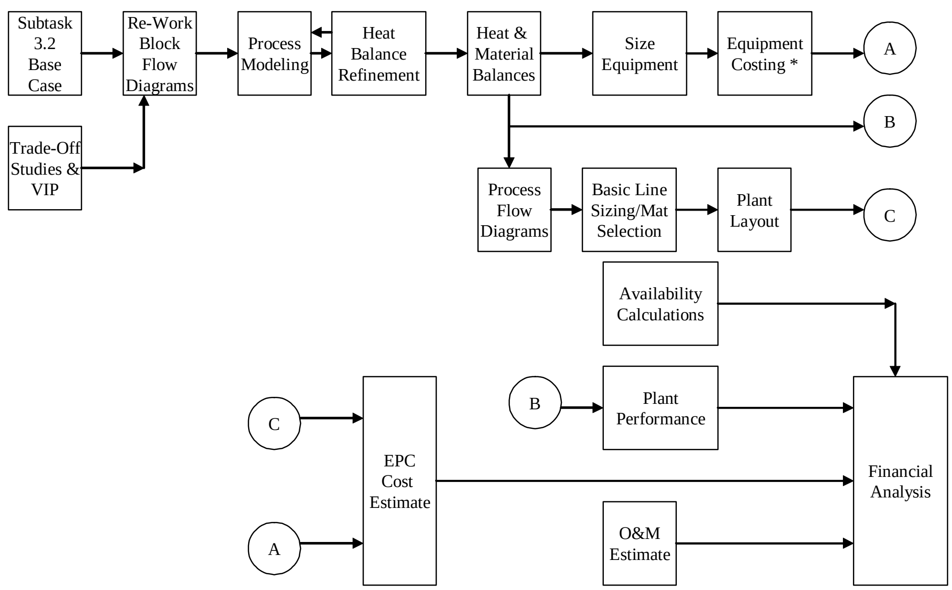
6 "Topical Report – Task 3: Gasification Alternatives for Industrial Applications, Subtask 3.2: Preliminary Design for Eastern Coal Case," Gasification Plant Cost and Performance Optimization, United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, November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

7 "Topical Report – Task 3: Gasification Alternatives for Industrial Applications, Subtask 3.4 – Lignite-Fueled IGCC Power Plant," Gasification Plant Cost and Performance Optimization, United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, March 2005.

Based on the model results, PFDs, sized equipment lists, line sizing, and other information necessary to calculate the plant cost were developed. The mid-year 2004 plant cost was built-up based on cost information from selected equipment quotes, information from similar Bechtel projects, and from commercially available cost estimating software.

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.

3.4 AVAILABILITY ANALYSIS

The common measures of financial performance, such as return on investment, net present value, 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 annual 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.[8] 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.[9] Details of the GTI's availability estimate for the gasification island and of the quantitative estimates are provided in Addendum F of the Subtask 3.2 Topical Report.[10]

Recent data presented at the 2002 Gasification Technologies Council conference show further reliability improvements in the on-stream performance of the Wabash River Repowering Project.[11] 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 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

8 "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.

9 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.

10 "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.

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

scheduled maintenance outages to determine the annual feed and product rates, and from this, the annual revenues.

This study also assumes a mature facility, as compared to a first-of-a-kind 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(s) to maximize power production. Natural gas is preferred for these applications. When this situation occurs, the power output from the turbines is reduced. Furthermore, the internal power consumption also is reduced by that of the non-operating units. 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, power demand, etc. However, this study does not consider the use of natural gas to fuel the gas turbines when syngas is not available.

In addition, all plants use some natural gas 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 used in the gasification model came from information provided by US government agencies. This includes data from the DOE’s Energy Information Administration (EIA) Annual Energy Outlook 2004[12] for commercial electricity values, natural gas, and coal, and from the US Geological Survey (USGS) for sulfur. The steam value was calculated using natural gas as the marginal fuel for production, while the gasifier bottoms value was estimated using previous values for Nexant gasification studies. Each of the values was normalized where necessary to reflect the current nominal value, using a 3% inflation rate. The preliminary model runs were made using these inputs. Table 3.1 below lists the major assumptions for the 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.

[12] 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

	Price	Escalation (%/yr)
Feeds		
Coal	27.20 $/short ton	2.0
Natural Gas, HHV	4.68 $/MMBtu	4.0
Products		
Electric Power	8.0 cents/kWh	3.0
Steam	12.00 $/short ton	3.0
Sulfur	26.52 $/short ton	3.0
Gasifier Bottoms	10.00 $/short ton	3.0

The assumptions made for Subtask 3.2 were reviewed by the project team, including representatives from the industrial facility, for accuracy. Modifications were made to both the electricity and steam values to better reflect expected costs for 2005 that will be incurred by the industrial facility. The electricity value is based upon the marginal supplier of electricity to the industrial facility. The rate schedule enrolled in by the industrial facility is the cost basis. Sulfur, gasifier bottoms, natural gas, and coal values were all left unchanged from EIA and USGS estimates to adequately reflect a "typical" industrial facility in this part of the country.

For the most part, EIA factors also were used to predict price escalation during the life of the project. These factors basically are consistent with the values that Nexant has used on previous gasification studies. 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%, was used for the electricity price. EIA predictions for natural gas follow a similar trend, with a slight decrease, followed by price increases after 2011. This increase, however, is expected to lead to natural gas slightly outpacing the rate of inflation during the life of the project. Therefore, natural gas escalation was set at 4%. Since natural gas is not a main plant feed, the small amount of natural gas that is used is included in the variable O&M costs, making this input insignificant. Natural gas costs will be relevant only when co-firing will be used or in comparison with other power producing alternatives.

In keeping with previous Nexant studies and expectations of oversupply in the coal industry, the coal escalation rate was kept to 2%, below what is expected for future general inflation rates. This is between current EIA estimates and escalation factors used in previous Nexant studies. While there may be 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. These escalation rates were maintained throughout the life of the gasification facility.

The gasifier bottoms product 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, including negative values.

3.6 FINANCIAL ANALYSIS METHODOLOGY

The results reported for rate of return and discounted cash flow come from the Nexant developed IGCC Financial Model Version 3.01. This version of the model was developed in May 2002 specifically for NETL under a task order from NETL on-site support contractor E2S. 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 results.

In order to develop the appropriate financial assumptions for the industrial facility under consideration, a number of sources were reviewed and conversations held with team experts. 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 US refining market, and previous gasification optimization studies performed by Nexant, namely the "Gasification Plant Cost and Performance Optimization" study (DOE Contract number DE-AC26-99FT40342) for NETL. Details of the financial assumptions can be found in Addendum C of the Subtask 3.2 Topical Report.[6] A few of the major assumptions and some of the areas that will be explored via a sensitivity analysis are listed below:

- Start-up in 2015
- + 30/-15% accuracy 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 downtimes for 21 days of the calendar year based off gasifier requirements. This is coupled with the availability analysis to calculate the operational time per year.
- 8% cost of capital
- Total operation and maintenance (O&M) costs of 5% per year (fixed and variable)
- 32-month construction period
- 20-year plant life
- Fees added to 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 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, steam production, sulfur produced, and quantity of gasifier bottoms. 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 the Plant Cost section of this report (Section 6.2). Appropriate scale-up factors used in previous gasification projects allowed any equipment not reflecting installed cost to be properly estimated.

4.1 STUDY BASIS

This study investigated the cost for installation and operation of a combined heat and power (CHP) facility at an industrial site. The goal of the study was to identify alternatives for improving the return on investment and lowering plant emissions associated with power generation as developed for Subtask 3.2.[1] The location for this facility is the same as for Subtask 3.2 (i.e., at an industrial site in upstate New York).

The design criteria for Subtask 3.3 are the same as for Subtask 3.2 and are repeated here for convenience:

- Two GE 10 gas turbines @ ~12.5 MW each (total = 25 MW)
- Maximize co-generation of steam from the gasifier and HRSG
- Export 400 psig/550°F steam to industrial site
- Southeast Ohio coal (assume 15% moisture for design, 8.4% moisture normal), as defined in Table 4.1

Table 4.1 Southeast Ohio Coal Analysis

Ultimate Analysis, wt%	**Dry Basis**	**As Received**	**ASTM Method**	**Ash Fusion Temperature**	**°F**
C	74.65	68.38	D3176	IT	1974
H	5.79	5.3	D3176	ST	2025
N	1.54	1.41	D3176	HT	2049
S	3.32	3.04	D4239	FT	2067
Ash	5.91	5.41	D3176		
O	8.79	8.06	D3176	**Coal Ash Analysis**	**wt%**
Total	100.0	91.6		SiO_2	33.3
				Al_2O_3	29.6
Proximate Analysis, wt%				Fe_2O_3	29.3
Residual Moisture			D3173	TiO_2	0.6
Total Moisture		8.4	D3302	CaO	2.9
Ash	5.91	5.41	D3174	MgO	0.7
Volatile Matter	43.24	39.6	D3175	Na_2O	0.4
Fixed Carbon	50.85	46.59	D3172	K_2O	0.5
Total	100	100		SO_3	2.1
Air-Dry Loss		5.53	D2013	P_2O_5	< 0.1
Sulfur	3.32	3.04	D4239	BaO	< 0.1
Gross Caloric Value, Btu/lb	13,590	12,448	D1898	Mn_2O_3	< 0.1
Dry, Ash Free, Btu/lb	14,443			SrO	< 0.1
Pounds SO_2/MMBtu	4.88			Total	99.4

[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.

Environmental performance, based on the DOE target emission and performance goals, is the same as for Subtask 3.2. and repeated here for convenience:

- S > 99% removal
- NOx < 0.05 lb/MMBtu
- Particulates < 0.005 lb/MMBtu
- Mercury > 90% removal
- Thermal Efficiency = 45-50%
- Capacity factor = 85%

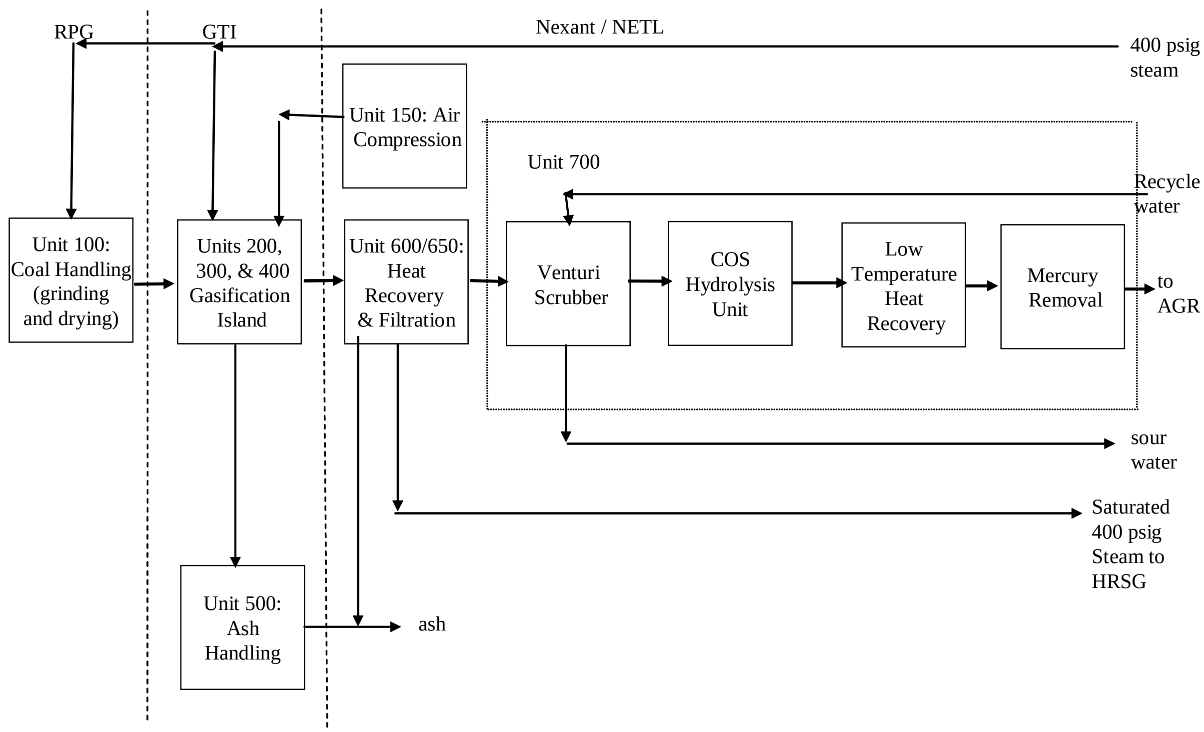
4.1.1 Plant Description

The U-GAS® plant at the industrial site consists of the following process blocks and subsystems:

- Unit 100: Coal Prep/Handling
- Unit 150: Air Compression
- 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, and Sour Water Stripper (SWS)
- Unit 900: Power Block including two GE 10 combustion turbines (CT) and two heat recovery steam generators (HSRGs)
- 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 in two parts. The first part shows the syngas generation and processing areas, with the second part showing the sulfur removal, sulfur recovery, sour water stripper, and power block.

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

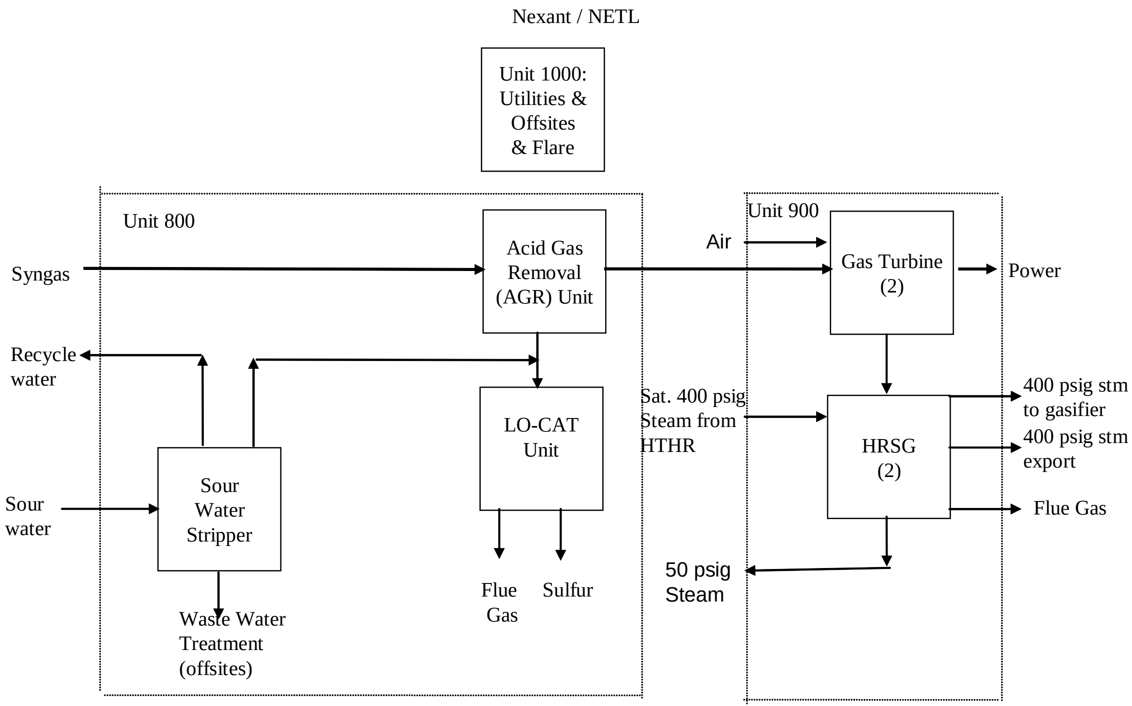


4.1.2 Site Selection

The upstate New York industrial facility is a large site of over 1,800 acres. There are 5 locations that have been identified where this gasification facility could be located. Critical site issues include:

- Sufficient open space for all equipment
- Distance for power interconnections
- Ability to balance steam from the IGCC into the existing industrial site infrastructure
- Access for coal storage and handling

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



4.1.3 Feed Stock – Eastern Bituminous Coal

A Southeastern Ohio coal was chosen as the design fuel to represent the Eastern bituminous coal fields. Southeastern Ohio coal was selected because there are significant quantities mined, and there is excellent transportation from this region via barge, rail, and truck to many of the industrial facilities in the Eastern industrial belt of the United States. This coal is typically higher in sulfur than coal from other areas mined in the Appalachian coalfields, and thus, is discounted compared to those fuels. It is anticipated that coal from this area could be delivered to industrial facilities at a cost of about 1.00 to 1.50 \$/MMBtu. This fuel has not been specifically tested by GTI in its pilot plant facilities, but is similar to Pittsburgh and Kentucky seam coals that have been extensively tested. The properties of this Southeastern Ohio coal are given in Table 4.1.

Coal properties are important parameters for determining if the fuel is a suitable candidate for reactor operation in the agglomerating mode. Ash agglomeration produces a hard glassy ash product that is very low in carbon and facilitates discharge from the gasifier. This Southeastern Ohio coal is likely to possess these qualities based on its ash deformation temperature and chemistry. Chlorine concentration in this coal seam averages about 460 ppmw, and mercury averages 0.12 ppmw.

4.2 PROJECT OVERVIEW

This Subtask 3.3 Topical Report is the third in a series of studies of preferred designs for upgrading the industrial IGCC power and steam facility. The first study (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. The Subtask 3.4 study considered a larger, grass-roots, stand-alone power plant consisting of an oxygen-blown gasification train producing sufficient syngas to fully load a GE 7FB combustion turbine and fueled by North Dakota lignite. This Subtask 3.3 study examined a variety of alternatives for improving plant costs including new sulfur removal technologies.

The objective of Subtask 3.2 was to design an industrial-size, IGCC coal-fired gasification system focusing on plant operability. Subtask 3.3 looked to improve this design by investigating options for cost reduction that would not negatively impact plant performance. 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 the continued use of premium fuels or 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 of coal allows for energy stability and security at the facility. Use of 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.

Industrial facilities in the United States are facing stricter environmental regulation in the next few years. In the past, many industrial and large commercial boiler facilities have switched to fuel oil or natural gas to avoid the expense of installing post combustion emission controls. However, during the past few years, the increasing price volatility and expense of using these fuels has placed a financial burden on US industry. Using coal as the fuel source at an industrial site gives the owner the knowledge that he will have low relatively stable fuel costs. Furthermore, there are abundant coal resources (over 240 years supply at current usage rates) in the United States compared to more limited amounts of oil and natural gas. As environmental rules tighten, industry will be forced to choose between expenditures: 1) for emission controls on coal boilers; 2) fuel switching to more costly premium fuels; or 3) shutdown of non-competitive facilities.

IGCC plants can provide industry with a viable alternative. Gasification offers several advantages as a long-term solution. These relate to lower cost, lower emissions, increased efficiency, and improved reliability.

First, coal is an abundant, low-priced energy source that is expected to have a stable low price over the foreseeable future. IGCC systems have higher thermal efficiencies than steam boilers, which reduce the fuel costs by reducing the amount of coal that is consumed to produce a given amount of power. Industrial facilities that purchase electricity and natural gas from power and energy suppliers must pay “retail” rates for their energy use. Self generation of electricity by an industrial site will often be lower in cost than what can be purchased from the grid. Similarly, coal transportation is not subjected to the transportation costs associated with purchase of gas from local suppliers or pipelines. Finally, self generation of power and steam avoids payment of state and local taxes typically added to retail energy purchases.

Second, pollution reduction also is simplified. Sulfur removal is easier because the sulfur is removed from the syngas stream where it is more concentrated than in the flue gas. NOx reduction is accomplished by the use of low NOx combustors in the gas turbine. If syngas is used as a fuel other than in the turbine, low NOx burners can be used to reduce NOx emissions. Mercury and heavy metal removal from syngas has been demonstrated at Eastman Chemical by adsorption on sulfur-impregnated carbon. Lastly, the higher efficiencies associated with IGCC reduces the quantity of carbon dioxide that is generated compared to burning coal in conventional boilers.

Increased efficiency has two major benefits. As the efficiency is improved, less fuel is required to produce the same amount of product resulting in lower fuel costs. Furthermore, since less fuel is used, the front end of the facility (coal handling and preparation, coal feeding, ash handling, etc.) becomes smaller and less costly. Finally, since less fuel is burned, less CO_2 and other pollutants are generated and released into the atmosphere.

Lastly, on-site reliability of the energy supply is enhanced. This is accomplished via several means. First coal can be conveniently stored to avoid fuel supply disruptions. It can be transported by truck, rail, or barge. Self generation of electricity can protect a facility from supply disruptions such as the power failure that covered the Northeastern U.S. and Canada in August, 2003. Similarly, during very cold weather, natural gas is sometimes curtailed to large industrial customers to ensure an adequate supply to residential consumers.

The basis for this study is to develop a CHP facility producing nominally 25 MW. The study uses GTI’s fluidized bed U-GAS® gasifier coupled with GE 10 gas turbines and heat recovery steam generators (HRSG) to co-produce power and superheated high pressure steam (400 psig/550°F) at a typical industrial complex. The steam can be used at various locations throughout the complex.

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 is not mature. In Subtask 3.2, the philosophy that was used for the plant design of was to maximize availability by keeping the design as simple as possible. For Subtask 3.3 the heat integration was improved and the resulting plant is more highly integrated. While this creates greater plant complexity, the resulting design reduces plant costs without sacrificing performance or availability.

For Subtask 3.2 the syngas cooling section of the plant was designed to minimize deposition and erosion problems as a result of dust carried in the syngas. Therefore, only one heat exchanger was placed before the water scrubber. In the Subtask 3.3 design, the high temperature syngas cooler cools the syngas to about 480°F where candle filters are used to remove the particulates. This allows for further heat integration before the water scrubber and the use of a different type of scrubber (see next paragraph).

In Subtask 3.2 the water scrubber had two functions: scrubbing dust, light oils, HCl , and other contaminants out of the syngas, while simultaneously cooling the syngas from 600°F to 265°F. Because of the absence of essentially all of the particulate matter, a venturi scrubber is used in Subtask 3.3 to reduce the water requirement. The syngas is then processed to remove contaminants such as mercury and sulfur.

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 $/MMBtu, 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[2] 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 natural gas are for prices to decrease (2002 dollars) to 4.16 $/MMBtu by 2010 and then slowly increase to 5.10 $/MMBtu in 2025.[3] This represents a 4% 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 over 9.00 $/MMBtu by 2025. Natural gas prices have demonstrated significant volatility over the past few years. Price volatility is not expected to dampen considerably since price variations are expected to continue responding to 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 by bypassing the LDC, the price is about 0.90 $/MMBtu 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 $/MMBtu 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. Delivered coal prices to industrial users are typically between 1.25 to 2.00 $/MMBtu (highly dependent on fuel type and delivery cost). Furthermore, coal prices are projected by the EIA to remain flat over the next 20 years. This nets a fuel cost differential in favor of coal of roughly 3.00 to 5.00 $/MMBtu, 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 4.50 $/MMBtu. 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

2 Malita Marie Garze, Chicago Tribune, Energy Costs an Offshore Factor, 4/25/2004.
3 Annual Energy Outlook 2004 with Projections to 2025, www.eia.doe.gov/oiaf/aco/economic.html

basis, there are clearly many facilities that can justify a serious evaluation of this technology as long term solution to meeting its energy and environmental 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 MMBtu/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. Costs associated with loss of manufacturing and industrial lost productivity have been studied by EPRI and others. These studies reflect the importance of an uninterrupted supply of electric power and steam to an industrial facility. Often, a brief outage of only a few minutes 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 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 two parallel GE 10 (or similar sized) gas turbines and HRSGs with a total electrical output of nominally 25 MW. This output size was selected for several reasons:

- This size fits well within the existing Industrial Partner’s facility
- This size can fit well within numerous industrial facilities nationwide
- There are many gas turbine vendors that may be able to supply engines ranging in size from 10 to 30 MW that can readily benefit from this cost study
- Industry practice is to use multiple utility systems to ensure high availability
- Multiples of this size equipment can be readily developed to provide facilities of a larger scale
- Facilities of this scale could be developed in a modular structure to allow a significant amount of shop fabrication for more cost effective construction

Syngas to power the gas turbines is supplied by a single gasifier using GTI’s U-GAS® fluidized bed gasifier technology. For the purposes of the study, the GE 10 engine was selected for the gas turbine. Each turbine requires 140.5 MMBtu/hr to produce 14.74 MW (gross). Waste heat from the engines and the gasification system is used to produce about 120 Mlb/hr of steam; a portion of which is used internally and the rest can be used for additional power generation, heating, and/or cooling in the industrial facility.

The gasification system contains several subsystems:

- Coal Handling and Preparation

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

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

A project construction schedule is shown in Addendum D.

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 are 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, modified IGCC applications (i.e., combined heat and power, CHP) at an industrial site. 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 industrial-scale electricity and steam generation using modified IGCC or CHP concepts. Additional cost savings have been identified for study, but not yet quantified.

Place (location) – The northeast location seems to be the best location for an eastern coal evaluation because there are many industrial facilities in this region of the country that were originally constructed to use coal for their steam and on-site electric power generation. These facilities will be required to retrofit emission control equipment to convert these facilities to less polluting premium fuels as new emission standards are enacted. The past use of coal and availability of existing coal related infrastructure makes implementation of gasification related technologies for replacement of old power systems more cost effective in the near term.

Product (or Market Penetration) – The initial application of a small industrial CHP will further develop the technology leading to improved designs; reduced costs; and increased efficiencies.

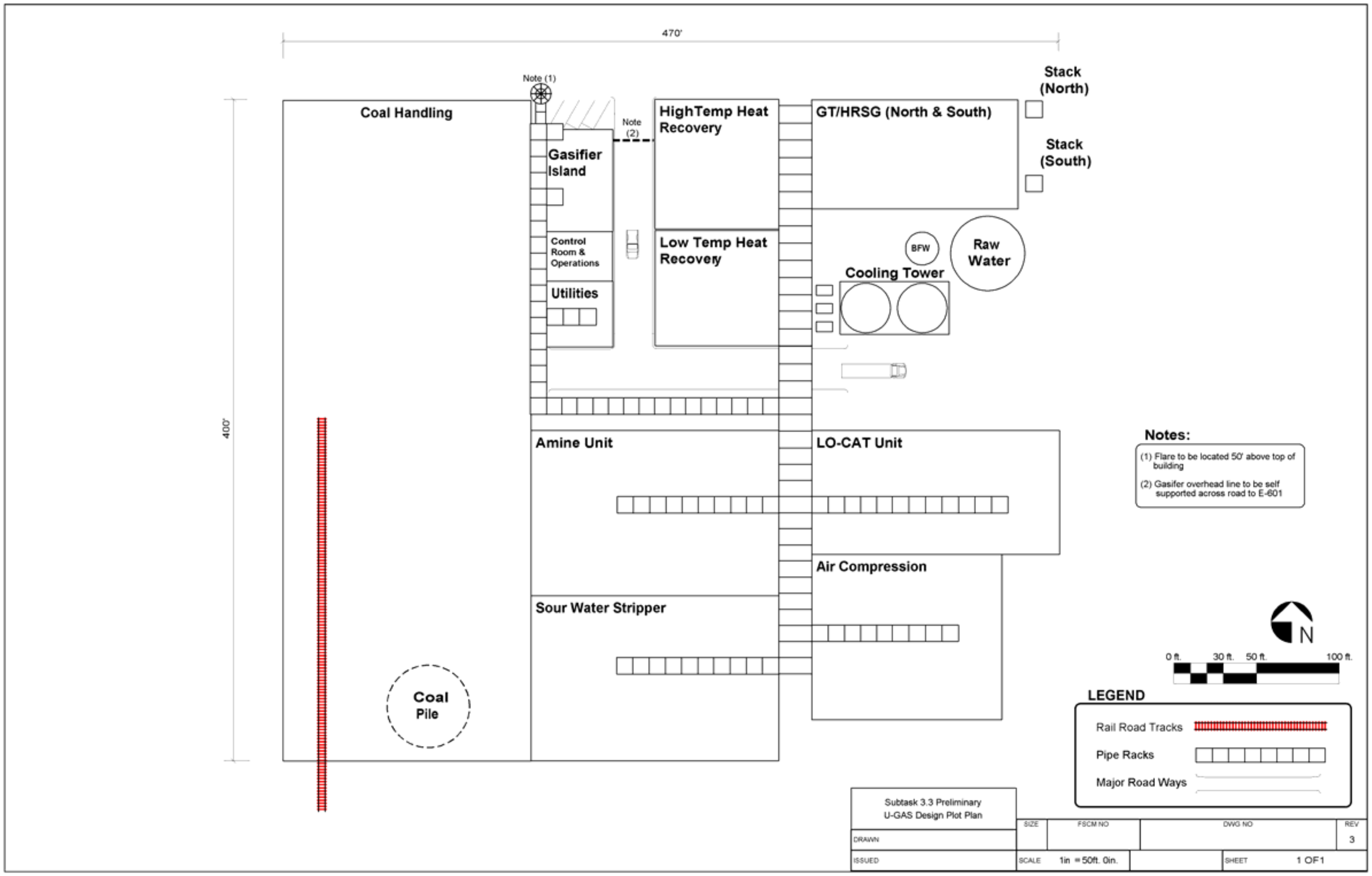
Proliferation – As more IGCC plants are built, their costs will decrease, availability will improve, and companies will be more willing to proceed with the construction of additional IGCC plants.

Promise – IGCC plants have higher efficiencies than pulverized coal 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 industrial applications of IGCC by demonstrating that it is possible to build a low cost 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



5.1 INTRODUCTION

The study here presents an alternate case for an air-blown gasifier in an industrial application sited in upstate New York. The alternate design employs improvements that either increase the overall efficiency or decrease the investment or both. These include:

- A single train gasifier island
- The use of the Stamet "solids" pump in place of the feed lockhopper system
- A combined bottom and fly ash handling system
- Candle filters for the removal of solid particles
- A venturi scrubber in place of the impingement scrubber
- Improved heat integration
- Simplified sour water stripper
- LO-CAT® system for sulfur removal

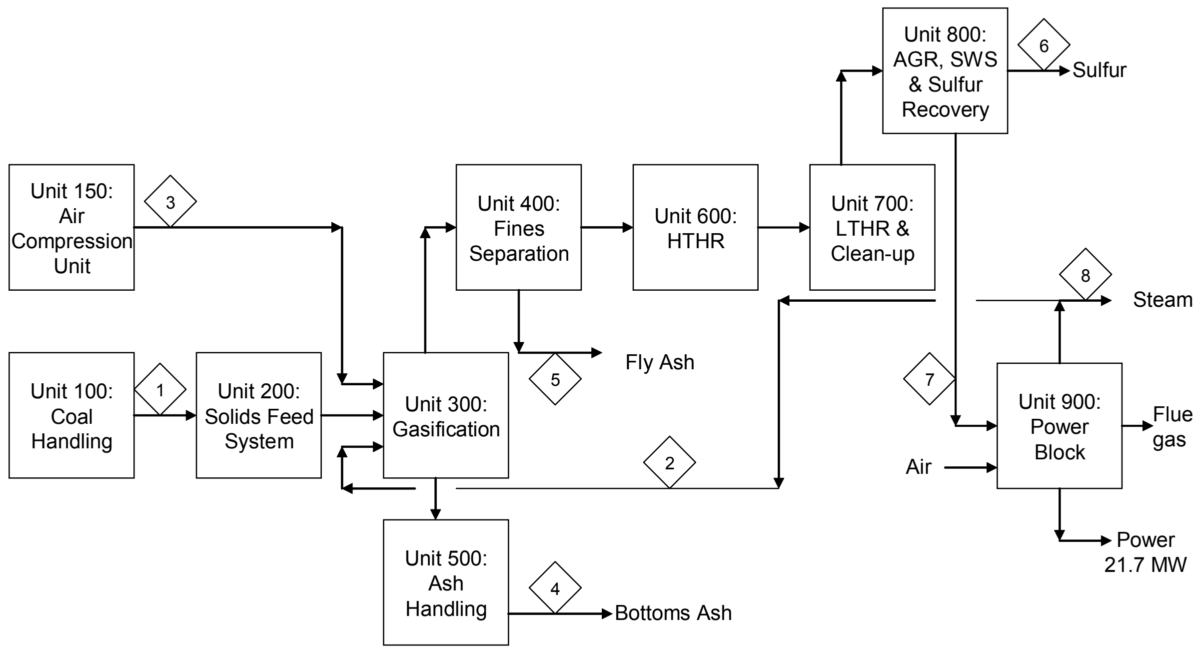
For convenience of the reader the base case flow scheme for the air-blown design (Subtask 3.2) is summarized in Section 5.2. The changes made for the alternate design (Subtask 3.3) are described in Section 5.3.

5.2 SUBTASK 3.2 AIR-BLOWN CASE DESIGN (BASE CASE)

The Eastern Coal base case is presented in detail in the Subtask 3.2 Topical Report[1]. The overall material balance generated using ASPEN is shown in Figure 5.1.

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

Figure 5.1 Simplified Block Flow Diagram and Material Balance for the Base Case



Stream Number	1	2	3	4	5	6	7	8
Description	**Coal to Gasifier**	**Steam to Gasifier**	**Oxidant to Gasifier**	**Bottom Ash from Gasifier**	**Fly Ash from Cyclones**	**Sulfur Product**	**Clean Syngas to Gas Turbine**	**Export Steam Production**
Pressure, psig	14.7	420	415	34.7	34.7	NA	295	400
Temperature, F	70	550	500	1850	1850	NA	120	550
Flow Rate, lb/hr	30,250	9,653	96,106	1,465	632	899	123,098	101,700

5.2.1 Coal Handling (Unit 100)

The coal handling system is designed to receive and unload coal from unit-train rail shipments delivered to the plant once per week. Rail cars are separated by the plant rail car handling system and delivered to the unloading area where the cars are dumped and unloaded. This system can handle about 300 tons/h of fuel which is transferred to a ready pile. Coal from the ready pile is delivered via a reclaim hopper, vibrating feeder and conveyor to a crusher/dryer that prepares the as-received coal for feeding to the gasifier. The gasifier feed specifications are:

- No more than 4% > ¼ - inch
- No more than 10% < 100 mesh
- No more than 5% surface moisture

After treatment, the coal is delivered to a silo that contains a one day inventory of prepared fuel. The coal is transferred from the silo to either of two bucket elevators for delivery to the Gasifier Island. A second silo, that is used to store startup coke, is located next to the main fuel silo.

5.2.2 Air Compressor (Unit 150)

Two parallel Ingersoll-Rand air compressors, each with 50% capacity, compress the air to the gasifier inlet pressure, 415 psia. Two rotary compressors provide operating flexibility during operations at reduced capacity. Each compressor requires a 4,000 BHP electric motor drive, and has five intercooler stages. The air is discharged from the compressors at 224°F and is heated to 500°F with superheated steam before entering the gasifier.

5.2.3 Gasification Island (Units 200, 300, 400 & 500)

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

The gasifier is lined with refractory to minimize heat losses. The outer layer is designed to minimize heat loss, and the inner layer is made of abrasion resistant material to withstand the rigorous environment of the gasifier. The gasifier bed is supported by a grid. Oxidant (air) and steam enter to the gasifier below the grid. Fuel is fed to the gasifier just above the grid. Solids that are recycled from the dust cyclones by gravity through the dip legs also are returned to the gasifier bed at a level 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 45½ feet tall which is of sufficient height for the grid, bed, and disengaging zones. The syngas exits the gasifier (when operated on bituminous coal) at approximately 1850°F. The gasifier operates at 340 psig to provide adequate available pressure throughout the plant ahead of the gas turbines.

The dust (or fly ash) removal system consists of a series of equipment to cool the dust and to transport it via a lockhopper system from the gasifier pressure to storage at atmospheric pressure. A pressurized cooling screw cools the dust from the high temperatures of the gasifier to a temperature of about 500°F to protect the lockhopper valves and to allow the use of carbon steel equipment downstream. A refractory lined surge hopper collects dust when the screw is not rotating (lockhopper closed). When the lockhopper is full, the upper valve is closed, the vessel pressure is lowered to atmospheric, 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 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.

The bottoms ash removal system works in the same way.

5.2.4 High Temperature Heat Recovery (Unit 600)

The high temperature heat recovery system comprises a fired-tube, steam boiler and a steam drum. The steam boiler is a vertically oriented 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 raw syngas goes to the syngas cleanup unit after exiting the steam boiler at 600°F. A thermosyphon loop is employed between the steam boiler and the steam drum. Boiler feed water enters the steam drum at 250°F and 435 psig from the 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 saturated steam at 425 psig is routed to the heat recovery steam generator (HRSG) to produce superheated 400 psig steam at 550°F.

5.2.5 Low Temperature Heat Recovery & Clean-up (Unit 700)

The syngas cleanup system consists of two syngas scrubbers, a COS hydrolysis unit, and a low temperature heat recovery unit. The syngas streams from the two trains merge before they enter the preheater for the COS hydrolysis reactor.

An impingement column was selected for the syngas scrubber. 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 particulate laden sour water leaves the bottom of column and goes to the SWS. A combination of three different water sources is used in the scrubbers. They are (1) clean process water, (2) process condensate, and (3) recycled water from the SWS. By using the process condensate and recycled water from SWS, the amount of fresh make-up water is minimized. Half of the process condensate is recycled to the wash column, while the other half is routed to the SWS for further treatment. This is done to prevent buildup of contaminants in the system. The cleaned syngas leaving the wash column is saturated with water and contains twice the amount of water than it had when entering the scrubber. All remaining particulates, light oils, most of the chlorides, and a part of the ammonia are removed from the syngas in the wash columns.

Most of the sulfur in the coal is converted to hydrogen sulfide (H_2S) during the gasification process. However, a small portion is converted to carbonyl sulfide (COS). The COS concentration downstream of the water scrubber is about 315 ppm by weight.

In a Claus unit, only H_2S is converted to elemental sulfur. Thus a system is needed to convert the COS to H_2S to achieve 99% total sulfur removal.

The Süd-Chemie group designed the COS hydrolysis unit in which COS reacts with water over a 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 water scrubber at 265°F is saturated with water. A small heater is used to raise the syngas above its dew point. The syngas enters the hydrolysis reactor at 275°F which favors the shifting of the hydrolysis reaction towards the formation of H_2S. A conventional shell-and-tube heat exchanger is used to heat the syngas with 400 psig steam.

The mercury removal catalyst bed and the amine acid gas removal system require that the incoming syngas be at about 110°F. A low temperature heat recovery system recovers a portion of the sensible heat from the syngas exiting the COS hydrolysis reactor, and further cools the syngas before it goes to the mercury removal bed and subsequently the amine system.

With most of the COS being converted to H_2S, the syngas leaves the COS hydrolysis at a temperature 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 250°F. Then the syngas is cooled to 140°F in an air fin cooler before being further cooled to 110°F with cooling water in a shell-and-tube exchanger.

5.2.6 Mercury Removal, Acid Gas Removal, Sour Water Stripper, & Sulfur Recovery (Units 700 and 800)

Mercury is removed from the syngas by adsorption on Calgon Carbon HGR® sulfur impregnated activated carbon. Over 90% of the mercury in the syngas is removed by adsorption on the activated carbon. The carbon bed is expected to have a life of over 3 years, after which it is discarded.

The syngas leaving the mercury adsorption 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 process, which selectively removes most of the H_2S, but allows most of the CO_2 to remain in the syngas stream. By allowing the CO_2 to slip through the system, the sizes of the amine regenerator and sulfur clean-up equipment can be made smaller than otherwise possible with other process designs. This reduces the capital and operating costs associated with this system. The amine based acid gas removal unit consists mainly of an absorber and a regenerator. The treated syngas then flows to the gas turbines.

The acid gas stream leaving the regenerator is converted to elemental sulfur. Based on demonstrated performance on syngas and on the required scale of production (10-11 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 the amine system and 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 very high overall sulfur removal from the syngas, on the order of 99.1% or higher. The elemental sulfur produced in the Claus reactor is sold as an additional source of revenue.

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 both high and low pressure steam. This steam along with the steam generated in the Claus reactor are used in the reboiler of the amine stripper. The vent gas containing 1.8 lb/hr of sulfur is dispersed to the atmosphere at about 550°F.

Sour water from the syngas scrubber is mixed with the process condensate in the flash drum and is flashed at 24 psig and 240°F. Most of the inlet CO_2 and approximately half of the inlet H_2S leaves the flash in the vapor stream. The liquid stream is then cooled to 186°F (approximately 10°F below the bubble point at atmospheric pressure) to reduce the chance for any off-gassing in the settling tanks.

Settling tanks were selected to remove the particulates and possibly some oils in the sour water from the syngas scrubber column and process condensate. Due to the small size of the particles, 5 to 10 μm, a flocculent is added to agglomerate the small particles and increase their settling velocities. The design basis was adopted from the successful operation at the Polk Power Station. Two settling tanks in parallel are used for reliability and to provide extra capacity for excess particulates. Specific details of the settling tanks will require particulate samples in order to optimize the settling tanks and identify the types and amounts of flocculants to be added to the particle laden stream. Additional chemical treatment also may be used. Other chemical treatments may be added to the water to agglomerate any oils.

It was assumed that the slurry of agglomerated particulates at the bottom of the tank will contain 75% water. Two pneumatic positive displacement pumps are used to transport the slurry from the bottom of the settling tank to filter presses that are used to dewater the slurry. Each settling tank will have its own filter press. The filter presses were quoted by USFilter and are designed to operate in an automated batch mode once per shift. The effluent from the presses is collected in a sump.

Water exiting the settling tank and filter press is then pumped to a day tank. This day tank has two purposes; first it dampens changes in the composition and flow rate of the sour water, and secondly, it has sufficient storage capacity to account for a one-day outage of the sour water system. The day tank typically is operated with a two-hour hold-up time. Liquid from the day tank then is 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, and the reboiler is heated with 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-finned 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 (WWT) 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 WWT is within the existing industrial site and, thus, outside the scope of this project. The industrial facility has indicated that it can handle the additional waste streams.

5.2.7 Power Block (Unit 900)

Clean syngas is sent to the combustion turbine (CT) at 120°F and 295 psia. The CTs are rated at 11.25 MW at ISO conditions fired using natural gas. The use of syngas can produce higher generator output due to a higher mass flow rate to the turbine. This phenomenon is sometimes referred to as the “syngas boost”. Modeling estimated that each CT would generate approximately 14.93 MW of net power. This is consistent with prior performance estimates provided by GE for the use of syngas in the combustion turbine.

The following describes the exhaust gas and water/steam flow for each individual HRSG train.

Flue Gas Flow – Exhaust gas exiting the CT flows through the 400 psig steam superheater, 400 psig evaporator, 50 psig steam superheater, 50 psig evaporator, economizer, and then out through the stack.

Water/Steam Flow – Boiler feedwater enters the economizer at 150°F and 80 psia. The heated water then flows to the 50 psig evaporator. Approximately 2/3 of the water flow entering the 50 psig evaporator is extracted as liquid water and sent to the 400 psig evaporator. The remaining 1/3 of the water exits the evaporator as 50 psig saturated steam. This flow then goes to the 50 psig superheater where it is heated to approximately 353°F.

The liquid water exiting the 50 psig evaporator flows to the 400 psig evaporator. Approximately 4% of the inlet water mass flow is blowdown from the system. The saturated steam exiting the evaporator is mixed with the 400 psig saturated steam coming from the waste heat boiler of the gasifier. The mixed saturated steam is then sent to the 400 psig superheater.

5.2.8 Offsites/Utilities (Unit 1000)

The outside battery limits (OSBL) facilities consists of systems provide to support the gasification units in terms of utilities and other auxiliary facilities. These are described in detail in Section 5.2.8 of the Subtask 3.2 Topical Report. Some of these facilities take advantage of the synergy with the existing industrial site and include:

- The steam system at two levels (400 psig and 50 psig)
- The condensate collection system
- The wastewater collection, treatment and disposal system

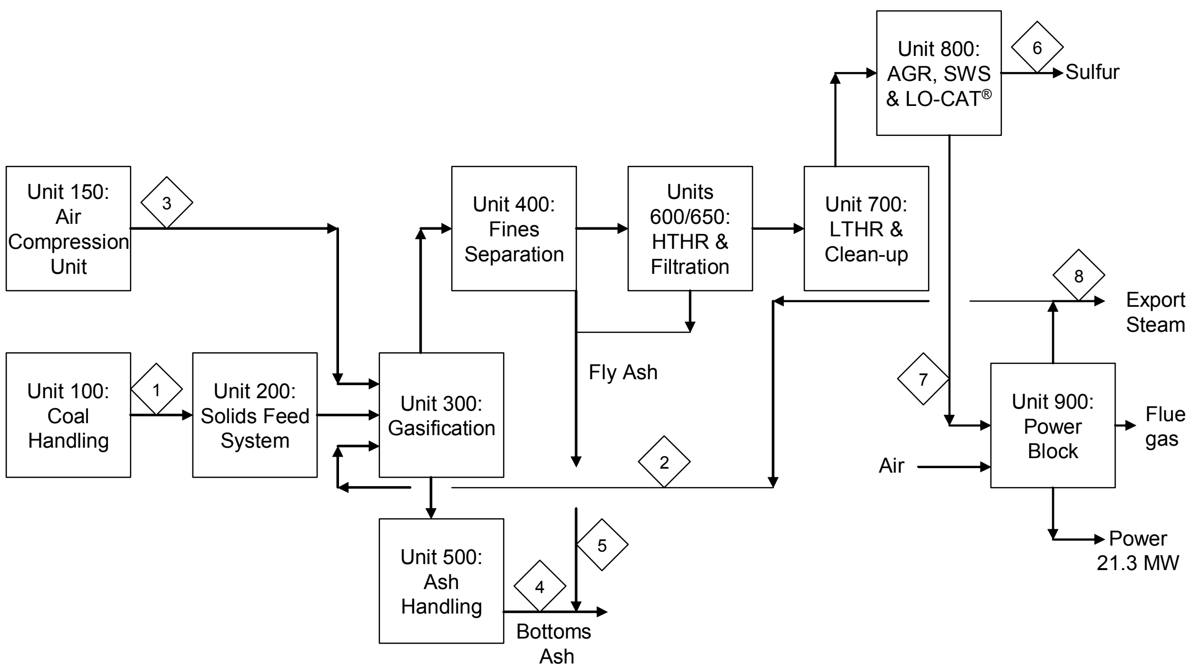
Other utilities that are grass-roots systems include:

- The safety shower and eye wash system
- The cooling water system
- The raw water and fire water systems
- The drinking (potable) water system
- The compressed air system
- The natural gas system
- The flare system
- The nitrogen system
- The electrical distribution
- Miscellaneous facilities

5.3 IMPROVEMENTS TO THE SUBTASK 3.2 AIR-BLOWN BASE CASE

The overall material balance generated using ASPEN is shown in Figure 5.2. The complete material balance is shown in Addendum C.

Figure 5.2 Simplified Block Flow Diagram and Material Balance for the Alternate Case



m Number	1	2	3	4	5	6	7	8
ription	Coal to Gasifier	Steam to Gasifier	Oxidant to Gasifier	Bottom Ash	Fly Ash	Sulfur Product	Clean Syngas to Gas Turbine	Export Steam Production
sure, MPa (psig)	0.10 (14.7)	2.90 (420)	2.86 (415)	0.24 (34.7)	0.24 (34.7)	NA	2.03 (295)	2.76 (400)
perature, K (F)	294 (70)	583 (550)	533 (500)	1283 (1850)	1228 (1750)	NA	322 (120)	561 (550)
Rate, kg/s (lb/hr)	3.8 (30,250)	1.2 (9,653)	12.1 (96,106)	0.2 (1,465)	0.2 (1,264)	0.1 (899)	15.5 (123,221)	13.3 (105,343)

For Subtask 3.3 numerous improvements have been made to the base case design shown in the Subtask 3.2 Topical Report. The offsites and utilities sections are essentially the same as for Subtask 3.2 except for minor capacities differences. The following subsections discuss these modifications.

5.3.1 Gasifier – One Train

As noted previously in this report, the base case used two 50% gasifier vessels operating in parallel to produce the syngas required by the two GE 10 turbines. The logic was that it was not practical to feed two gas turbines from a single gasifier on a regular basis, since it places too large a turndown demand on the equipment. However, it is expected that by 2015 GTI will have developed more operating experience with the U-GAS® technology and that turndown to 50% will not be a limiting factor. Therefore, economy of scale would advocate a single large gasifier be used to reduce the capital investment.

5.3.2 Feed Lockhopper System Replaced with Stamet Solids Pump

5.3.2.1 Background

Stamet, Inc., of North Hollywood, CA, has 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.[2] A trade-off study was performed to determine if an alternative coal feeding system could benefit the plant by reducing costs and/or improve operability. This trade off study evaluated Stamet's Posimetric® solids pump (so called "rock pump") against the traditional coal feeding system that was employed in Subtask 3.2. The traditional single train coal feeding system, which the Stamet system would replace, includes a lockhopper, a rotary valve, a surge drum and a screw feeder. This system also uses nitrogen to pressurize one of the lockhoppers. Two parallel lockhopper feed trains per gasifier were used in Subtask 3.2 to provide redundancy in the event of a disruption of the coal feed in one of the feed trains.

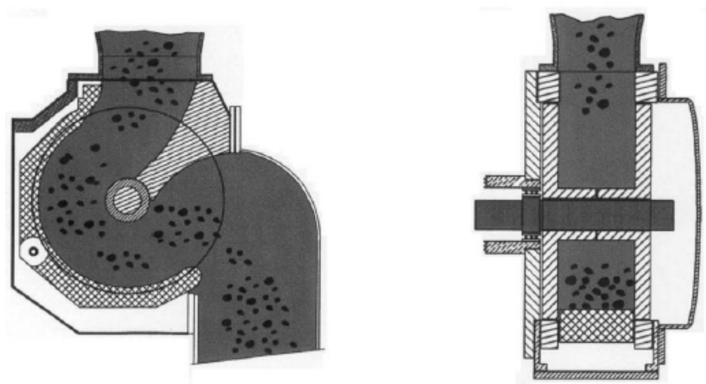
Stamet's solid feeding system uses a innovative technology known as Posimetric® solids feeding. The machine relies on a simple continuously rotating element, without valves or pressure vessels that also provides precise flow control. No nitrogen or other gas is needed to pressure the system or to maintain 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 5.3. 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 rotating 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.

More than 10 years of intensive research and development by Stamet has resulted in its successful commercialization with more than 200 Posimetric® feeder systems now installed in U.S. coal fired power plants. Posimetric® feeders are installed in power plants feeding pulverizers at capacities as low as 10 tph and up to 700 tph for large rotating hammermills. Stamet is currently testing this pump at higher pressures under two DOE-funded programs. This application would be a first of a kind at a higher pressure (greater than 250 psi). It is expected that by plant start-up (i.e., 2015) that operation of the Stamet pumps at higher pressures will be commercial and, therefore, justification for consideration herein.

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

Figure 5.3 Diagram of Stamet's Posimetric® Solids Pump[3]



Stamet, Inc. provided the budgetary quotes, operational data and background for the trade-off study that follows.

5.3.2.2 Trade-off Study

The Stamet pump system reduced the investment by about $465,000 compared to the conventional lockhopper system. However, the Stamet pumps use considerably more power than the screw feeder (157 kW versus 19 kW). The payout time for two Stamet pumps with 100% capacity to replace the conventional feed systems is 5 years. Although this payout is only marginal, the advantage of the simpler, more reliable feed system justifies the use of the Stamet pumps for the alternate design.

5.3.2.3 Process Description

Dried and sized coal is fed to one of two weigh hoppers on load cells that give accurate weights of material transferred to the gasifier. When one of the hoppers is discharging coal, the other hopper is being filled. The processed coal is fed to the gasifier via a Stamet pump. The Stamet pump effectively transfers the coal from atmospheric pressure to the operating pressure of the gasifier. Each Stamet pump is designed to deliver 15 tph to the gasifier (100% of the total gasifier design rate). A 100% spare pump is available that allows for complete redundancy in the event of disruption of coal feed.

5.3.3 Combined Ash Handling Systems

The bottoms ash and fly ash handling systems have been combined in the alternate design case. This not only simplifies the design, but 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

[3] www.stametinc.com/html/technology.html

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.

5.3.4 Candle Filters versus Third Stage Cyclone

Operating experience at the Wabash River plant with metallic candle filters has shown that this technology has evolved to the point that consideration be given to employing this technology in the Eastern Coal design. In addition information from Pall Corporation indicates that in the future ceramic filters may allow operation as high as 1850°F. This would allow the third stage cyclone to be replaced by ceramic candle filters and simplify the design of the downstream equipment. A trade-off study was made to determine the best location for the candle filters.

5.3.4.1 Background

Syngas filters remove the residual particulates that are not captured by the series of cyclones downstream of the gasifier vessel. The filter system selected is a Pall Corporation Gas Solid Separation System using metallic filters 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.

5.3.4.2 Trade-off Study

Two locations were considered. The warm filter location is downstream of the high temperature heat recovery (HTHR) unit. The operating temperature is about 480°F which allows for a metallic filter system above the gas dew point. The third stage cyclone and HTHR are unaffected by this change. The hot filter location would consider replacing the third stage cyclone with ceramic candle filters. The operating temperature is 1800ºF and would require ceramic filters, requiring a higher investment. However, the latter location makes the design and operation of the HTHR simpler and less expensive resulting in a higher overall availability. Budgetary estimates were obtained from Pall Corporation for both cases. The estimate for the hot filter case was adjusted downward assuming greater commercialization by 2015 and, accordingly, a lower cost (re. Section 6.2.2 of this report).

Table 5.1 Candle Filter Location Trade-off Study Results

	Hot Filter	Warm Filter
Plant Investment		
MM$	84.1	82.1
$/kW*	2,925	2,755
Overall Availability, %	91.74	89.86
Return on Investment, %	8.16	8.41

* Based on converting the steam export to power using an average turbine efficiency

The warm filter location results in a lower capital investment, but also a lower availability than the hot filter location. Based on the above cost and availabilities, the return on investment was higher for the warm filter location, and, thus, was the preferred option.

5.3.4.3 Process Description

Removal of the particulates from the syngas (480°F and 345 psia) exiting the HTHR system will reduce downstream complications due to the presence of the fine solids (e.g., erosion, agglomeration, etc.). Based on operating experience at the Wabash River plant, sintered metal candle filters were selected. The filters remove >99.99% of the particulates, and leave less than 0.15 lb/hr of solids remaining in the syngas stream exiting the filters. Maximum pressure drop across the filter assembly is approximately 5 psig.

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, 162 filter elements hang downward, with the particulates gathering on the outside of the filter elements. Each vessel is divided into six sections, with each section containing 27 individual filter elements. The individual filter elements have an outside diameter of 2.375 in and a length of 110 in. 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. Particulates are collected at the bottom of the vessel and removed by a lockhopper system.

5.3.5 Venturi Scrubber Replacing the Impingement Scrubber

The addition of the upstream candle filters allowed for a redesign of the downstream syngas scrubber. For Subtask 3.2 an impingement scrubber was employed to remove the particulates, as well as the ammonia, chlorides, light oils, and other contaminants. To ensure the proper operation of the acid gas removal (AGR) system and the gas turbines, it is critical to remove these undesirables from the syngas.

The presence of the upstream candle filters allows the impingement type scrubber to be replaced with a venturi scrubber. The venturi scrubber uses the differential between high velocity gases and free-flowing water to create droplets which entrap contaminants. Venturi scrubbers are widely used to remove particulates in similar applications and are more economical in water usage compared to scrubber columns. The upstream candle filters reduce the particles remaining in the syngas to a manageable size (less than 1.3 microns) and allow the water use to be reduced by about 35% of that for Subtask 3.2 (based on a conservative 20 gpm per thousand cubic feet per minute of syngas). Data available on the vendor (SLY, Inc.) web site shows that the venturi scrubber is estimated to be over 99% efficient with a pressure drop of approximately 1 psi. In addition the lower water usage results in significant capital savings since the sour water stripper becomes smaller (see Section 5.3.7).

5.3.6 Improved Heat Integration

5.3.6.1 Background

Heat integration refers to schemes for using available heat sources in the plant to improve overall efficiency. In order to improve a plant's efficiency, typically some level of heat integration has to be achieved. A Second Law type of analysis of plant performance often sheds light on where improvements can be made to enhance the efficiency. Such an analysis is often called an exergy analysis. Exergy is loosely defined as a universal measure of the work potential or quality of different forms of energy, relative to a given environment. Two streams can have the same amount of energy relative to a given ambient condition, but their potential to produce work can differ significantly. An exergy analysis of a system seeks to minimize exergy loss, a generally applicable quantity that measures inefficiency. Unlike a traditional energy balance analysis, which bases itself solely on the First Law of Thermodynamics, the exergy approach incorporates both the First and Second Laws.

As with any other theoretical analytical approaches when applied to a realistic scenario, the exergy method has its own limitations. Some of these limitations are listed here:

- It does not capture the effect of economics. An exergy analysis leads to ways of maximizing the work from a given energy source. Some schemes may result in significant efficiency improvement when implemented, however, they may require a large financial commitment that may not be economically attractive. For instance, in this study, waste heat could be used to heat up the air, which in turn can be used to dry the incoming coal. To implement such a scheme, the capital cost for the heat exchanger can be exorbitant due to relatively low heat transfer coefficients and low temperature differences across the heat exchanger.

- The exergy approach fails to capture any realistic difficulties associated with a specific process. For instance, the syngas heat recovery subsystem of this design can be improved further in terms of minimizing exergy loss by fully utilizing the sensible heat to produce steam. Besides the issue related to the cost, the presence of undesirables in the syngas prevents such a scheme from being implemented. The undesirables include chlorine, oils, and ammonium chloride, all of which present implementation difficulties due to possible corrosion and the severe reduction of process reliability.

- To realize some of the efficiency gains indicated by an exergy analysis, unreasonably large equipment may have to be used, and some of them may not even work. For example, a heat exchanger can minimize its exergy loss by using a very small temperature difference between the cold and hot streams. This could lead to a very large heat exchanger and large pressure drops for the two streams.

The exergy approach usually leads to the so called pinch analysis, which emphasizes:

- Minimizing the temperature difference between the hot and cold fluids in a heat exchanger.
- Minimizing the amount of heat rejected to the surroundings.
- Using high quality heat (high temperature heat) to produce high quality products (i.e. high temperature steam).
- Using low quality heat (i.e. low pressure steam or low temperature streams) to facilitate internal energy usage as much as possible. The high quality heat can be used to produce useful work, since it has a better work potential.

Overall, the design for Subtask 3.3 adhered to such principles. However, a number of ideas which could lead to a higher overall plant efficiency from an exergy perspective were not selected because they were impractical. These include:

- Fully utilizing the low pressure steam generated in the HRSG and the syngas heat recovery system by routing it through a low pressure condensing turbine. This would allow the system to generate an additional 1.5 MW of electricity. However, the investment for a condensing turbine of this scale for this application could not be economically justified.
- Generating additional electricity by producing 1,000 psig steam. Superheated 1,000 psig steam can be generated in the HRSG and the high temperature heat recovery system; this steam can be routed to a steam turbine before being extracted at 400 psig. By doing so, an additional 2 MW of electricity could be generated. Similar to the previous argument, the cost for the steam turbine did not justify the implementation of this idea.
- Using waste heat to dry the coal. Currently the system has considerable low quality heat which is not being utilized. For instance, there is more than 10

MMBtu/hr of low quality heat between 250°F and 110°F. This heat and other low quality heat in the HRSG can be used to heat air, which in turn, could be used to dry the coal. Large equipment costs are the primary reason why this idea was not adopted.

- Generating superheated steam in the high temperature heat recovery system. The syngas leaves the gasifiers and the ceramic candle filters at about 1750°F. It is possible to generate high temperature steam using this heat. After considering several factors, this idea was discarded, and only 400 psig saturated steam was produced in the high temperature heat recovery system. The rationale was:
    - Both the syngas and the steam are at high pressures. The cost for such an exchanger will be more than that if syngas is superheated in the HRSG.
    - The heat transfer coefficient is relatively poor for a gas-to-gas shell and tube heat exchanger compared to that of a heat exchanger producing saturated 400 psig steam. This means the heat transfer area will be relatively large for such an exchanger which will translate to an added cost.
    - By producing both saturated steam and superheated steam in the high temperature heat recovery system, the reliability of the entire system will suffer.

The improvements in heat integration are described below.

5.3.6.2 High Temperature Heat Recovery (HTHR)

Two changes were made to the high temperature heat recovery (HTHR) system:

1. The boiler feed water (BFW) preheated in the ash transport screw (S-501) was integrated into the design (versus returning the hot BFW to the industrial site).

2. An additional BFW preheater (E-602) was added downstream of the candle filters (F-651). This change is now practical because once the particles are removed by the candle filters, agglomerization of the light oils and particles is not a concern. Although ammonium chloride deposition is still a concern at this lower temperature, periodic water washing will remove any buildup in the exchanger. The result is that the temperature of the syngas to the venturi scrubber is reduced to 351°F (versus 600°F for Subtask 3.2). At the same time the boiler feed water to the tube fired boiler (E-601) is hotter (338 versus 250°F). This allows more energy from the syngas to be used for vaporization instead of sensible heat in the tube-fired boiler.

5.3.6.3 Low Temperature Heat Recovery (LTHR)

The temperature of the syngas entering the air cooler (E-702) is reduced from 236 to 215°F by preheating the BFW going to the HRSG.

5.3.6.4 Heat Recovery and Steam Generation (HRSG)

The HRSG design for the Subtask 3.2 was improved by adding a syngas preheater exchanger in the power block. 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, boiler feed water (BFW) is heated in the HRSG and then cooled by preheating the syngas to 340°F outside the HRSG. The BFW is heated in a coil within the HRSG between the medium pressure steam evaporator and the medium pressure steam superheater. For Subtask 3.3 clean syngas at 120°F and 295 psia is sent to the power block, where it is first heated to 340°F by BFW. Preheating the syngas in this manner not only increases the efficiency of the CTs, but also allows the low quality heat in the HRSG to be effectively used. This also allows the stack temperature to be reduced to 250°F which is still above the acid-dew point. Figure C.5 in Addendum C illustrates the CT/HRSG system.

5.3.7 Simplified Sour Water Stripper

With the addition of the candle filters, the sour water stripper (SWS) unit becomes simpler. Since there is very little if any solid particles present in the sour water (and those that are present would be less than 1.3 microns), the settling tank and filter are not needed. A three phase separating drum is used to separate any light oils that may be present. Any coal particles that are present will agglomerate with the oil. A day tank is still used as a back up to the SWS. However, the sour water rate is reduced by about 35% when compared to Subtask 3.2 due to the change in scrubber type (see Section 5.2.6). The temperature of the sour gases to the LO-CAT® unit (see Section 5.3.8) is maintained at less than 140°F.

5.3.8 LO-CAT® System Used for Sulfur Recovery

5.3.8.1 LO-CAT®

The sulfur removal/recovery unit was designed by Gas Technology Products LLC (GTP). The patented LO-CAT® process is a liquid reduction-oxidation (Redox) system that uses a chelated iron solution to convert H_2S to innocuous elemental sulfur. LO-CAT's environmentally safe catalyst does not use toxic chemicals and produces no hazardous waste byproducts. LO-CAT® units can be designed for better than 99.9% H_2S removal efficiency.

The LO-CAT® process was developed to provide an isothermal, low operating cost method for carrying out the modified Claus reaction:

$$H_2S + 1/2\ O_2 \rightarrow H_2O + S° \text{ (l)}$$

This reaction is accomplished in an aqueous scrubbing system using a water soluble metal ion capable of being oxidized by oxygen present in either ambient air or in the process gas stream, and has a suitable electropotential for oxidizing the sulfide ion to sulfur. More simply stated, the reaction is carried out in a water solution which contains a metal ion capable of removing electrons (negative charges) from a sulfide ion to form sulfur and in turn can transfer the electrons to oxygen (O_2) in the regeneration process. Although there are many metals which can perform these functions, iron was chosen for the LO-CAT® process because it is inexpensive and non-toxic.

There are other advantages to using the iron catalyst in this process. The iron catalyst is readily available and continuously regenerated in the process. It also 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 adsorber section, which converts the H_2S to elemental sulfur, and one in the oxidizer section, which regenerates the LO-CAT® catalyst.

In this application, the overall unit consists of two absorber units and one LO-CAT® autocirculation unit. The H_2S is removed from the syngas in the 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 the amine. The H_2S rich gas stream then is sent to the LO-CAT® unit where the H_2S is converted to sulfur. Any ammonia that is present in the H_2S rich gas stream will be vented from the LO-CAT® unit to the flare.

The LO-CAT® process replaced the conventional Claus/SCOT sulfur recovery system employed in Subtask 3.2. The LO-CAT® process has a 15% lower capital investment (9.2 versus 10.8 MM$), achieves the same sulfur removal and recovery as the conventional system, and is commercially proven. Although the operating costs (i.e., chemical consumption and utilities requirements) are higher, the lower capital cost offsets this, producing a process showing a lower total cost over five years of operation.

5.3.8.2 Other Sulfur Removal Technologies Considered

Two other sulfur removal systems were considered for the alternate case: CrystaSulf® and Morphysorb®. These are described below.

CrystaSulf®

CrystaSulf® is a non-aqueous sulfur recovery technology that can be used for direct treatment of gas streams containing H_2S. 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 via crystallization.

Sour gas enters the absorber where it contacts the lean CrystaSulf® solution. H_2S reacts with dissolved SO_2 to form dissolved sulfur. Sweet gas exits the absorber at 140-150°F. Rich solution flows from the absorber bottom to a flash vessel, where it is flashed to atmospheric pressure. The sweet gas exiting the flash tank can be recompressed. Liquid exiting the flash tank is sent to a crystallizer where its temperature is reduced to 105°F, causing the sulfur to crystallize. The excess solution overflows the crystallizer via a weir and flows by gravity to a heated solution tank. Lean solvent in the surge tank is heated and then pumped back to the absorber. As the slurry density increases, a batch filtration process is initiated, and a slipstream of the slurry from the bottom of the crystallizer is sent to a filter with a small positive displacement pump. The slurry is 15-20 wt% solids and flows from the crystallizer to an indexing-belt pressure filter. Filtrate is sent to the surge tank. The sulfur is rinsed with water and a wash solvent. The product sulfur nominally is 90+ wt% sulfur on a wet basis and 98+ wt% sulfur on a dry basis.

The LO-CAT® system provided more attractive economics than that of the CrystaSulf® system. LO-CAT® has a lower capital cost, lower yearly solvent cost, and a lower annual operating cost than the CrystaSulf® system.

Morphysorb®

Morphysorb® is a physical solvent-based process used for the bulk removal of H_2S and/or CO_2 from natural gas and other gaseous streams. The solvent consists of N-formylmorpholine and other morpholine derivatives. This process is particularly effective for high-pressure and high acid-gas applications. The Morphysorb® process takes the place of the amine unit and removes 99% of the H_2S from the syngas stream. A COS hydrolysis unit would be needed to remove any COS in the acid gas stream. Also a dehydration unit unit would have to be incorporated in the design to remove the moisture in the syngas stream. Both a Claus unit and a tailgas treatment unit are required after the Morphysorb® process.

The Morphysorb® system was not selected because of the high capital cost, high solvent cost, and the required removal of water from the syngas.

5.3.9 Spare Gasifier

A trade-off study was conducted for the addition of a spare gasifier train. The spare equipment would include the Solids Feeding System (unit 200), Gasification (unit 300), Fines Separation (unit 400), Ash Handling (unit 500), High Temperature Heat Recovery (unit 600) and Particulate Removal (unit 650). The first location in the syngas flow scheme where block valves can be located (which will insure a proper seat) is downstream of the candle filters. Upstream of the candle filters the presence of particulates will not allow the valve to seat properly, making isolation impossible. It is assumed that the spare gasifier is maintained in a condition such that is can be brought on-stream fairly rapidly. The additional equipment will cost about 9 MM$, but will improve the overall availability from 89.8 to 94.8 percent for the warm filter case. The results of a financial analysis showed that the return on investment will be lower with the spare gasifier, and therefore, it was not included in the final design.

Table 5.2 Spare Gasifier Trade-off Study Results

	Hot Filter		Warm Filter	
Number of Gasifiers	1	2	1	2
Plant Investment				
MM$	84.1	95.5	82.1	91.7
$/kW*	2,925	3,320	2,855	3,190
Overall Availability,** %	91.74	94.84	89.86	94.76
Return on Investment, %	8.16	5.56	8.41	6.77

***** Based on converting the steam export to power using an average turbine efficiency
** Overall availability is without scheduled maintenance.

5.3.10 Other

Many IGCC designs employ the use of air extraction from the CT compressor as the initial stage of compression for the gasifier air (or for 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 design because the GE 10 turbine has not been thoroughly evaluated for air extraction.

5.3.11 Capital Savings

The investment due to the changes described above has been reduced by 8.8%, from 90.0 to 82.1 MM$. Details are shown in Section 6.2.3.

5.4 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 control 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 are smaller and less expensive systems. In addition hydrogen sulfide is more reactive and, therefore, easier to remove than sulfur dioxide.

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

5.4.1 Sulfur

Sulfur is removed from the syngas by the LO-CAT® process. It is substantially different from the multi-step process employed in Subtask 3.2, which removes H_2S from the syngas by UOP's Selective AGFS process, and that finally produces elemental sulfur by a Claus process with a SCOT unit. The LO-CAT® process does not use any toxic chemicals and does not produce any hazardous waste byproducts. In this alternate case, the LO-CAT® process replaces the Claus and SCOT processes.

The combined SO_2 release rate from the gas turbine and the incinerator is 10.2 lb/hr or 0.026 lb per MMBtu (HHV) of energy input. This is about the same as the Subtask 3.2 base case (air-blown). The net result of this processing scheme is an overall sulfur removal rate of 99.1%.

5.4.2 NOx and CO

The firing of combustion turbines on coal-derived syngas requires the proper design of turbine components. The specific design influences the emission rates of NOx and CO. For this application, two GE 10 combustion turbines are used. The GE 10 turbine is not yet commercially available for use on coal-derived syngas. Communication with GE engineers indicates that although they expect to be able to deliver the turbines within the next two years, they are not yet able to guarantee NOx, CO or other emission levels without successful combustion testing. GE currently estimates NOx emission levels for this application ranging from 65 – 90 ppmvd @ 15% O_2 (0.25 lb/MMBtu – 0.35 lb/MMBtu). A prior CO emissions estimate for the GE 10 was 20 ppmvd @ 15% O_2 (~0.04 lb/MMBtu). Including balance of plant emissions (e.g., flare), total facility CO emissions are estimated at less than 0.05 lb/MMBtu.

The NOx estimates from GE are approximately twice the current new source performance standards for coal-fired utility boilers (currently 1.6 lb/MWh gross energy output or ~0.15 lb/MMBtu fuel input). Furthermore, the specific NOx emission rate required for this type of facility would be highly site specific and depend on a number of factors including local area designation (attainment vs. non-attainment), proximity to sensitive areas, and others factors including corporate emission control philosophy.[4] The use of post combustion means to reduce NOx (e.g., selective catalytic reduction (SCR)) has been suggested for IGCC applications. However, such a requirement would result in increased capital and O&M costs as well as reduced performance at the power block due to increased backpressure on the turbine. The negative effects on the efficiency and economics will make it hard for industries to endorse the system.

Clearly, before such a system can be deployed the NOx emissions need to be reduced, preferably without the use of controls down stream of the turbine. Despite the relatively high NOx emission estimates for the GE 10, low NOx gas turbines have been developed and used for many applications, including IGCC. In the past two decades, significant progress has been achieved in reducing the NOx emissions without the use of SCR. For example, the 7FA turbine used at TECO's Polk Power station initially operated with NOx emissions less than 25 ppmvd @ 15% O_2. Recently, its emissions

[4] Because this study represents a replacement power application at an existing industrial facility, the permitting process would likely fall under New Source Review requirements and would therefore undergo a thorough evaluation in respect to site specific conditions, including the opportunity to buy or trade emissions credits. The determination of likely emission permit limits for NOx and other pollutants are beyond the scope of this study. The discussion included here is only to highlight the need for additional emission performance data of the GE 10 turbine fired with coal-derived syngas.

were further reduced to less than 15 ppmvd @ 15% O_2 by supplementing diluent nitrogen with water dilution. The GE 7FA also was used for the Wabash River repowering project. The NOx emission level for that application is about 25 ppmvd @ 15% O_2. Other facilities that combust syngas in large stationary gas turbine combined-cycle projects have had NOx limits ranging between 16 and 20 ppmvd.[5] By improving gas turbine combustor designs (i.e. using the can-annular combustion system) for industrial-scale engines employing low Btu gas and supplementing diluents such as H_2O, CO_2, and N_2, GE has consistently demonstrated that reducing NOx to low levels without the use of SCR is achievable. GE believes that 0.04 lb/million Btu is an achievable target for IGCC applications.[6]

Carbon monoxide emissions results due to incomplete combustion of carbon based fuels and are primarily a result of highway and off road transportation sources. While CO emissions are not regulated with 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, the flare system, and possible fugitive emissions from equipment leaks.

Most of the upstate New York area is classified as attainment[7] for CO and therefore would be subject to Best Available Control Technology (BACT) for CO control. Control technologies for carbon monoxide emissions identified as potential BACT by Global Energy include good combustion techniques and possibly the use of an oxidation catalyst.

While specific emission limits 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/MWh to 2.2 lb/MWh 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/MWh, equivalent to 15 ppm on syngas.

Most of the developments, however, have been focused on larger systems, which include GE models 6B, 6FA, 7EA, 7FA, 9E, 9EC, 9FA, and the newer H-type gas turbines. It is unclear what level of NOx and CO emissions can be attained using industrial size gas turbines, such as the GE 10. GE claims low and ultra-low NOx emissions can be achieved. The task is how to transfer the reduction technologies achieved in large turbines to smaller machines.

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5 Ratafia-Brown, J., et.al., Major Environmental Aspects of Gasification-Based Power Generation Technologies, Final Report. US DOE National Energy Technology Laboratory, December 2002.

6 Outlook on Integrated Gasification Combined Cycle Technology, Testimony before subcommittee on Clean Air, wetlands and climate change, Edward Lowe, GE Gas Turbine-Combined Cycle Product Line Manager, January 29, 2002

7 The Syracuse area is identified as a maintenance area for CO, previously identified as marginal non-attainment for CO ($\leq$ 12.7 ppm). http://www.epa.gov/oar/oaqps/greenbk/cmcs.html#NEW%20YORK

5.4.3 Mercury

Mercury emissions for larger coal-fired electric generators are not currently regulated although several proposed regulations are currently under review. 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 0.00036 lb/hr (0.95 lb/TBtu). Mercury emissions of this rate are equivalent to a stack gas concentration of around 1µg/Nm^3, approaching the detection limit of current mercury measurement technologies.

5.4.4 Water

For this study preliminary modeling of syngas organic content suggests that traditional aerobic wastewater treatment would be the most effective technology for destruction of trace organic compounds of the type expected in this study. Since this study represents a repowering of an existing chemical facility, it is expected that their existing water treatment system can readily handle the levels of organic contamination in the wastewater. This is because of three reasons: 1) the volume of water sent to the wastewater treatment plant from the gasification operations is small compared to the daily quantities handled by the existing wastewater treatment operations (< 1% of the 15 to 20 million gallons treated per day); 2) the total loading of organic material to the wastewater treatment plant (all wastewater streams combined) is extremely low and 3) the organic material is expected to be of a type that is readily consumed in such wastewater plants.

In the case of a greenfield plant design, treatment of the wastewater using standard methods may be sufficient to assure adequate destruction of similar mass loadings of trace organic material. However, for an equivalent sized gasification only plant, discharge of water from gasification operations would result in similar mass loading but higher concentrations (due to the absence of mixing with other large volume wastewater streams). Water quality requirements for the receiving streams should be reviewed to determine the method and degree of destruction required. The cost of wastewater treatment for a greenfield system has not been included in this study.

5.4.5 Emissions Summary

Particulate emissions are considered to be negligible. All particulates in the syngas are removed by cyclones attached to the gasifiers, metallic candle filters downstream of the high temperature syngas cooler, and a venturi scrubber upstream of the sulfur removal unit. 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.

Depending upon the specific situation and the emission levels of the facilities that this gasification plant will replace, this may allow the industrial facility to adjust their policy with respect the sulfur dioxide and nitrogen oxide credits. If they are selling credits, they may have more credits to sell, and if they are purchasing credits, they may be able to purchase less credits. Either case would be beneficial to the facility and increase the net return on the gasification facility.

5.5 DESIGN COMPARISON BETWEEN SUBTASKS 3.2 AND 3.3

Table 5.3 compares the yields for the base case (Subtask 3.2) and the alternate case (Subtask 3.3). In both cases the coal feed rate is the same, 345.7 tpd (moisture-free). The alternate case also recovers 622 lb/hr more dry ash by removing the fly ash from the syngas as a dry solid via the candle filters instead of as a wet sludge from the sour water stripper unit.

The export power is 2% less for the alternate case due to the addition of the Stamet solids pump and LO-CAT® unit. The high-pressure steam production is increased by 3-1/2% as a result of the improved heat integration and reduced low-pressure steam requirements. The boiler feed water (BFW) from the ash screws is used internally for the alternate case, reducing the overall BFW requirements. The condensate returned to the industrial facility has decreased because less 50 psig steam is consumed in the LO-CAT® unit as opposed to the Claus unit. The overall thermal efficiency is increased from 48.4% to 49.7%.

Table 5.3 Comparison of Designs

	Alternate	Base	Difference
<u>Design Inputs</u>			
Coal Feed, moisture-free tpd	345.7	345.7	0
Coal Feed, moisture-free lb/hr	28,810	28,810	0
Fuel (Natural Gas), MMBtu/hr	-	5.1	-5.1
Makeup Water Input from the Industrial Facility			
Boiler Feed Water, gpm	418	495	-77
Quench Water, gpm	0	30	-30
Cooling Tower Makeup Water, gpm	58	53	+5
<u>Design Outputs</u>			
Export Power, MW	21.3	21.7	-0.4
Export Steam (400 psig, 550°F), Mlb/hr	105.34	101.72	+3.62
Sulfur, lb/hr	899	899	0
Ash, lb/hr	2,719	2,097	+622
Condensate (to industrial facility), Mlb/hr	54.43	60.86	-6.43
Cold Gas Efficiency, % (HHV basis)	79.3	79.3	0
Net CHP Efficacy, % (HHV basis)	49.7	48.4	+1.3

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). 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) facilities. 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 an upstate New York site. The labor rates associated with the construction have been adjusted for the labor rates and productivity in upstate New York.

6.2.2 Methodology

6.2.2.1 Equipment Design

The equipment for the coal fueled IGCC case was designed using the material and energy (M&E) balances developed specifically for Subtask 3.3. Multiple groups had input to the M&E balances. Raymond Professional Group (RPG) developed the coal handling and preparation area. GTI developed the gasification island. Gas Technology Products LLC (GTP) prepared the acid gas removal and LO-CAT® 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 ASPEN and Gatecycle computer simulations along with previous experience with similar systems. The M&E balances and PFDs are given in Addendum C.

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, coal handling and drying. Air Products provided the design and cost for the air compressor, Unit 150. The equipment sizing for Units 200-500 (ex. Stamet pump) was prepared by GTI. The design for the equipment in Units 600 through 1000 (excluding the acid gas removal and LO-CAT® systems) were prepared by Nexant and NETL using the ASPEN and Gatecycle heat and material balances as a basis. The acid gas removal and LO-CAT® systems design was provided by GTP. 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 upstate New York 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), air compressor (from Air Products), gas turbine (from General Electric), HRSG (from Vogt Power), mercury removal (from Calgon Carbon), solids pump (from Stamet), and the gas removal and LO-CAT® units (from GTP).

It is expected that some advanced technologies being presently developed will be commercially proven by 2015, when the current study's start-up is based. These technologies, high-temperature ceramic candle filters and the Stamet pumps, have been considered for this design. The commercial price quotes from vendors for first-of-a-kind equipment will likely drop in order to be competitive and as the technology develops. It is an accepted principle for advancing new technology to commercial maturity that the first-of-a-kind commercial plant is significantly higher in cost to build than subsequent plants and does not provide adequate information on all operating, maintenance, and cost issues[1]. These factors have been taken into account when estimating the cost for advanced technologies in 2015.

It was determined that a reduction in equipment cost of 3% per year (in nominal cost) can be expected for the next 10 years for both the ceramic candle filters and high pressure Stamet pumps. This assumes greater commercial use of both these technologies. Both historical analogies and vendor input were used in developing an estimate for the extent of the cost reduction. Cost data for the wider commercial application of both Flue Gas Desulfurization and Selective Catalytic Reforming technologies showed similar cost reductions during periods of wider technological acceptance[2]. While it is difficult to predict the expected future demand for advanced filter and pumping technologies, it can be reasonably assumed that the demand will correspond to some extent to the construction of gasification facilities. With the potential for a number of new facilities between 2005 and 2015, it can be assumed that the advanced technologies considered in this design will have significant commercial and research interest.

6.2.3 Results

Table 6.1 shows the EPC (engineering, procurement and construction) cost for the Subtask 3.3 Alternate Design for the Eastern Coal Case. These costs are on a second quarter 2004 basis. The investment is adjusted for labor rates and productivity in New York.

1 National Academies of Science, Commission on Engineering and Technical Systems, Coal: Energy for the Future, National Academies Press, 1995.

2 Professor Edward Rubin, Carnegie Mellon University, "The Government Role in Environmental Technology Innovation", Clean Coal Technology Roadmap Workshop, Calgary, AB, Canada, 20 March 2003.

Table 6.1 Capital Cost Summary, Alternate Eastern Coal Case
(US$)

Description	Total Project Cost*	Percent of Total
Coal Preparation and Handling	7,551,000	9.2
Air Compressor	4,321,000	5.3
Coal Feeding	1,012,000	1.2
Gasification	3,738,000	4.6
Dust Removal	937,000	1.1
Ash Removal	1,427,000	1.7
Gas Cooling (HTHR)	1,794,000	2.2
Particulate Removal	1,029,000	1.3
Gas Cleaning and LTHR	2,014,000	2.5
Sour Water Stripper	1,644,000	2.0
Acid Gas Removal and Sulfur Recovery	9,200,000	11.2
Gas Turbine, and HRSG	29,939,000	36.5
Offsites and Auxiliaries	14,583,000	17.8
Buildings	2,913,000	3.5
TOTAL	**82,103,000**	**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).

In order to keep the investment cost as low as possible, modular construction was considered wherever possible. However, the size of the plant made this prohibitive.

Table 6.2 compares the investment for Subtasks 3.2 and 3.3.

Table 6.2 Capital Cost Compared, Eastern Coal Cases
(US$)

Description	Base Case	Alternate Case	Delta	Percent Reduction
Coal Preparation and Handling	7,551,000	7,551,000	0	0
Air Compression	4,321,000	4,321,000	0	0
Coal Feeding	1,739,000	1,012,000	727,000	41.8
Gasification	5,168,000	3,738,000	1,430,000	27.7
Dust Removal	2,629,000	937,000	1,692,000	64.4
Ash Removal	2,217,000	1,427,000	791,000	35.7
Gas Cooling (HTHR)	2,373,000	1,794,000	579,000	24.4
Particulate Removal	0	1,029,000	(1,029,000)	-
Gas Cleaning and LTHR	2,812,000	2,014,000	799,000	28.4
Sour Water Stripper	2,979,000	1,644,000	1,334,000	44.8
Acid Gas Removal and Sulfur Recovery	10,800,000	9,200,000	1,600,000	14.8
Gas Turbine, and HRSG	29,890,000	29,939,000	(49,000)	(0.2)
Offsites and Auxiliaries	14,583,000	14,583,000	0	0
Buildings	2,913,000	2,913,000	0	0
TOTAL	**89,976,000**	**82,103,000**	**7,873,000**	**8.8**

The 41.8% cost reduction in the coal feeding system comes from the removal of the conventional screw feeder system and replacing it with the Stamet solids pump. The

savings in the gasification, ash removal, and gas cooling areas primarily comes from the economy of scale of using a single larger gasifier system instead of two smaller ones. The capital cost of the dust removal unit is reduced by combining portions of the fly ash and bottoms ash handling systems. The new particulate removal area contains the cost for the metallic filters which now remove the residual particulates from the syngas instead of the water scrubber. The gas cleaning area is less expensive because the single venturi scrubber replaced the two impingement scrubber columns. The cost savings in the sour water stripper are the result of reduced water usage in the syngas scrubbing area. The savings in the acid gas removal and sulfur recovery is the direct result of replacing the Claus and SCOT units with the LO-CAT® sulfur recovery system. The HRSG is slightly more expensive because a new coil was added to preheat the syngas going to the gas turbine. Overall the savings is 7.9 MM$ or 8.8% less than the base case.

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.

Availability analyses were performed for the Subtask 3.3 design to account for forced and scheduled outages to determine expected annual cash flows. Based on these cash flows, financial analyses were performed to evaluate the comparative economics of two possible Subtask 3.3 plant configurations with and without a spare gasification train.

The effect of sparing (back-up equipment or parallel trains of reduced capacity) can have a significant effect on the 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. However, sparing results in an increase in the investment cost. As stated in the trade-off study in Section 5.3.9, sparing the gasifier has a negative impact on the economics because the high cost of the gasifier and the associated equipment do not compensate for the increased availability.

For this analysis, most operations of Subtask 3.3, exclusive of the gasifier island, coal handling and sulfur recovery, 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 3.3 is similar to the analysis in Subtask 3.2 with the following notable exceptions: candle filters are present in the gasification train and the HRSG availability is improved over that in Subtask 3.2. In Subtask 3.2, the availability of the HRSG is quite low, since it was based on the operation data obtained from the Wabash River Repowering Project. However, the low availability of the HRSG for Wabash River was a result of design flaws which did not allow for tube expansions in the HRSG. It is expected that lessons will be learned prior to the start-up of this case, creating solutions for the past flaws. Make-up of the individual blocks as well as availabilities of the component units is presented in Table 6.3. 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

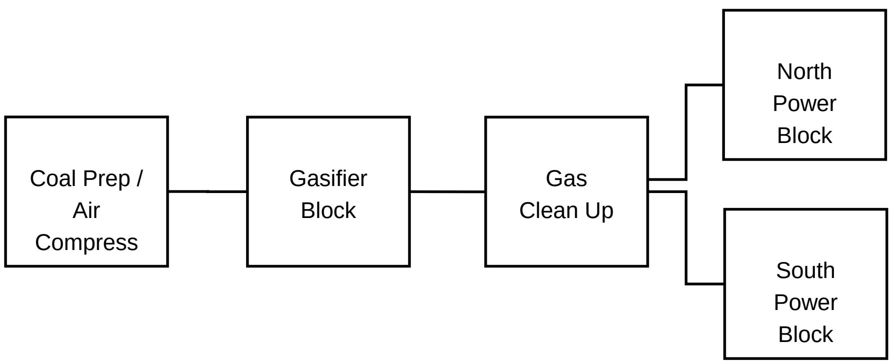


Table 6.3 Availability Estimates

Plant Section	Availability
Coal Prep/Air Compression	
Coal Drying	100%
Coal Crushing	99.95%
ASU	99.84%
Gasifier Block	
Gasifier Island	97.49%
High Temperature Heat Recovery	98.95%
Candle Filter	99%
Medium Temperature Heat Recovery	99.0%
Water Scrubber	99.87%
Gas Clean-up	
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	97.4%

* 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 6.4 presents availability calculations for individual state capabilities; probability of operating an individual state (e.g., 2 of 2 parallel power trains, 1 of 2 parallel power trains operating excluding scheduled maintenance) as well as equivalent availability, both with and without 21 days per year of scheduled outage.

Table 6.4 Calculated Power Availabilities

Coal Prep*		99.95%
Syngas Operations**		94.01%
Power Block***		
	2 of 2	91.46%
	1 of 2	99.81%
Equivalent Availability		
	w/o Scheduled Maintenance	89.86%
	w/ Scheduled Maintenance	84.69%

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 and CT/HRSG in operation at a given time) and export power. The equivalent availability for the alternate case is 84.69%, which is a percentage point lower than the base case presented in Subtask 3.2 (85.67% availability). Table 6.5 is a summary of the design and annual average plant flow rates.

Table 6.5 Design and Annual Average Flow Rates

	Export Power (MW)	Export Steam (klb/hr)	As Received Coal (lb/hr)	Sulfur (lb/hr)	Ash (lb/hr)
Design	21.33	105.34	30,250	899	2,729
w/o Scheduled Maintenance	19.17	94.66	27,183	808	2,452
w/ Scheduled Maintenance	18.06	89.21	25,619	761	2,311

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 financial parameters for the Eastern Coal Case are given in the Subtask 3.2 Topical Report.

The plant EPC cost used in the financial model is shown in Table 6.1. An owner's contingency fee of 25% was added to the cost of the gasifier island (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 additional capital that may be required during plan construction. Based on the cost of the gasification island and the plant EPC cost, the overall contingency for the entire plant was revised to 15.46% to reflect the higher contingency value of the gasifier. The plant feed and product rates are adjusted from those 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 was calculated to be 89.86%. 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 84.69% is the "Guaranteed Availability" of 89.86% times the percentage of time the plant is scheduled to operate (8,256 hours/year, or 94.25% of the time).

6.4.1 Results

For an air-blown facility with EPC costs of 82.1 MM$ and a project life of 20 years, the return on investment (ROI) is expected to be 8.4%, with a net present value (NPV) of

-5.3 MM$ given a 10% discount factor. Table 6.6 shows the rate of return, NPV, payback year, and required electricity and steam selling prices to obtain a 12% ROI with the other items fixed. There are two major products from this facility, electricity and steam, and both must be considered when determining the suitability of this project. Besides the base case, a “High” and “Low” sensitivity is listed reflecting the current cost accuracy assumption of +30/-15%.

Table 6.6 Alternate Case Financial Cost Summary

	Alternate	**Low -15% EPC**	**High +30% EPC**
ROI, %	8.4	12.9	0.15
NPV, MM$ (10% Discount Rate)	-5.3	8.6	-38.8
Number of years till payback	14	9	>20
Electricity Selling Price for 12% ROI, cents/kWh	8.5	7.1	11.3
Steam Selling Price for 12% ROI, $/ton	14.1	8.4	25.4

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

Table 6.7 Alternate Case Total Project Costs

Construction/Project Cost (in Thousand Dollars)		
Capital Costs	**Category**	**Percentage**
EPC Costs	$82,103	72%
Initial Working Capital	$1,089	1%
Owner's Contingency (% of EPC Costs)	$12,689	11%
Development Fee (% of EPC Costs)	$3,284	%3
Start-up (% of EPC Costs)	$1,642	%1
Initial Debt Reserve Fund	$0	%0
Owner's Cost (combined with development and start-up costs)	$3,284	%3
Additional Capital Cost	$0	0%
Total Capital Costs	**$104,092**	**91%**
Financing Costs		
Interest During Construction	$7,816	7%
Financing Fee	$2,216	2%
Additional Financing Cost #1	$0	0%
Additional Financing Cost #2	$0	0%
Total Financing Costs	**$10,032**	**9%**
Total Project Cost/Uses of Funds	**$114,124**	**100%**
Sources of Funds		
Equity	$38,802	34%
Debt	$75,322	66%
Total Sources of Funds	**$114,124**	**100%**

When compared to the air-blown base case of Subtask 3.2, Tables 6.6 and 6.7 illustrate the improvements in financial performance of the plant that resulted from the Subtask 3.3 process design improvements. All parameters in Table 6.6 demonstrate a significant improvement compared to the base case as shown from an increased ROI and a shorter payback period. Table 6.8 compares Subtask 3.3 with the air-blown Subtask 3.2 base case.

Table 6.8 Financial Cost Summary Comparisons, Subtask 3.3 Alternate Air-Blown Case vs. Subtask 3.2 Air-Blown Base Case

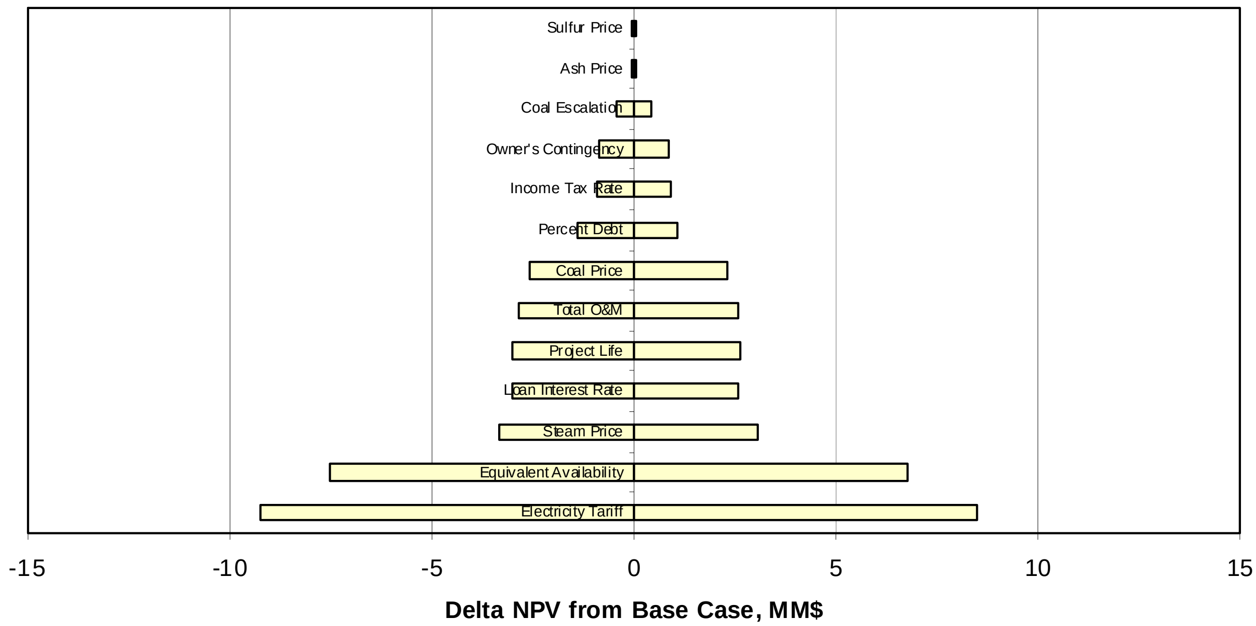
	Subtask 3.3	**Subtask 3.2**
ROI, %	8.4	5.9
NPV, MM$ (10% Discount Rate)	-5.3	-14.6
Number of years to Payback	14	17
Electricity Selling Price for 12% ROI, cents/kWh	8.5	9.0
Steam Selling Price for 12% ROI, $/ton	14.1	17.6

6.4.2 Sensitivities

A number of financial parameters that were likely to influence overall economic performance were varied to determine the project financial sensitivities. Changes to the input that were deemed to be reasonable based on previous sensitivity analysis, commodity input ranges, and team estimates were entered into the model. The impact that these changes had on the NPV and ROI were recorded, along with the percent change to the parameter that was modified. The financial impacts were normalized by calculating the overall impact relative to the size of the modification. The variables and their impact on the financial outputs were then ranked to determine the parameters with the highest sensitivity.

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 changes of NPV from the base case are shown. 10% changes were used to give a common ground by which all variables were evaluated. It is worthwhile to note, however, that the range of realistic possibilities for each variable differs significantly. For example, 10% changes in the availability or income tax rate should capture the majority of long-term variations. This would not be the case with variables such as coal price and electricity tariff which could vary by much more than 10%. The relative significance and range of possible values were considered in determining which items have the most impact on the model.

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 nearly 8 MM$ while decreasing it by 10% results in a decreased net present value by nearly 9 MM$. In this case, "Electricity Tariff" is used to refer to the sales value for the electricity that the plant generates. This variable also was the most significant in Subtask 3.2. The guaranteed availability also is very significant. Although a theoretical 10% increase of the guaranteed availability (as defined on page 6-8) would result in an unrealistic 100% guaranteed availability, operating at or near 100% would result in a net present value increase of more than 6 MM$. By reducing the availability by 10%, the net present value is reduced by more than 7 MM$. All other variables associated with the amount of time the plant is operating (e.g., operating hours (impacting equivalent availability) and plant life) also had a significant impact on the plant economics.

As was the case for Subtask 3.2, the remainder of the input variables impacted the plant economics to a lower extent. The steam price, plant life, and interest rate were next in importance, with all other items showing a less significant impact. While the remaining items had a less significant impact relative to those described above, many could push the project to a near zero or negative net present value within the ±10% range evaluated here.

The model relies more heavily on the electricity tariff for the economic outcome because electricity accounts for 72% of the total revenue stream for the base case facility.

Although steam also is a primary product for this facility, the contribution to total revenue is only 27%, making this variable less sensitive to fluctuations in price. Figure 6.3 shows the relationship between the ROI and electricity tariff. The reference power price of 80 $/MWh is indicated by an arrow on the abscissa.

Figure 6.3 Effect of Electricity Tariff on Return on Investment

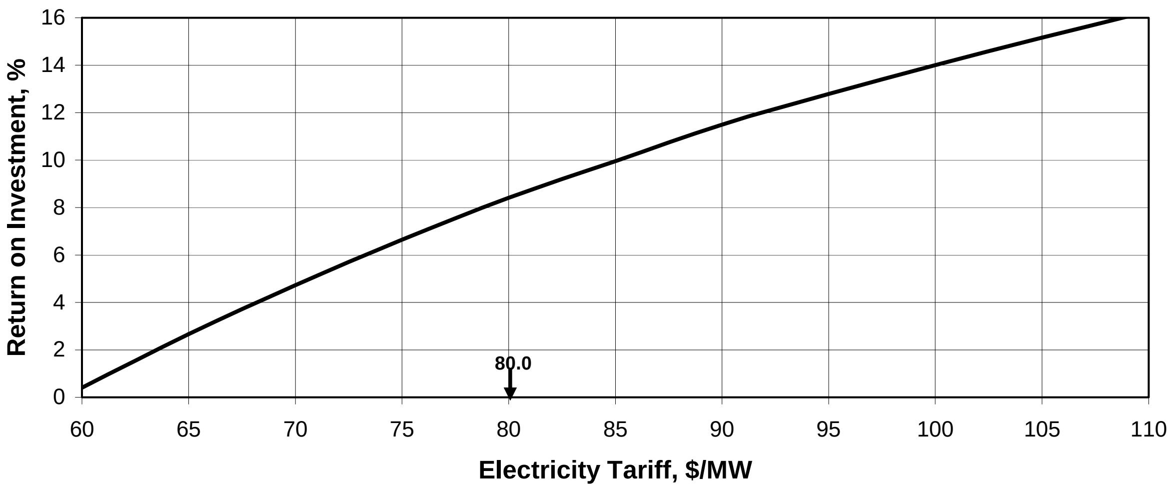
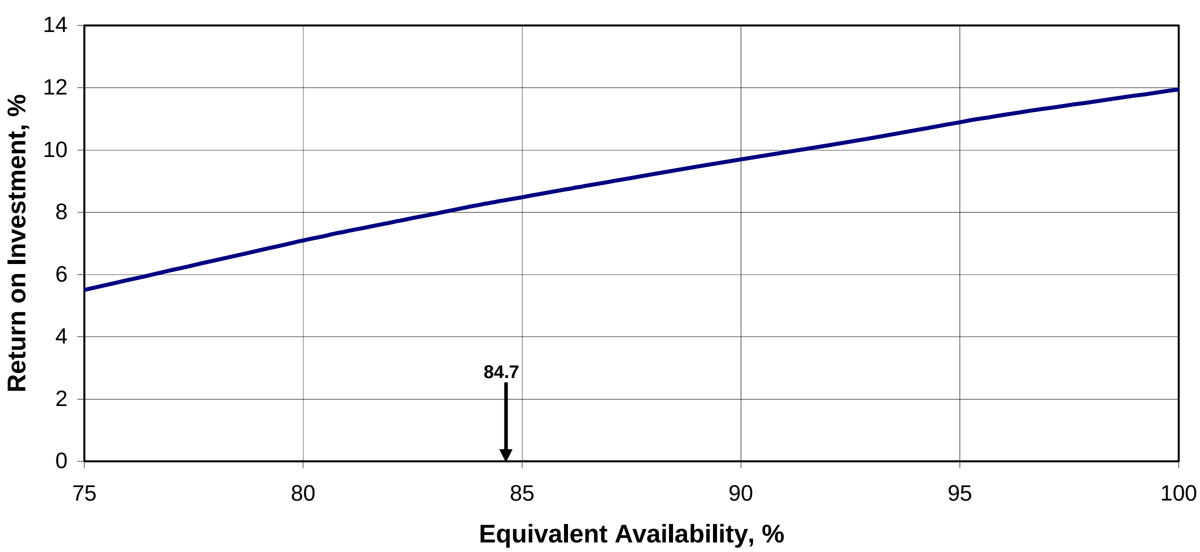


Figure 6.4 shows the effect of varying the guaranteed availability has on the NPV assuming a 10% discount rate. At the projected availability of 84.7%, the alternate case has an ROI of 8.4%.

Figure 6.4 Effect of Availability on Return on Investment



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 the overall plant economics is similar to that of Subtask 3.2. 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.

As with Subtask 3.2, 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, will not have the impact of these two items.

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 for 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 gasification. While the inputs 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, the above sensitivity analysis can provide insight into the outcome for plants with somewhat different base assumptions.

7.1 INTRODUCTION

Because of scope and because the impacts were expected to be minimal, some items that could improve the process either by increasing the efficiency or reducing the cost were identified, but not rigorously analyzed. Another item, high temperature ceramic filters, was rejected because of high cost. It should be reexamined in the future to determine if improved manufacturing techniques could reduce the cost and make it economic.

7.2 IMPROVED SULFUR REMOVAL METHODS

Other methods for removing the sulfur from the syngas were considered including Morphysorb® and CrystaSulf®. 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. The Morphysorb® process is an alternative adsorption process for scrubbing H_2S from a gas stream. It can remove H_2S to leave a residual H_2S content of about 10 ppm compared to MDEA which leaves a residual H_2S content of about 30 ppm. It requires a pretreatment step to convert COS to H_2S just like MDEA.

While capital costs were low, both of these processes were rejected because their chemical costs were too high, creating operating costs that made these technologies uneconomic relative to the amine, Claus, and SCOT processes. The LO-CAT® process was accepted because it has a lower total cost (capital and operating cost) than the Claus and SCOT processes.

The SCOHS (Selective Catalytic Oxidation of Hydrogen Sulfide) process being developed by NETL has the potential to be a simple and cost effective system for sulfur removal. However, the process is still in the early stages of development and the developers were not able to provide cost estimates for this application. This process should be revisited when development is further along.

Deeper sulfur removal methods may be required in the future if NOx reduction requirements become more stringent. There are other solvents that can be used instead of MDEA for removing acid gases from the syngas, but they also remove CO_2, which reduces the power output from the combustion turbine. Furthermore, the removal of the CO_2 diluent from the syngas require that other diluents be mixed with the syngas to control NOx.

7.3 WARM MERCURY AND SULFUR REMOVAL METHODS

The current design cools the syngas to 110°F before passing it over a bed of activated carbon for mercury removal. The cleaned syngas is then reheated to 300°F in the

HRSG before going to the combustion turbine. Since the syngas has to be cooled to 110°F for the amine system, there is no incentive to remove the mercury at a higher temperature because direct adsorption on activated carbon is effective and commercially proven. Mercury removal systems at warmer temperatures are being developed. However, until an effective sulfur removal system that operates at a warmer temperature is available there is no incentive to use a warm mercury removal system.

7.4 HIGH TEMPERATURE CERAMIC FILTERS

The use of high temperature ceramic filters, operating at the gasifier outlet temperature, located just after the second stage cyclones, would simplify the design and improve the operation of the high temperature syngas cooler. The high temperature ceramic filters, which would replace the third stage cyclones, would effectively remove all the particulates from the syngas before it enters the high temperature syngas cooler. The third stage cyclones allow a portion of the particulates to remain in the syngas and enter the syngas cooler thus making the design of the cooler more complicated. Hard facing is required for erosion protection. A careful design is required to prevent particle deposition and plugging.

Replacing the third stage syngas cyclones with ceramic filters would eliminate the erosion and deposition problems in the high temperature heat recovery system. This service has been shown to be a maintenance problem in past gasifier designs. Furthermore, downstream particulate removal equipment would be eliminated. Based on past experience, it would be expected that a gasifier design employing a ceramic filter system would have a somewhat higher overall availability than the present system.

However, a financial analysis showed that the present system consisting of third stage cyclones, and lower temperature metallic filters following the high temperature syngas cooler was more economic because of the high cost of the ceramic filter system. The high temperature ceramic filter systems should be reevaluated in the future to see if any cost reductions have been obtained that could make them more economic.

8.1 SUMMARY

Subtask 3.3 developed an alternate design for a combined heat and power plant for an industrial facility located in upstate New York. Table 8.1 summarizes the major input and output streams from this alternate design along with some key operating parameters and compares them to the original Subtask 3.2 base case.[1]

Table 8.1 Overall Plant Summary

	Alternate	Base	Difference
Design Inputs			
Coal Feed, moisture-free tpd	345.7	345.7	0
Coal Feed, moisture-free lb/hr	28,810	28,810	0
Fuel (Natural Gas), MMBtu/hr	-	5.1	-5.1
Makeup Water Input from the Industrial Facility			
Boiler Feed Water, gpm	418	495	-77
Quench Water, gpm	0	30	-30
Cooling Tower Makeup Water, gpm	58	53	+5
Design Outputs			
Export Power, MW	21.3	21.7	-0.4
Export Steam (400 psig, 550°F), Mlb/hr	105.34	101.72	+3.62
Sulfur, lb/hr	899	899	0
Ash, lb/hr	2,719	2,097	+622
Condensate (to industrial facility), Mlb/hr	54.4	60.9	-6.5
EPC Cost, MM$*	82.1	90.0	-7.9
Plant EPC Cost, $/kW**	2,755	3,090	-335
Plant Energy Input, k$/MMBtu/hr	209.7	229.9	-20.2
Plant Energy Output, k$/MMBtu/hr	421.6	469.2	-47.6
Equivalent Availability, %	84.7	85.7	-1.0
Return on Investment, %***	8.4	5.9	+2.5
Cold Gas Efficiency, % (HHV basis)	79.3	79.3	0
Net CHP Efficacy, % (HHV basis)	49.7	49.0	+0.7

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

** Based on converting the steam export to power using an average turbine efficiency

*** Based on 8.0 cents/kWh and 12 $/ton of steam

[1] "Topical Report – Subtask 3.2, Preliminary Design for Eastern Coal," Gasification Alternatives for Industrial Applications, United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, November 2004

Compared to the base case, the alternate case has about 2% less export power, 3.5% more export steam, about 9% less capital investment, a higher net CHP efficacy and a higher return on investment.

8.2 CONCLUSIONS

This study has shown that:

- Improvements that were made to this Subtask 3.3 Alternate Case increased the return on investment by 2.5 percentage points (5.9 to 8.4%) over that of the Subtask 3.2 Base Case.

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

- A ROI of 8.4% is achievable at the current market price of electricity in upstate New York. Future optimization of this plant design should identify several additional enhancements that will further improve the economics of IGCC power plants (see below for a list of potential enhancements and improvements).

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

- 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 5.4).

- 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. These include:

    - Improved sulfur removal methods including warm sulfur removal
    - Warm mercury removal systems
    - Improved particulate removal systems resulting in reduced capital costs and higher efficiency

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

    - Studying improved coal drying techniques since the coal preparation and drying account for nearly 10% of the total project cost
    - Investigating the effect that the coal moisture content has on the U-GAS® gasifier operation
    - Updating the database for gasification reactivity of the desired coal

- Characterizing the particulate properties
- Characterizing the hydrocarbon content of the syngas to confirm the sour water stripper design and effluent water treatment facilities
- Investigating cyclone performance at high temperatures (greater than 1000°F)
- Determining the combustion turbine performance capabilities for the desired engine(s) (both output and emissions)
- Determining the characteristics of the ash associated with the char

Another objective which was realized 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 above-described 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 and industrial facilities. The following areas are recommended for further development through additional systems analysis and/or R&D efforts:

- Additional optimization work is required for coal. These include further optimization of the plant configuration, such as with the heat integration and/or sulfur recovery. One example is integration of the gas turbine and ASU, which could reduce compression costs. This change may significantly reduce the cost and improve the efficiency of the gasification plant. A commercial demonstration of this type of integration would be valuable to all gasification systems.

- 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 made more thermally efficient.

- Develop a R&D program that will address critical issues such as
    - Improving the availability of the gasification system and various sub-systems
    - Determining combustion turbine performance (both power output and emissions) on syngas in order to prepare for widespread commercialization of gasification

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

- The physical characteristics and properties of coal must be studied further in order to better predict gasification system performance. These include:
    - Determination of the gasification reactivity of the desired feedstock.
    - Determination of the ash characteristics associated with the char
    - Characterization of the particulate properties
    - Characterization of the hydrocarbon content of the syngas to confirm that the design of the sour water stripper and effluent water treatment facilities are sufficient to handle tars and oils
- Determination of cyclone performance at higher temperatures (above 1000°F).
    - 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 domestic company that is a world leader in cyclone design).

A.1 ASPEN

A.1.1 Gasification Island

The modeling of the gasification island is identical to that for Subtask 3.2 except that for this subtask there is only one gasifier train in operation (vs. two gasifier trains). For details the reader is referred to the Topical Report for Subtask 3.2.[1]

A.1.2 Syngas Cleanup System

The syngas cleanup systems for this subtask were modeled in the same manner as was used for Subtask 3.2, only the flow scheme has changed slightly. The changes are noted below:

- Metallic candle filters are used to remove particulates upstream of the scrubber.
  - This allows the use of a venturi scrubber in place of the impingement scrubber employed for Subtask 3.2. The venturi scrubber reduces the water requirements by about 35%.
  - It also allows the temperature of the syngas leaving the fired tube boiler to be lowered to 480°F (vs. 600°F in Subtask 3.2). The additional high temperature heat recovery is used to generate additional 400 psig steam.
- LO-CAT® technology is used in place of the Claus and SCOT tail gas cleanup units for sulfur recovery.

ASPEN Plus provides a number of physical property methods for calculation of stream thermodynamic parameters under various conditions; different property methods will yield different results, and sometimes these results can have significant repercussions on the entire design. For our current system, caution needs to be exercised in evaluating the syngas water scrubber and the flash drum downstream of the low temperature heat recovery system, since gases are dissolved in the sour water and process condensate, both of which are treated in the sour water stripper. It is important to realistically estimate the gas content 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 PR-BM physical property method set. However, for applications involving electrolytes, such as the 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 make sure that the gases in the sour water and process condensate are correctly predicted, the

[1] "Topical Report – Subtask 3.2, Preliminary Design for Eastern Coal," Gasification Alternatives for Industrial Applications, United Stated Department of Energy, National Energy Technology Laboratory, Contract No. DE-AC26-99FT40342, November

ASPEN Plus simulation developed for the current design incorporates the results obtained using both the ElectrolyteNRTL and PR-BM equations of state.

A.1.3 Sour Water Stripper

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

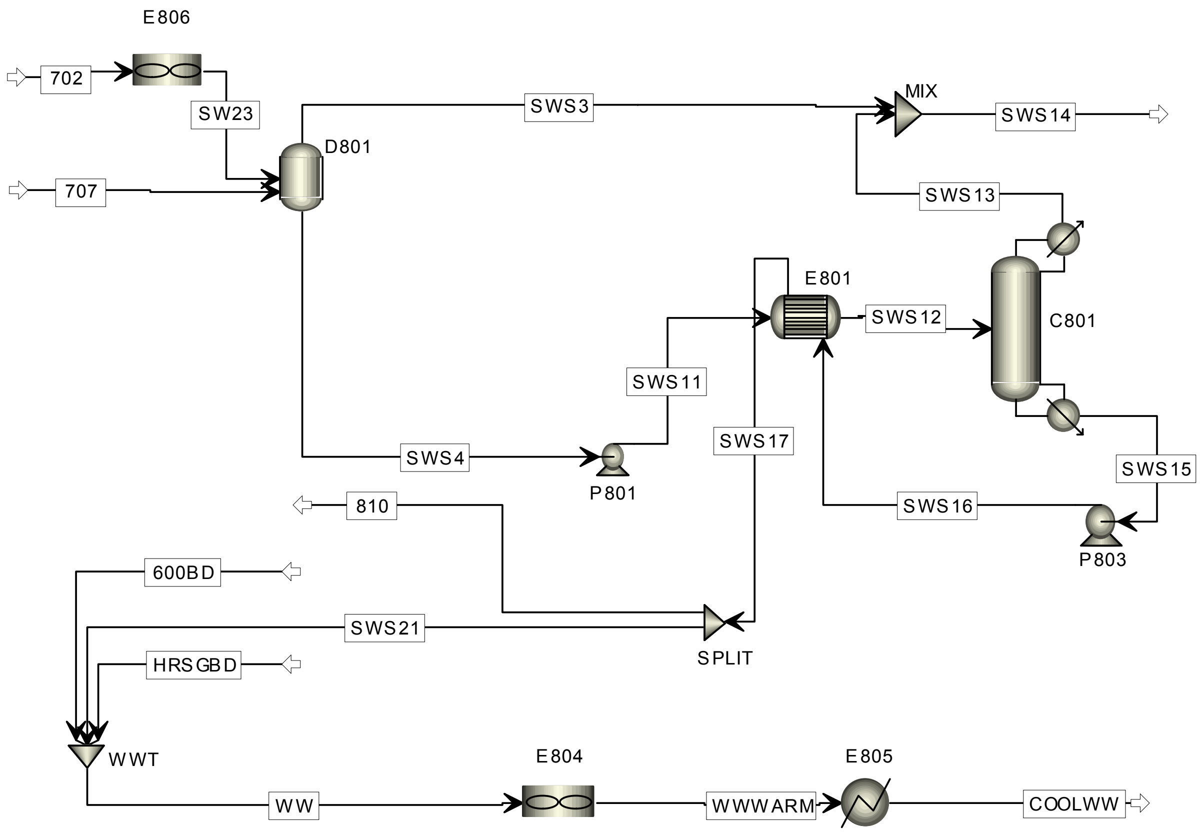
The largest sour water feed stream comes from the venturi scrubber down stream of the high temperature heat recovery (HTHR) unit. A portion of the process condensate also is mixed with the scrubber water and treated in the sour water treatment system. In addition to the dissolved impurities to be removed 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 (primarily benzene and naphthalene derivatives) also may be in the sour water.

A.1.3.3 Sour Water Stripper

Figure A.1 shows the ASPEN flow diagram for the sour water treatment (SWS) unit. The SWS processes the effluent from the venturi scrubber and the process condensate from the flash drum upstream of the amine system. It consists of a three-phase flash drum, sour water stripping column, and associated heat exchangers and pumps. In addition a 24 hour sour water storage tank is provided as a back-up in case the sour water stripper column is unavailable. Vapors from the flash drum and stripping column are sent to the LO-CAT® unit. Stripped water from the bottom of the column is recycled to the venturi scrubber. Excess water not required by the venturi scrubber along with the blowdown streams from the HTHR and HRSG are cooled and sent to the waste water treatment plant.

The distillation column was designed using the same techniques and bottoms specifications (i.e., no more than 50 ppmw ammonia and less than 10 ppmw hydrogen sulfide) that were used for Subtask 3.2.

Figure A.1 ASPEN Flow Diagram - Sour Water Stripper



A.2 GATECYCLE

A.2.1 Power Block

The power block consists of the combustion turbine set (CT and generator), and heat recovery steam generator (HRSG). The requirements of the facility call for two identical parallel CT/HRSG trains. The basis for the CT/HRSG is described in Appendix E of this report. The power block was modeled using the GateCycle computer program for Windows Version 5.52.0.r by simulating one of the two individual parallel trains. Syngas composition was generated separately using ASPEN Plus and is described in other sections of this report.

A.2.2 Combustion Turbine Modeling

The GE 10 turbine (11.25 MW ISO conditions, natural gas DLE) was selected for this facility. For the purpose of modeling, a Nuovo Pignone PGT10B turbine (a forerunner to the GE 10) was selected from the GateCycle turbine library. Syngas composition was generated using ASPEN Plus and was the basis for fuel input into GateCycle. The specific fuel inputs were calculated using the Excel spreadsheet fuelcalc.xls that accompanies the GateCycle software.

As noted in Subtask 3.2, there is some degree of uncertainty when modeling coal-derived syngas (or any low Btu syngas) with the stock turbines provided in the GE software turbine library. Because the turbines in the library are based on existing performance data, modeling a turbine with fuel gas of a significantly different composition than that on which the data are based may result in model predictions that vary from acutal performance. GateCycle also allows for the use of a modeling block called data gas turbine. This option allows for the specification of turbine performance including the use of gas turbine curve sets. Because the GE 10 turbine has not been commercially demonstrated for use on coal-derived syngas, there is not sufficient data available for use. Use of the GE software library PGT10B data provided results reasonably consistent with GE performance data mentioned above.

A.2.3 HRSG Modeling

The individual HRSGs were modeled such that four specific process conditions were met:

- Stack temperature remained above the acid-dew point so that condensation and corrosion did not occur within the system (~240°F).
- Sufficient 50 psig superheated steam (~353°F) was generated such that the process steam demands of all gasifier and gas clean-up processes were satisfied (including gas clean up operations and sour water treatment).
- Balance of steam generation was 400 psig superheated steam (~548°F).
- Take advantage of the excess low quality heat to preheat the syngas entering the gas turbine.

The modeling was accomplished by inserting the appropriate HRSG components downstream of the turbine exhaust. Figure A.2 provides a screen capture of the GateCycle flow diagram.

Figure A.2 Gate Cycle Process Flow Diagram

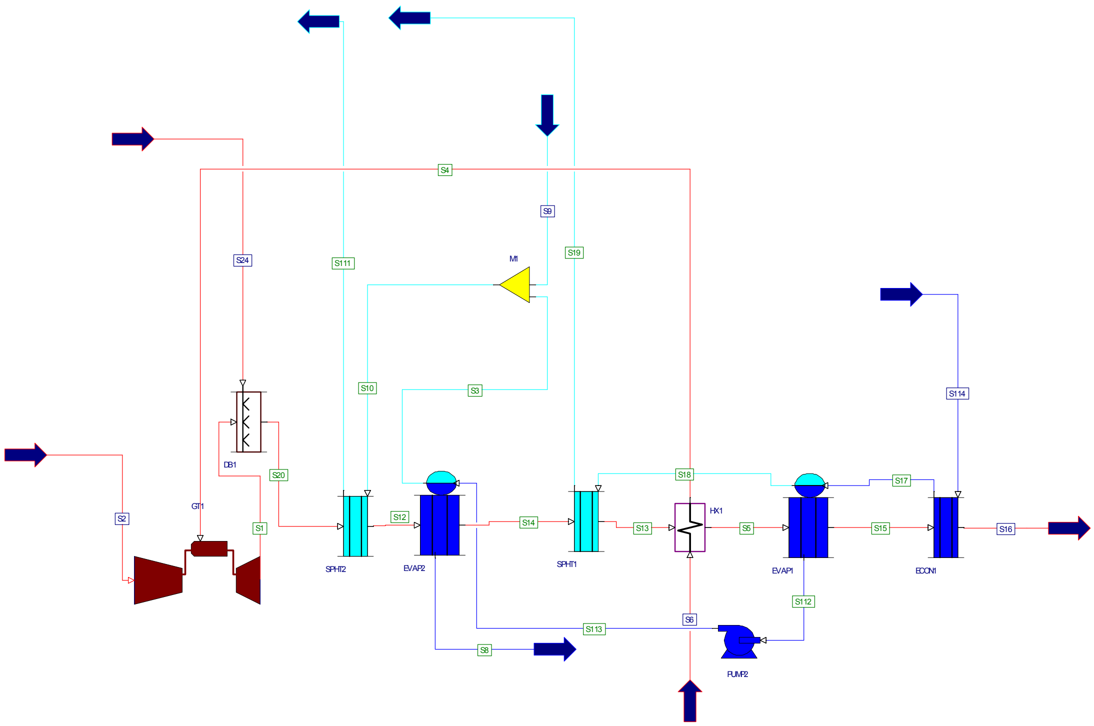


Figure A.2 shows the HRSG was modeled to produce both 400 psig and 50 psig superheated steam to meet the requirements described above. The key input parameters included degree of superheat for 50 psig and 400 psig steam, 400 psig steam input (from gasification operations), and pinch ΔT of the 400 psig evaporator and 50 psig evaporator. To achieve the necessary 50 psig steam production, the pinch ΔT for the 400 psig evaporator was manually adjusted such that the desired steam production was achieved. The only difference between this design and that for Subtask 3.2 is an addition of syngas preheater. This heater preheats the syngas entering the gas turbine to about 300°F. A detailed description of the model can be found in the topical report of Task 3.2.

B.1 COAL PREPARATION

The equipment included in the coal preparation area (Area 100) of the Subtask 3.3 Alternate Case includes:

- Car Shaker - The car shaker is manufactured by Kinergy with a pneumatic operator, 15 hp motor. The shaker covers half the length of the rail car.
- Thawing Equipment - The railcar thawing equipment consists of infra-red heaters that require 200 kW of electric power to heat the contents of the car to 500°F and keep the air space around the car at 300°F during thawing. The radiant heaters are mounted on the side walls of the building and on the middle of the rail track.
- Under Track Feeder - Kinergy Screw Feeder, 300 tph capacity, 10 hp motor, horizontal type
- Belt Conveyor to Transfer Coal to Active Pile - Incline belt conveyor, 300 tph capacity, trough type, 36 inch wide x 220 feet long, with head end metal detector and magnet, 40 hp motor
- Active Pile Discharger - Kinergy Discharger, 15 hp motor, design for 40 tph capacity.
- Reclaim Screw Feeder - Kinergy Screw Feeder, 40 tph capacity, 5 hp motor, horizontal type
- Belt Conveyor - Coal Feed to Crusher - Incline belt conveyor, 40 tph capacity, trough type, 30 inch wide x 120 feet long, 20 hp motor
- Coal Crusher - Williams two stage, heavy duty, single and double roll crushers, 40 tph capacity, 50 hp motor, to reduce 2" x 0" coal to ¼" top size.
- Crushed Coal Feeder – Transfer Coal to Elevator, 40 tph capacity
- Kinergy Fluidized Bed Dryer - 40 tph capacity, with steam coil heater, FD and ID fans, dryer, cyclone separator, dust collector, ductwork, motors with 15, 40 and 50 hp.
- Kinergy Vibratory Screen - 40 tph capacity, 10 hp motor, separating dried coal sized above 120 mesh to ¼" for delivery to primary coal silo. Larger coal is recirculated to coal crusher coal finer than 120 mesh is collected and pneumatically transported by a pressure blower (with a 40 hp motor) to the other plant boilers and used as fuel.

- Vibratory Screen Discharge Screw Feeder, 40 tph capacity, 5 hp motor, horizontal type
- Continental Bucket Elevator - 40 tph capacity
- Primary Coal Silo - 363 Tons for 24 hour storage, 21 ft. diameter x 42 ft. cylinder height, with bottom hopper and discharge gate, top dust collector and vent, with 15 hp exhaust fan motor.
- Continental Screw Conveyor - 32 tph, 10 hp motor
- Continental Bucket Elevator - 32 tph capacity
- Redundant Primary Silo Discharge Screw Conveyor - 32 tph, 10 hp motor
- Redundant Continental Bucket Elevator - 32 tph capacity
- Coke Truck Receiving Hopper
- Continental Screw Conveyor for coke transport - 16” x 20”; discharge screw conveyor, 7.5 hp motor
- Continental Bucket Elevator for coke transfer
- Start-Up Coke Silo - 8 hour storage, 14 ft. diameter cylinder x 32 ft. height, with bottom hopper and discharge gate, top dust collector and vent, with 15 hp exhaust fan motor.
- Continental Screw Conveyor for coke transport to Redundant Elevator - 5 tph Capacity, 16” screw conveyor, 10 hp motor
- Surge Hopper
- Distribution Screw Feeder (with grab sample connection, and four drop off openings with motor operated knife gates) - 5 to 32 tph Capacity, 15 hp motor
- Fines Collector - 15 hp exhaust fan motor and 40 hp pressure blower motor.

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

B.2 SUBTASK 3.3 ALTERNATE CASE

The equipment in Areas 150 though 1000 for the Subtask 3.3 Alternate Case includes:

Area 150 Air Supply				
Identification	No.	Description	Comments	Unit Size
K-151	2	Air Compressor package	850 GPM cooling water W/ 25F delta	4,000 bhp
E-151 A/B	2	Heat Exchanger	SHELL: DP= 450 psig; DT= 600 F; TUBE: DP= 450 psig;DT= 600 F; CS Tubes , AEU	510 sq ft
Area 200 Coal Feeding				
Identification	No.	Description	Comments	Unit Size
T-201	2	Weigh Hopper		
S-201	2	Rotary Feeder		
S-202	2	Rock Pump/Live-wall hopper/controls	$100k for eng and test and $200k/pump for installation	
Area 300 Gasification				
Identification	No.	Description	Comments	Unit Size
R-301	1	Gasifier		
		Refractory		
		Internals		
H-301	1	Startup Heater		

Area 400 Dust Removal				
Identification	No.	Description	Comments	Unit Size
CY-401	1	Primary Cyclone		
		Refractory		
CY-402	1	Secondary Cyclone		
		Refractory		
CY-403	1	Tertiary Cyclone		
		Refractory		
		Connecting Refractory Pipe		
D-401	1	Cyclone Surge Hopper		
		Refractory		
Area 500 Ash Removal				
Identification	No.	Description	Comments	Unit Size
D-501	1	Ash Surge Hopper		
		Refractory		
S-501	1	Ash Transport Screw		
D-502	1	Ash Lock Hopper		
T-503	1	Ash Pneumatic Transport Hopper		
S-503	1	Ash Feeder		
T-504	1	Ash Storage Silo		
Areas 600/650 Gas Cooling & Filtration				
Identification	No.	Description	Comments	Unit Size
Fil-651	1	Gas Solids Separation System	quote from Pall Corp. (650 F spec)	72" ID vessel containing 162 filter elements; DP=500psig; DT=700^{o}F
P-651	1	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-601	1	High Pressure Steam Drum	CS Horizontal, P_{des} =470psig T_{des} = 500F	D = 11 ft, and L = 44 ft
E-601	1	High Pressure Steam Boiler	SHELL: DP= 470 psig; DT= 505 F; TUBE: DP= 375 psig;DT= 1800 F; Inconel Tubes , BEM	1303 sq ft
E-602	1	BFW Preheater	SHELL: DP= 365 psig; DT= 530 F; TUBE: DP= 480 psig;DT= 375 F; 410 SS, AEU	595 sq ft
P-601	1	HP Steam Boiler Start-up Pump		flowrate: 250 gpm; head = 100 ft, 9.9hp
P-602 A/B	2	Fresh Quench Water pump		47 gpm; delta P = 405 psi; head =994 ft
P-603 A/B	2	BFW Booster Pump		132 gpm; delta P = 360 psi; head = 871 ft

Area 700 Gas Cleaning				
Identification	No.	Description	Comments	Unit Size
C-701	2	Syngas Venturi Scrubber Column	Internals = 410SS, Vessel = CS, P_{des} = 355psig, T_{des}=300F	
E-701	1	BFW Preheater	SHELL: DP= 355 psig; DT= 300 F; TUBE: DP= 85 psig;DT= 250 F; CS Tubes , AEU	2062 sq ft
E-702	1	Effluent air cooler	air fin; DP = 340; DT = 260 F	5176 sq ft
E-703	1	Effluent trim water cooler	SHELL: DP= 340 psig; DT= 190 F; TUBE: DP= 100 psig;DT= 160 F; CS Tubes , AEU	810 sq ft
E-704	1	COS Hydrolysis Reactor Preheater	SHELL: DP= 450 psig; DT= 600 F; TUBE: DP= 350 psig;DT= 325 F; CS Tubes , AEU	88 sq ft
R-701	1	COS Hydrolysis Reactor	Vessel size supplied by Sud Chemie	9.0' ID by 12.0' TT, 11.1' bed depth, 706.1 cu ft of Sud Chemie C53-2-01 1/8" catalyst, DP=350psig, DT=325F, CS
D-701	1	Effluent condensate drum	vertical drum, DP=330 psig, DT=160F, CS	Dia = 6, H = 6.5 ft,
R-711	1	Mercury Adsorption Vessel	Information supplied by Calgon Carbon	9.0 ft ID by 10.0 ft TT
S-701	Lot	Sud Chemie C53-2-01 1/8" catalyst	Loading supplied by Sud Chemie	
S-711	Lot	Sulfur Impregnated Activated Carbon	Information supplied by Calgon Carbon	20,000 Lb

Area 800 Acid Gas Removal and Sulfur Recovery				
Identification	No.	Description	Comments	Unit Size
C-801	1	Sour Water Stripper Column	CS w/ 410SS internals; DP = 50 psig; DT = 320 F, 21 trays	Dia = 2.5 ft; T-T = 60 ft
D-801	1	Sour Water Three Phase Settler	horizontal drum; DP = 45 psig; DT = 295 F; CS clad w/410 SS	Dia = 6.5 ft; L = 19.5 ft
D-802	1	SWS Stripper Distillate Drum	horizontal drum; DP = 50 psig; DT = 215 F, CS clad w/ 410SS	Dia = 3.5 ft; L = 11 ft
E-801	1	SWS Feed Pre-Heater	S&T; AEU; DPshell = 85; DPtube = 80; DTshell = 320; DTtube = 305; 410 SS tubes; CS shell	Area = 1513sq ft
E-802	1	SWS Condensor - Air Fin	air fin; DP = 50; DT = 265 F	Area = 7400 sq ft
E-803	1	SWS Kettle Reboiler	S&T; BKU; DPshell = 50 ;DPtube = 150 ;DTshell = 320 ;DTtube = 350 ; CS tubes; CS shell	Area = 1849 sq ft
E-804	1	Waste water cooler (air fin)	air fin; DP = 75; DT = 350 F	Area = 996 sq ft
E-805	1	Waste water trim cooler (cooling water)	S&T; AEU; DPshell = 70 ;DPtube = 100 ;DTshell = 190 ;DTtube = 160 ; CS tubes; CS shell	Area = 94 sq ft
E-806	1	Scrubber SWS air cooler	air fin; DP = 350; DT = 300 F	Area = 1008 sq ft
P-801 A/B	2	SWS Feed Pump		58 gpm; delta P = 53 psi; head = 127 ft
P-802 A/B	2	SWS Reflux Pump		26 gpm; delta P = 42 psi; head = 104 ft
P-803 A/B	2	Stripper Bottom Pump		59 gpm; delta P = 30 psi; head = 74 ft
TK-801	1	Sour Water Storage (Day Tank)	Dia = 25 ft, height = 24 ft	85,000 gallons
S-801	1	Initial Fill of Amine Solution	Information supplied by Dow Chemical Co.	10,000 gal
S-802	1	Initial Fill of Activated Carbon		36.3 cu ft
	1	LO-CAT® Unit	quote from GTP	
Area 900 Gas Turbine and HRSG				
Identification	No.	Description	Comments	Unit Size
GT-901	2	Syngas Turbine		
F-901	2	Final Syngas Filter		
	2	HRSG	quote from Vogt Power	80,615.6 sq ft
E-901	2	Syngas Preheater	2 shells in series, DPshell = 300; DPtube = 450; DTshell = 370; DTtube = 470; CS	615 sq ft (total)

Area 1000 Offsites and Auxiliaries				
Identification	No.	Description	Comments	Unit Size
		Steam generation system	all steam generated on-site	
		Condensate collection system	**equipment:** stroage tank, booster pumps	109 gpm
		Demineralized water system	not reqwuired	
		Cooling water system	supply temp = 80^{o}F, return temp = 100^{o}F, **equipment:** cooling tower, circulation pumps	2,922 gpm normal; 3,800 gpm design
		Safety shower/eye wash system		
		Raw water/fire water system	**equipment:** raw water/fire water storage tank, CW makeup pump, fire water pumps	250k gal storage
		Drinking (potable) water system		
		Compressed air system	**equipment:** 2 compressors (2 working, backup from industrial site), desiccant air dryers, IA receiver tank, PA receiver tank	600 SCFM (each)
		Natural gas supply system		
		Flare system	**equipment:** elevated flare, pilot and knock out drum	280 million Btu/hr
		Nitrogen system	package	14.2 thousand SCFH
		Waste water collection, treatment and disposal system	**equipment:** sumps and sump pumps, transfer pumps	
		Electrical distribution system		
		Interconnecting piping		
		Telecommunications systems		
		Buildings		
		Miscellaneous		

Figure C.1 Simplified Flow Diagram – Coal Handling System

Figure C.2 Gasification – IGCC Process Flow Sheet

Figure C.3 Heat Recovery and Gas Clean Up Process Flow Sheet

Figure C.4 Mercury and Acid Gas Removal Process Flow Sheet

Figure C.5 Gas Turbine & Gas Recovery Steam Generation Process Flow Sheet

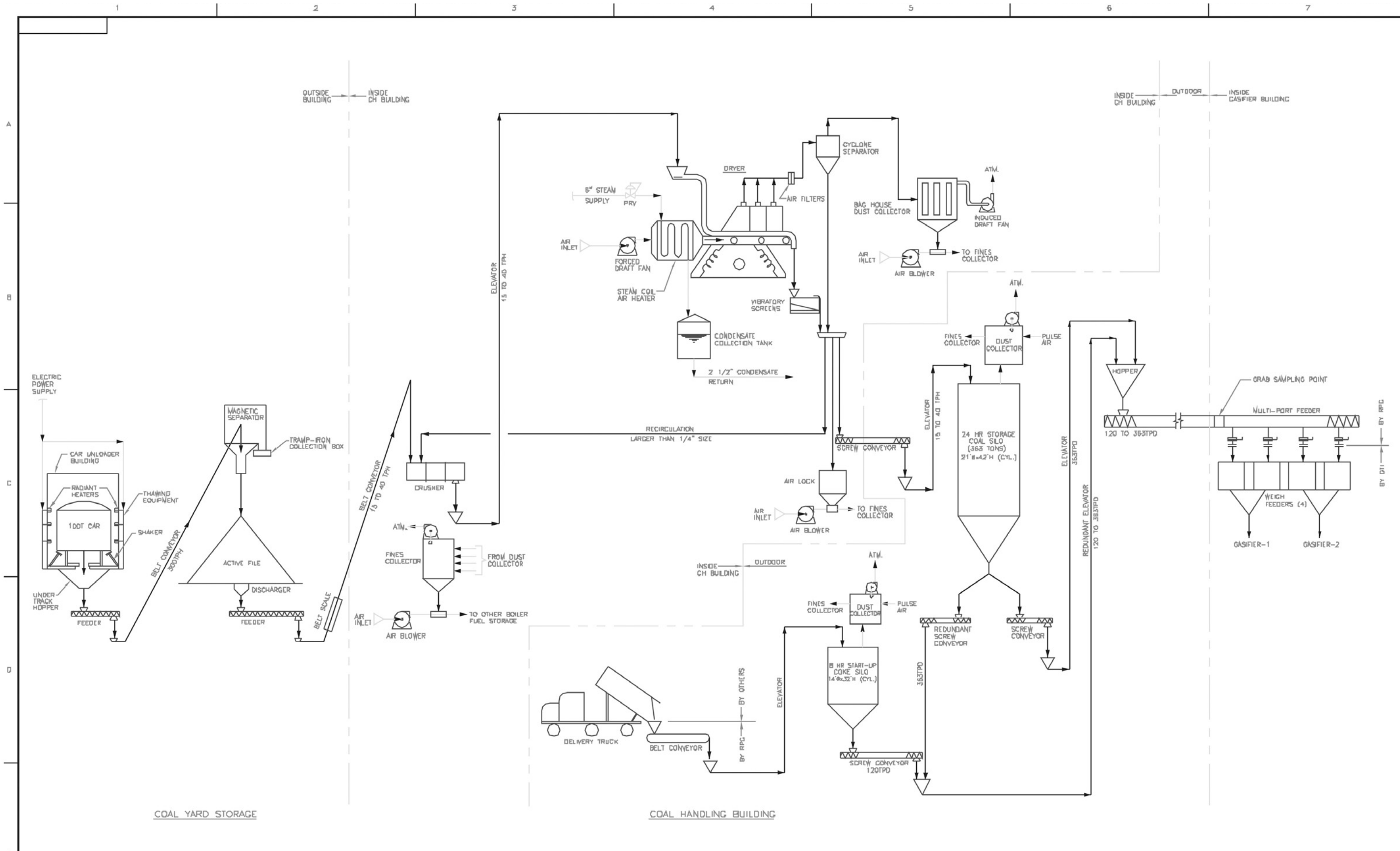
Table C.1 Gasifier Island Material and Energy Balance

Table C.2a/b Gas Cooling & Cleaning Material and Energy Balance

Table C.3 Sour Water Stripper Material and energy Balance

Table C.4 GT/HRSG Material and Energy Balance

Figure C.1 Simplified Flow Diagram – Coal Handling System



REV.	DATE	REVISION DESCRIPTION	DWN.	CHKD.	APPD.
A	5/11/04	REVISED PER 5/11/04 MEETING	VB	TO	
B	6/25/04	REVISED PER 8/25/04 CONFERENCE CALL	VB	TO	
C	7/01/04	REVISED PER NEXANT COMMENT	VB	TO	

DRAWN BY: V.R.BALUYOT
DESIGNED BY: T. ONANA
CHECKED BY: T. ONANA
APPROVED BY:
DRAWING STATUS: PRELIMINARY FINAL

NEXANT, INC.

SIMPLIFIED COAL HANDLING SYSTEM FLOW DIAGRAM

RAYMOND Raymond Professional Group, Inc.
Raymond Professional Group - A/E, Inc.
550 West Van Buren Street, Suite 400
Chicago, Illinois 60607
Tel: (312) 935-3200 Fax: (312) 935-3201

SCALE: NONE
FILE NAME: COALSCHEMATIC
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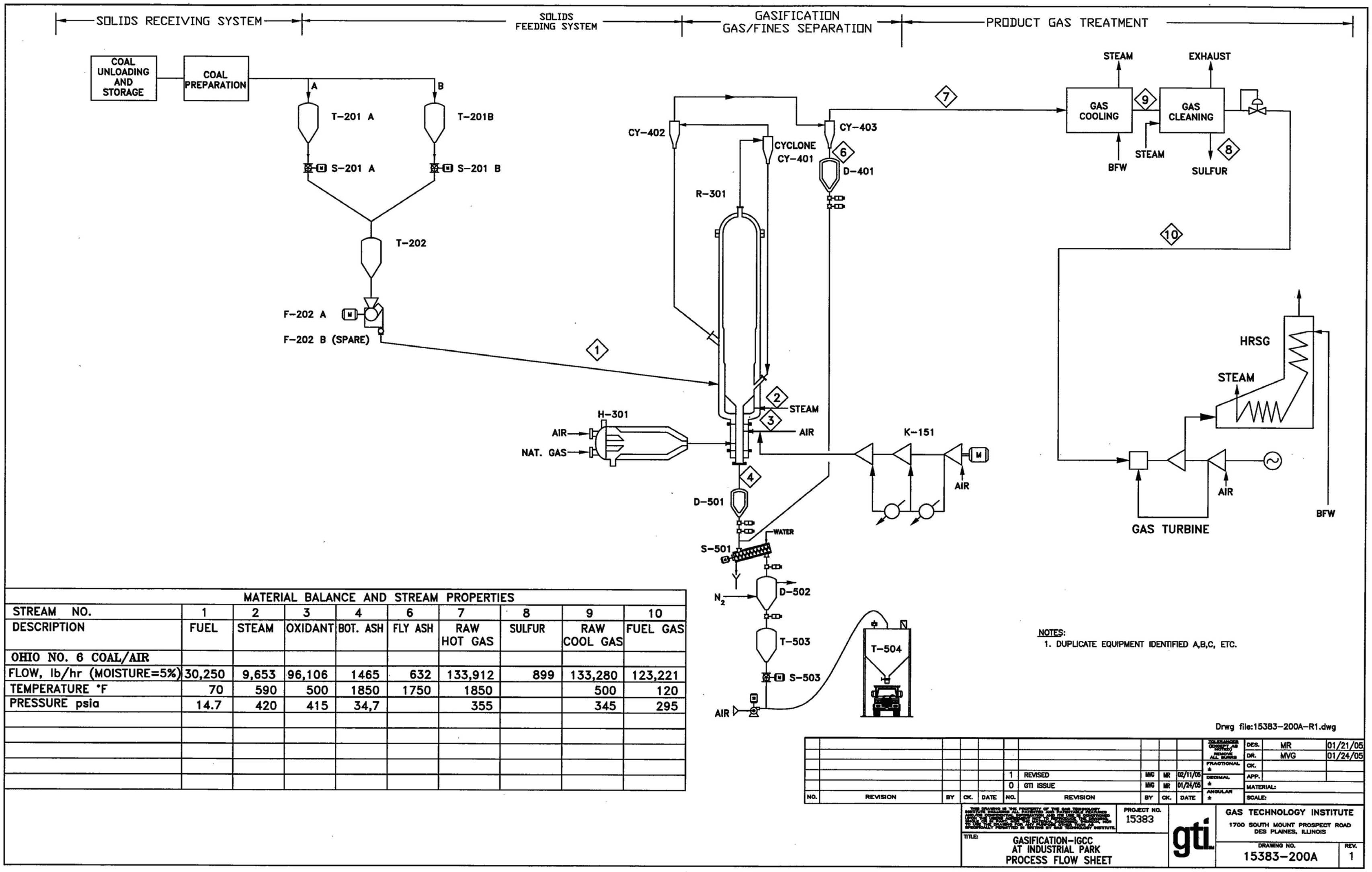
PDF FILE:
MICRO-STATION CELL LIBRARY:
DATE: 9/03/04
REV. C
PROJECT NUMBER 176678
DRAWING NO.
SHEET OF

Figure C.2 Gasification – IGCC Process Flow Sheet



MATERIAL BALANCE AND STREAM PROPERTIES

STREAM NO.	1	2	3	4	6	7	8	9	10
DESCRIPTION	FUEL	STEAM	OXIDANT	BOT. ASH	FLY ASH	RAW HOT GAS	SULFUR	RAW COOL GAS	FUEL GAS
OHIO NO. 6 COAL/AIR									
FLOW, lb/hr (MOISTURE=5%)	30,250	9,653	96,106	1465	632	133,912	899	133,280	123,221
TEMPERATURE °F	70	590	500	1850	1750	1850		500	120
PRESSURE psia	14.7	420	415	34,7		355		345	295

Figure C.3 Heat Recovery and Gas Clean Up Process Flow Sheet

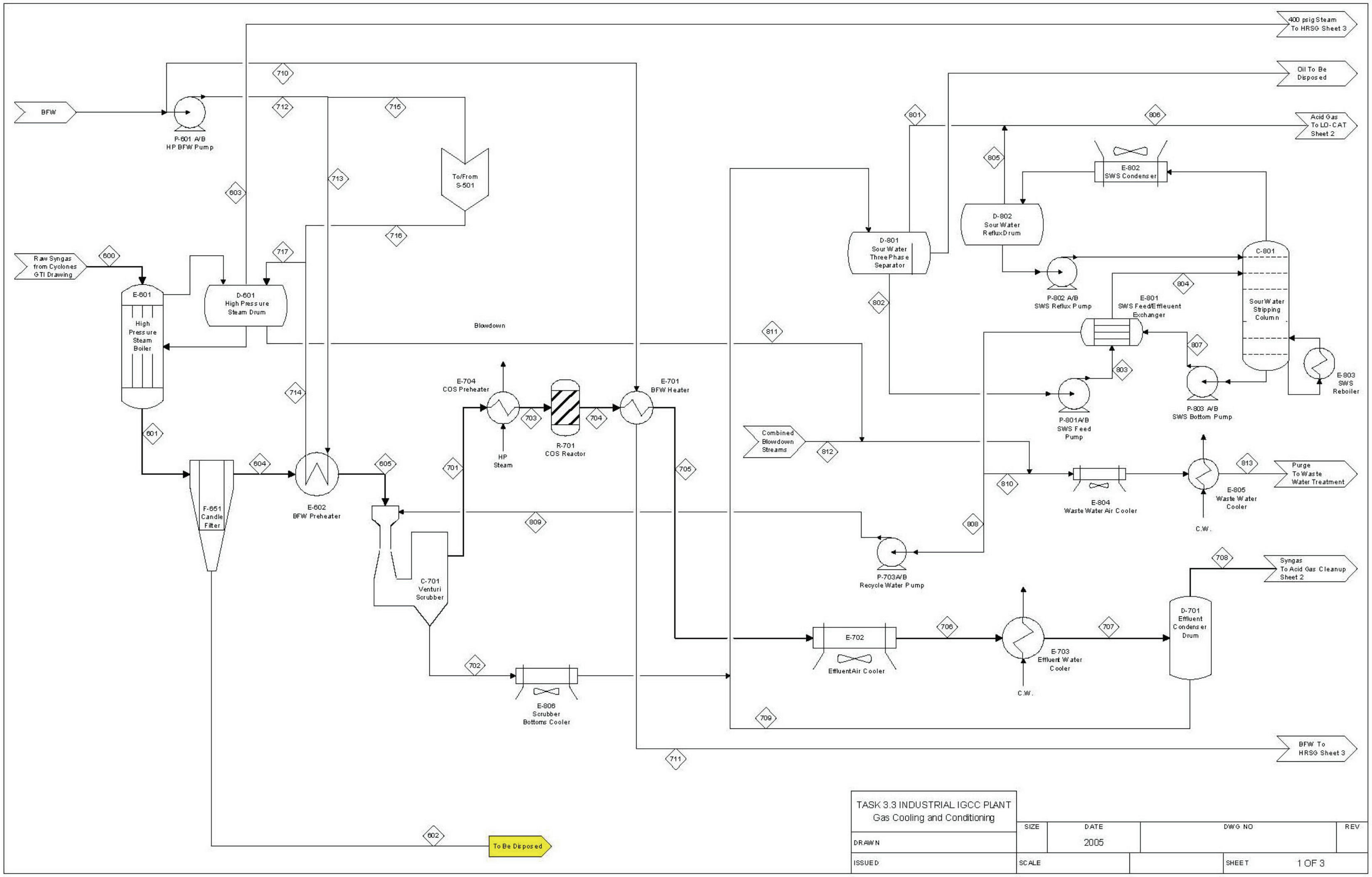


Figure C.4 Mercury and Acid Gas Removal Process Flow Sheet

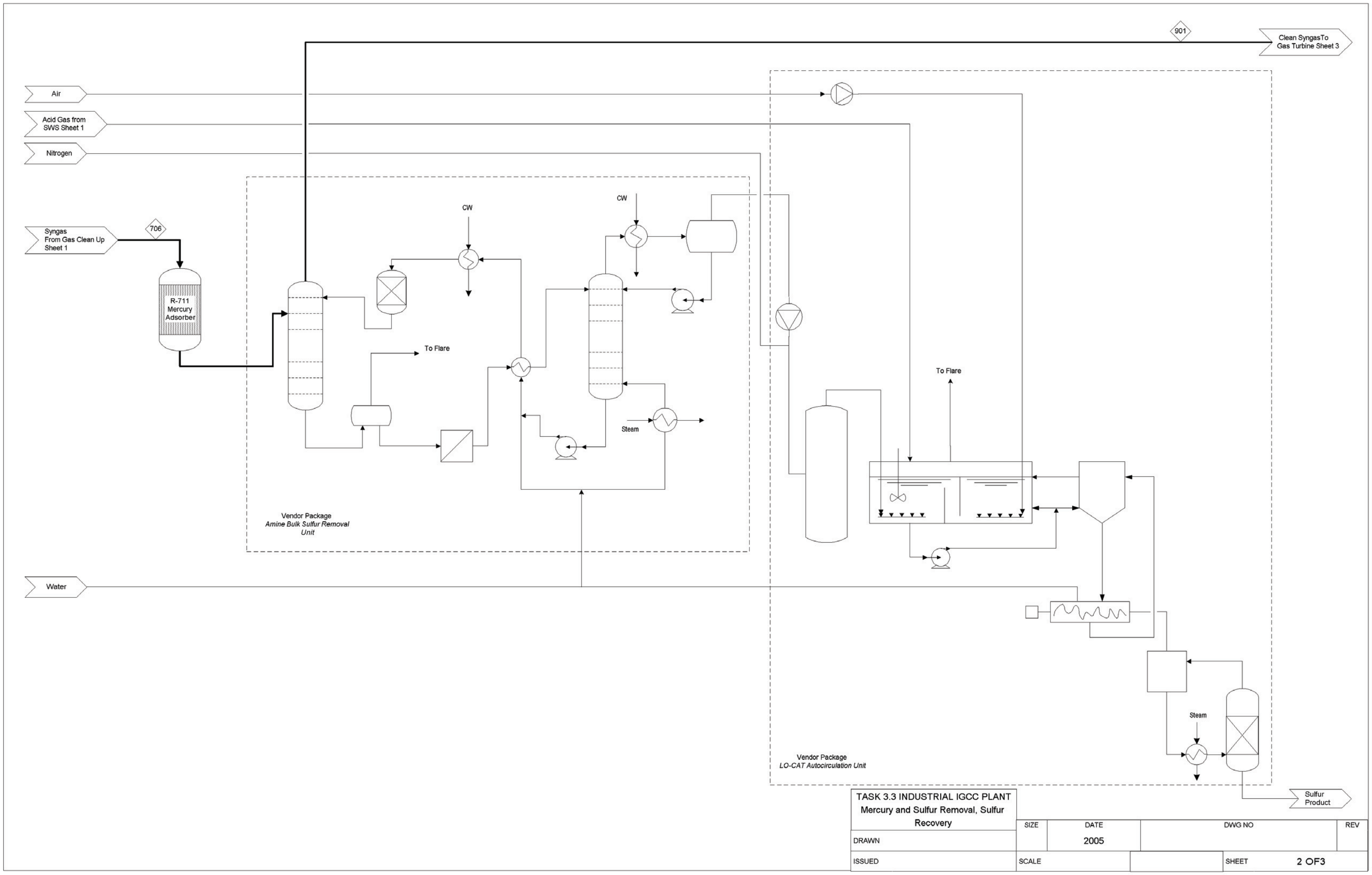


Figure C.5 Gas Turbine & Gas Recovery Steam Generation Process Flow Sheet

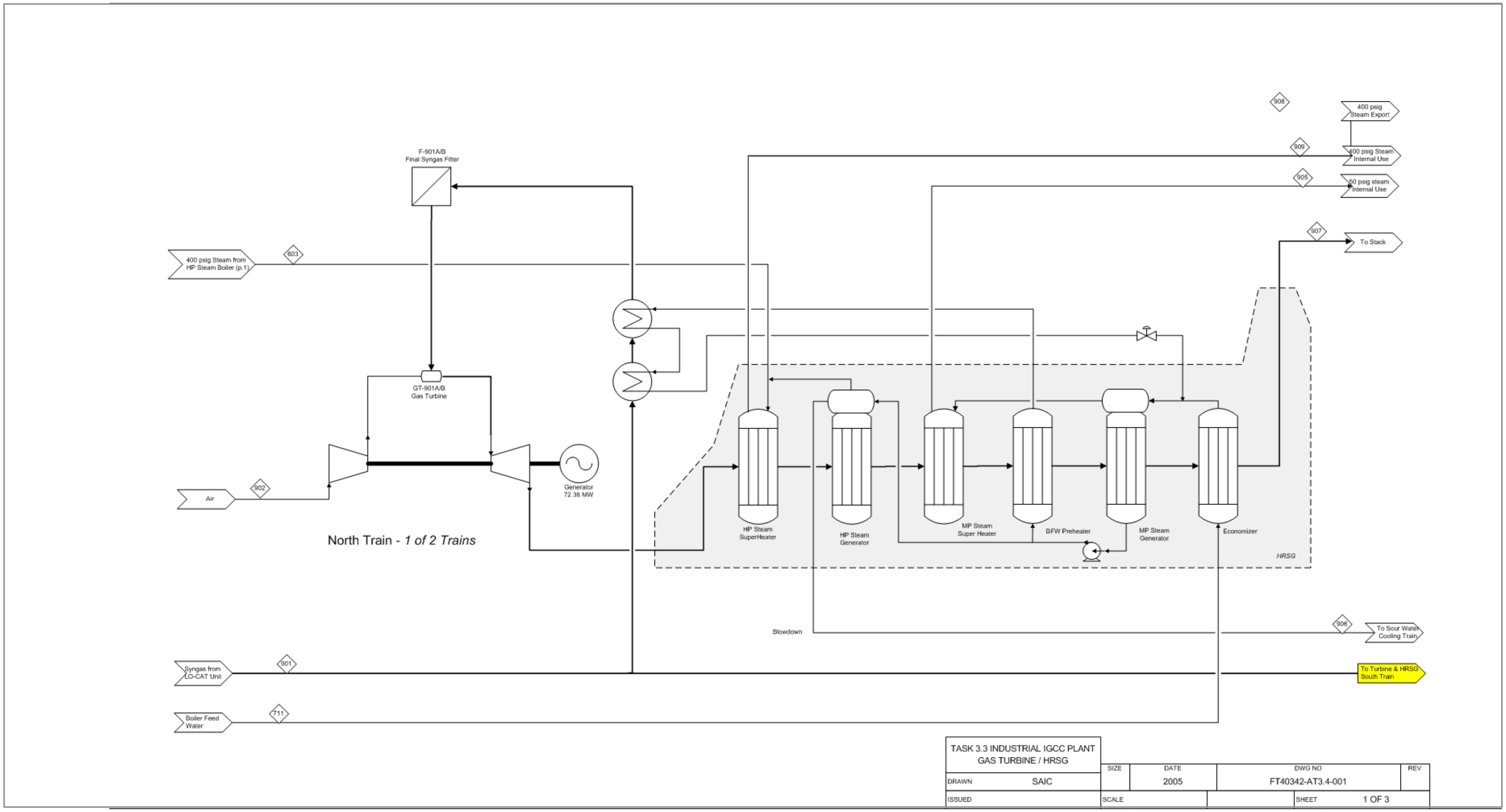


Table C.1Gasifier Island Material and Energy Balance

Stream No.	1	2	3	4	6
Stream Description	**Coal**	**Steam**	**Oxidant**	**Bottom Ash**	**Fly Ash**
Stream Composition, lb/h					
CO					
CO_2					
H_2					
H_2O	1,513	9,653	906		
CH_4					
H_2S					
COS					
NH_3					
HCN					
N_2			73,335		
O_2			21,865		
Coal/residue [1]	27,039			552	239
Mineral Matter/Ash	1,698			913	393
Total, lb/h	**30,250**	**9,653**	**96,106**	**1,465**	**632**
Temperature, F	70	550	500	1850	1750
Pressure, psia	14.7	420	415	14.7	14.7

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

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

Stream Number	600	601	602	603	604	605
Stream Description	Raw Syngas to tube fired boiler	Raw Syngas to Candle Filters	Ash from Filters	Saturated HP Steam	Raw Syngas from candle filters	Raw Syngas to Venturi Scrubber
Temperature, oF	1,750	480	480	451	480	351
Pressure, psia	355	345	345	430	340	335
Vapor Frac	1.00	1.00		1.00	1.00	1.00
Mole Flow, lbmol/hr	5,366	5,366		3,367	5,366	5,366
Mass Flow, lb/hr	133,280	133,280	632	60,651	133,280	133,280
Volume Flow, cuft/hr	360,368	157,616		67,461	159,934	139,626
Enthalpy, MMBtu/hr	-85.220	-142.690		-341.803	-142.690	-148.005
Density, lb/cuft	0.37	0.85		0.90	0.83	0.96
Mass Flow, lb/hr						
CO	31,158	31,158	0	0	31,158	31,158
CO2	16,221	16,221	0	0	16,221	16,221
H2	1,309	1,309	0	0	1,309	1,309
H2O	6,183	6,183	0	60,651	6,183	6,183
CH4	3,650	3,650	0	0	3,650	3,650
H2S	943	943	0	0	943	943
COS	42	42	0	0	42	42
H3N	135	135	0	0	135	135
CHN	17	17	0	0	17	17
N2	73,623	73,623	0	0	73,623	73,623
O2	0	0	0	0	0	0
Mole Flow, lbmol/hr						
CO	1,112	1,112	0	0	1,112	1,112
CO2	369	369	0	0	369	369
H2	649	649	0	0	649	649
H2O	343	343	0	3,367	343	343
CH4	228	228	0	0	228	228
H2S	28	28	0	0	28	28
COS	1	1	0	0	1	1
H3N	8	8	0	0	8	8
CHN	1	1	0	0	1	1
N2	2,628	2,628	0	0	2,628	2,628
O2	0	0	0	0	0	0
Ash, lb/hr	632	632	632	0	0	0

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

Stream Number	701	702	703	704	705	706	707	708	709	710	711	712	713	714	715	716	717	718	901
Stream Description	Syngas from Venturi Scrubber	Water from Venturi Scrubber	Syngas to COS Hydrolysis	Syngas from COS Hydrolysis	Syngas to Air Cooler	Syngas to Water Trim Cooler	Syngas to Separtor	Syngas to Amine Unit	Process Condensate to SWS	BFW to Preheater	Warm BFW to HRSG	BFW M/U @ HP	BFW to E-602	BFW from E-602	BFW to S-501	BFW from S-501	Warm BFW	Water to Venturi Scrubber	Syngas to HRSG
Temperature, °F	250	250	275	275	215	140	110	110	110	150	204	151	151	322	151	350	338	160	120
Pressure, psia	330	330	320	310	305	300	295	295	295	75	70	450	450	440	450	450	440	30	285
Vapor Frac	1.00	0.00	1.00	1.00	0.95	0.91	0.91	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00
Mole Flow, lbmol/hr	5,510	1,092	5,510	5,510	5,510	5,510	5,510	5,032	478	6,327	6,327	3,497	1,554	1,554	1,943	1,943	3,497	1,236	4,937
Mass Flow, lb/hr	135,845	19,697	135,845	135,845	135,845	135,845	135,845	127,091	8,755	113,982	113,982	63,000	28,000	28,000	35,000	35,000	63,000	22,263	123,221
Volume Flow, cuft/hr	126,857	313	135,514	142,712	115,247	99,823	95,623	4,296	145	1,915	1,979	1,059	471	528	588	674	1,202	388	106,963
Enthalpy, MMBtu/hr	-170.770	-126.774	-169.744	-170.168	-176.860	-184.130	-186.030	-123.750	-62.286	-773.660	-766.956	-427.490	-189.996	-184.681	-237.495	-229.710	-414.391	-149.658	-112.133
Density, lb/cuft	1.07	48.97	1.00	0.96				1.33	60.73	59.53	57.59	59.49	59.49	53.06	59.49	51.90	52.42	57.38	1.15
Mass Flow, lb/hr																			
CO	31,157	0.5	31,157.3	31,157.3	31,157.3	31,157.3	31,157.3	31,157.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	31,157.3
CO2	16,170	51.0	16,170.3	16,200.5	16,200.5	16,200.5	16,200.5	15,969.3	231.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	13,059.5
H2	1,309	0.1	1,308.5	1,308.5	1,308.5	1,308.5	1,308.5	1,308.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1,308.5
H2O	8,841	19,604.2	8,841.5	8,829.1	8,829.1	8,829.1	8,829.1	415.9	8,413.2	113,982.0	113,982.0	63,000.0	28,000.0	28,000.0	35,000.0	35,000.0	63,000.0	22,263.0	415.9
CH4	3,650	0.3	3,649.7	3,649.7	3,649.7	3,649.7	3,649.7	3,649.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3,631.7
H2S	934	9.0	933.9	957.3	957.3	957.3	957.3	943.3	14.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.3
COS	42	0.1	41.5	0.3	0.3	0.3	0.3	0.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.3
H3N	104	30.6	103.9	103.9	103.9	103.9	103.9	7.7	96.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	7.7
CHN	17	0.0	17.1	17.1	17.1	17.1	17.1	17.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	17.1
N2	73,622	1.2	73,621.8	73,621.8	73,621.8	73,621.8	73,621.8	73,621.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	73,621.8
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	0.0	0.0
Mole Flow, lbmol/hr																			
CO	1,112	0.0	1,112.3	1,112.3	1,112.3	1,112.3	1,112.3	1,112.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1,112.3
CO2	367	1.2	367.4	368.1	368.1	368.1	368.1	362.9	5.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	296.7
H2	649	0.0	649.1	649.1	649.1	649.1	649.1	649.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	649.1
H2O	491	1,088.2	490.8	490.1	490.1	490.1	490.1	23.1	467.0	6,327.0	6,327.0	3,497.0	1,554.2	1,554.2	1,942.8	1,942.8	3,497.0	1,235.8	23.1
CH4	227	0.0	227.5	227.5	227.5	227.5	227.5	227.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	226.4
H2S	27	0.3	27.4	28.1	28.1	28.1	28.1	27.7	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
COS	1	0.0	0.7	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	6	1.8	6.1	6.1	6.1	6.1	6.1	0.5	5.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.5
CHN	1	0.0	0.6	0.6	0.6	0.6	0.6	0.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.6
N2	2,628	0.0	2,628.1	2,628.1	2,628.1	2,628.1	2,628.1	2,628.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2,628.1
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	0.0	0.0

Table C.3 Sour Water Stripper Material and Energy Balance

Stream Number	702	707	801	802	803	804	805	806	808	809	810	811	812	813	814
Stream Decription	Sour Water from Venturi Scrubber	Process Condensate	Vapor from 3-Phase Separator	Liquid from 3-Phase Separator	SWS Feed	Pre-Heated SWS Feed	Sour Water Stripper Ovhd	Offgas to LoCat	Sour Water Stripper Btms	Cooled SWS Bttms	Recyle to Venturei Scrubber	Waste water from SWS	Blow Down from HP Steam Drum	Blow Down from HRSG	Cold Waste Water
Temperature, oF	250	110	142	142	142	249	140	140	267	157	157	157	451	450	110
Pressure, psia	330	295	20	20	75	70	35	20	40	65	65	65	440	420	15
Vapor Frac	0.000	0.000	1.000	0.000	0.000	0.002	1.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Mole Flow, lbmol/hr	1,092	478	0	1,570	1,570	1,570	17	18	1,552	1,552	1,236	317	131	179	626
Mass Flow, lb/hr	19,697	8,755	5	28,446	28,446	28,446	481	486	27,965	27,965	22,263	5,702	2,352	3,226	11,280
Volume Flow, cuft/hr	335	139	58	461	461	867	1,734	1,793	480	458	365	93	46	63	182
Enthalpy, MMBtu/hr	-130.690	-58.375	-0.016	-190.921	-190.916	-187.811	-1.708	-1.724	-185.520	-188.621	-150.160	-38.461	-15.145	-20.778	-76.619
Density, lb/cuft	51.4	58.7	0.09	61.7	61.7			0.27	58.3	61.1	61.1	61.1	51.4	51.5	61.9
Mass Flow, lb/hr															
CO	0.5	0.0	0.1	0.4	0.4	0.4	0.4	0.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	51.0	231.2	3.4	278.8	278.8	278.8	278.8	282.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2	0.1	0.0	0.1	0.0	0.0	0.0	0.0	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	19,604.2	8,413.2	0.5	28,017.0	28,017.0	28,017.0	53.7	54.2	27,963.3	27,963.3	22,261.3	5,701.9	2,352.0	3,226.0	11,279.9
CH4	0.3	0.0	0.0	0.3	0.3	0.3	0.3	0.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2S	9.0	14.0	0.2	22.8	22.8	22.8	22.8	22.9	0.1	0.1	0.1	0.0	0.0	0.0	0.0
COS	0.1	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H3N	30.6	96.2	0.0	126.8	126.8	126.8	124.8	124.8	2.0	2.0	1.6	0.4	0.0	0.0	0.4
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
N2	1.2	0.0	1.1	0.1	0.1	0.1	0.1	1.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Mole Flow, lbmol/hr															
CO	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
CO2	1.2	5.3	0.1	6.3	6.3	6.3	6.3	6.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2	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
H2O	1,088.2	467.0	0.0	1,555.2	1,555.2	1,555.2	3.0	3.0	1,552.2	1,552.2	1,235.7	316.5	130.6	179.1	626.1
CH4	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
H2S	0.3	0.4	0.0	0.7	0.7	0.7	0.7	0.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0
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
H3N	1.8	5.6	0.0	7.4	7.4	7.4	7.3	7.3	0.1	0.1	0.1	0.0	0.0	0.0	0.0
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
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

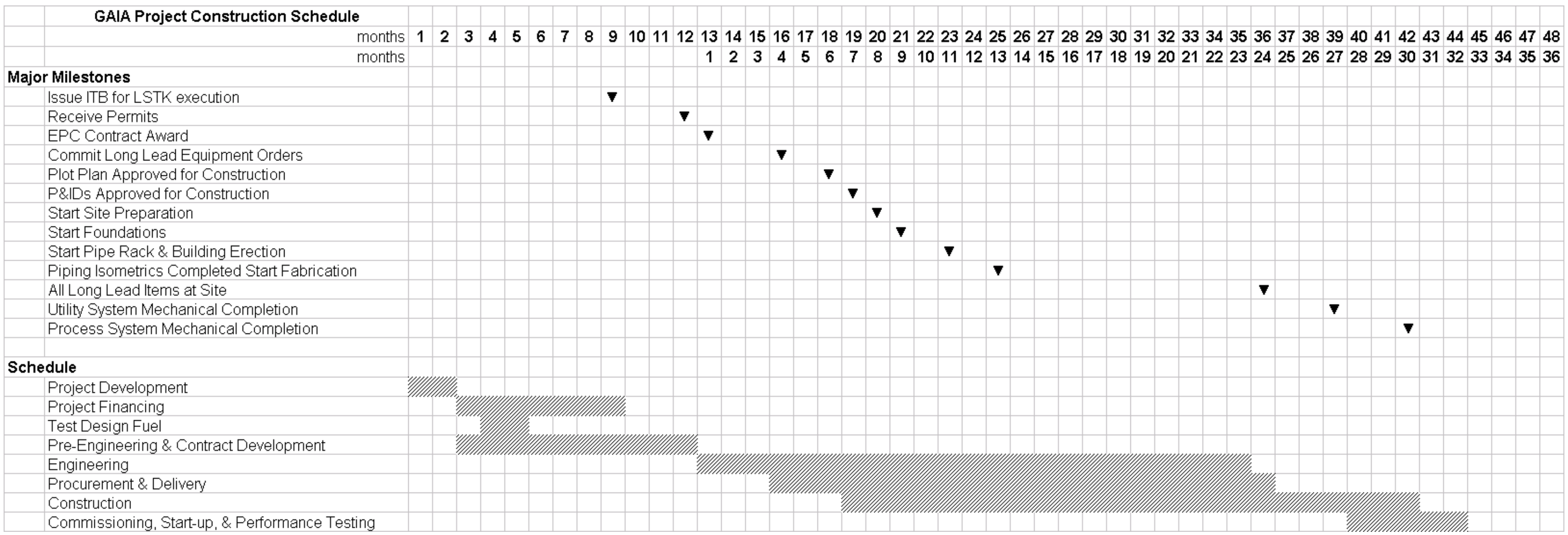
Table C.4 GT/HRSG Material and Energy Balance[1]

	901	903	904	905	906	907	908	909
	Air	BFW	400 Psig Super Heated Steam	50 Psig Super Heated Steam	400 Psig Evaporator Blowdown	Stack Exhaust	400 Psig Super Heated Steam - Export	400 Psig Super Heated Steam - Internal Use
Temperature F	65	150	548	304	449	254	548	548
Pressure psi	14.7	80.0	415.0	70.0	420.0	14.7	415.0	415.0
Vapor Frac	1.00	0.00	1.00	1.00	0.00	1.00	1.00	1.00
Mass Flow lb/hr	711396.8	113982.0	149518.4	22916.0	2221.2	834494.8	105343.2	44175.2
Volume Flow cuft/hr	9452980.0	2215.6	198620.1	144497.2	53.1	14959040.0	139937.9	58682.3
Enthalpy MMBtu/hr	-35.5	-773.2	-834.7	-130.1	-14.3	-390.5	-588.1	-246.6
Density lb/cuft	0.075	51.445	0.753	0.159	41.803	0.056	0.753	0.753
Mass Flow lb/hr								
AR	9069.4	0.0	0.0	0.0	0.0	9069.4	0.0	0.0
O2	163375.8	0.0	0.0	0.0	0.0	120698.8	0.0	0.0
N2	533054.1	0.0	0.0	0.0	0.0	606696.4	0.0	0.0
CO2	350.8	0.0	0.0	0.0	0.0	72109.9	0.0	0.0
CO	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CH4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2S	0.0	0.0	0.0	0.0	0.0	1.3	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	5546.8	113982.0	149518.4	22916.0	2221.2	25919.1	105343.2	44175.2
Mole Flow lbmol/hr								
AR	227.0	0.0	0.0	0.0	0.0	227.0	0.0	0.0
O2	5105.7	0.0	0.0	0.0	0.0	3772.0	0.0	0.0
N2	19028.5	0.0	0.0	0.0	0.0	21657.3	0.0	0.0
CO2	8.0	0.0	0.0	0.0	0.0	1638.5	0.0	0.0
CO	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CH4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	307.9	6327.0	8299.5	1272.0	123.3	1438.7	5847.4	2452.1

1 The results reflect the aggregate of both HRSG trains.

Figure D-1 shows the Project Construction Schedule. Project completion, as defined by completed performance testing, will occur 32 months after the award of the EPC contract.

Figure D.1 Project Construction Schedule



E.1 INTRODUCTION

The objective of this Design Basis is to define the process units and process support units including plant configuration for Subtask 3.3. This section includes the design basis and criteria for the subsequent engineering study and capital cost estimates. Subtask 3.3 is an alternate case for the preliminary case defined in Subtask 3.2. Subtask 3.3 is defined as follows:

- Revise the installation of an integrated coal gasification combined cycle (IGCC) facility in upstate New York to reduce capital and operating costs and to lower the plant emissions associated with power generation.

E.2 SUBTASK 3.3, ALTERNATIVE DESIGN FOR EASTERN COAL CASE

E.2.1 Plant Description

The U-GAS® plant located in upstate New York will consist of the following process blocks and subsystems:

- Unit 100: Coal Prep/Handling (Same as Subtask 3.2)
- Unit 150: Air Compression
- 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 Reactor, Low Temperature Heat Recovery and Mercury Removal
- Unit 800: Sulfur Removal, Sulfur Production, and Sour Water Stripper (SWS)
- Unit 900: Power Block including two GE 10 gas turbines (CT) and heat recovery steam generators (HSRG)
- 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 E.1 (in Section E.13 off this addendum) is a block flow diagram of the plant.

E.2.2 Site Selection

The upstate New York site is a large industrial site of over 1,800 acres. There are 5 locations that have been identified where this facility could be sited. Critical site issues include:

- Sufficient open space for all equipment
- Distance for power interconnect
- Ability to balance steam from the IGCC into the industrial infrastructure
- Access for coal storage and handling

The site is assumed to be level and cleared. Since the specific site within the industrial facilities has not been chosen, a generic plot plan will be prepared.

E.2.3 Feedstocks

The key to coal selection is to identify a cost effective candidate fuel for use at the industrial facility. Coal from Southeastern Ohio best fits these criteria. An existing Southeastern Ohio coal analysis will be used for design purposes. Seeking fuel bids and mine analysis at this time is not practical for the study. The coal analysis is shown in Section E.4 of this addendum.

Coal delivery to the site will be by rail. Drying facilities will be designed to handle delivered coal having up to 15 wt% moisture and will dry the coal to a maximum moisture content of 5 wt%. The expected feed coal moisture content will be 8.4 wt%.

E.2.4 Plant Capacity

The plant capacity will be about 25 MW of power generated from two GE 10 turbines. The determination of the exact coal-processing rate is part of this study. This rate is chosen so as to fully load two GE 10 gas turbines, and it is a function of the coal that is processed and the system design.

E.2.5 Configuration

The plant will be single train facility from gasification through syngas cleanup. There will be two GE 10 combustion turbines and two HRSGs.

E.2.6 Air Compression

The gasifier will be air-blown. Ambient air will be filtered and compressed for gasification.

E.2.7 Gasification Unit

The plant will have one gasifier that will operate at about 340 psig.

E.2.8 Power Block

Two (2) gas turbines (GE 10) will be used, each with a nominal rating of 12.5 MW each for a total power production of about 25 MW.

E.3 SITE CONDITIONS

Location	upstate New York
Elevation, feet	547
Air Temperature	
Maximum, ^oF	100
Annual Average, ^oF	48
Minimum, ^oF	-19
Seismic Zone	1

E.4 FEEDSTOCK

The following properties shall be used for the Southeastern Ohio coal.

Ultimate Analysis, wt%	Dry Basis	As Received	ASTM Method	Ash Fusion Temperature	^oF
C	74.65	68.38	D3176	IT	1974
H	5.79	5.3	D3176	ST	2025
N	1.54	1.41	D3176	HT	2049
S	3.32	3.04	D4239	FT	2067
Ash	5.91	5.41	D3176		
O	8.79	8.06	D3176	**Coal Ash Analysis**	**wt%**
Total	100.0	91.6		SiO2	33.3
				Al2O3	29.6
Proximate Analysis, wt%				Fe2O3	29.3
Residual Moisture			D3173	TiO2	0.6
Total Moisture		8.4	D3302	CaO	2.9
Ash	5.91	5.41	D3174	MgO	0.7
Volatile Matter	43.24	39.6	D3175	Na2O	0.4
Fixed Carbon	50.85	46.59	D3172	K2O	0.5
Total	100	100		SO3	2.1
Air-Dry Loss		5.53	D2013	P2O5	< 0.1
Sulfur	3.32	3.04	D4239	BaO	< 0.1
Gross Caloric Value, Btu/lb	13,590	12,448	D1898	Mn2O3	< 0.1
Dry, Ash Free, Btu/lb	14,443			SrO	< 0.1
Pounds SO2/MMBtu	4.88			Total	99.4

E.5 ELECTRIC POWER

Export Power, MW	Maximize
Voltage, kV	230

E.6 EXPORT STEAM PRODUCTION

Medium Pressure Steam	
Flow Rate, lb/hr	Maximize
Pressure at Delivery, psig	400
Temperature at Delivery, °F	550

E.7 WATER MAKE-UP

Source	Boiler Feed Water
Supply Pressure, psig	0
Supply Temperature, °F	150

E.8 NATURAL GAS

HHV, Btu/scf	1,050
LHV, Btu/scf	960

E.9 BY-PRODUCTS

Ash, tpd	TBD
Sulfur, tpd	TBD

E.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/MMBtu
Particulates	< 0.005 Lb/MMBtu
Mercury	> 90% removal
Target Thermal Efficiency	45-50%

E.11 FINANCIAL

Process Contingency (Gasifier block only)	25%
Project Contingency (ex. Gasifier block)	15%
Accuracy	+30/-15%
Capacity factor	85%

Fees (engineering, start-up, owner's costs)	10%
O&M	5%
Project, book and tax life	20 years
Tax rate	40%
Debt-to-equity ratio	2:1
Cost of capital	8%
Start-up	2015

E.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%.

E.13 BLOCK FLOW DIAGRAM

Figure E.1 Block Flow Diagram - Syngas Generation and Processing

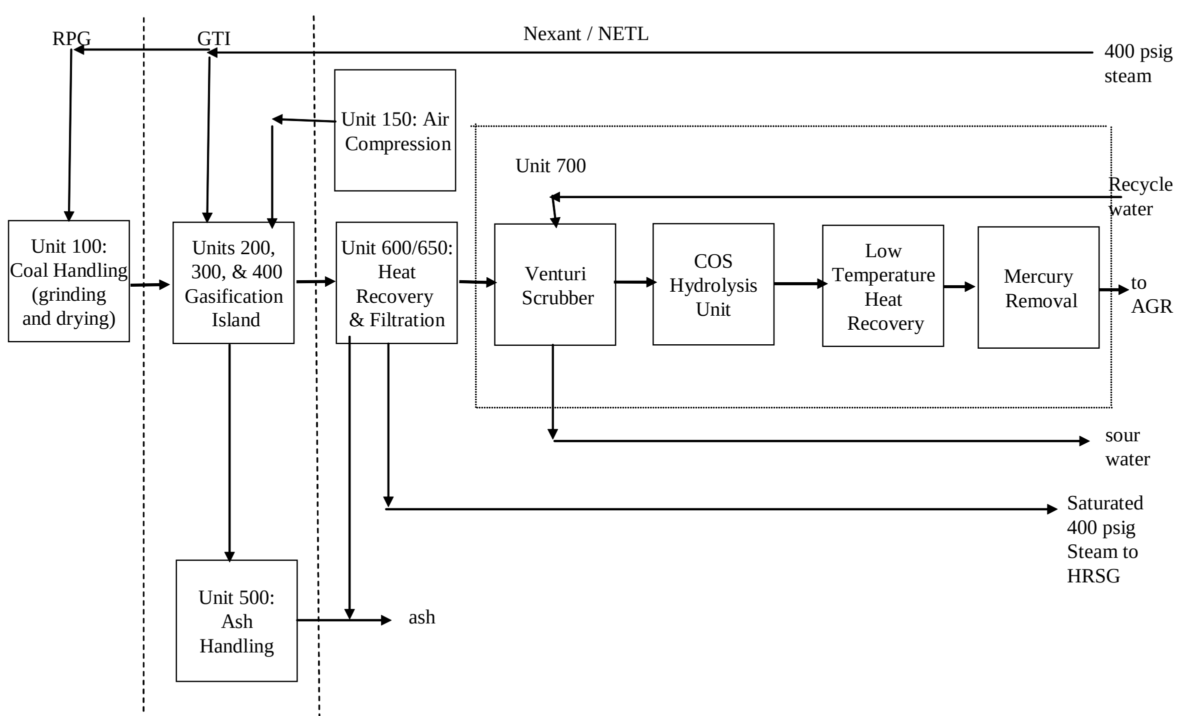


Figure E.2 Block Flow Diagram - Sulfur Removal and Recovery, Sour Water Stripper and Power Block