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Submitted To:
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Task 3 Gasification Plant Cost and Performance Optimization

DOE Contract No DE-AC26-99FT40342

Final Report

Submitted By:

Nexant

In association with:

gti

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Prepared For:

United States Department of Energy / National Energy Technology Laboratory

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Abstract

As part of an ongoing effort of the U.S. Department of Energy (DOE) to investigate the feasibility of gasification on a broader level, Nexant, Inc. was contracted to perform a comprehensive study to provide a set of gasification alternatives for consideration by the DOE. Nexant completed the first two tasks (Tasks 1 and 2) of the *Gasification Plant Cost and Performance Optimization Study* for the DOE's National Energy Technology Laboratory (NETL) in 2003. These tasks evaluated the use of the E-GAS™ gasification technology (now owned by ConocoPhillips) for the production of power either alone or with polygeneration of industrial grade steam, fuel gas, hydrocarbon liquids, or hydrogen. NETL expanded this effort in Task 3 to evaluate Gas Technology Institute's (GTI) fluidized bed U-GAS® gasifier.

The Task 3 study had three main objectives. The first was to examine the application of the gasifier at an industrial application in upstate New York using a Southeastern Ohio coal. The second was to investigate the GTI gasifier in a stand-alone lignite-fueled IGCC power plant application, sited in North Dakota. The final goal was to train NETL personnel in the methods of process design and systems analysis.

These objectives were divided into five subtasks. Subtasks 3.2 through 3.4 covered the technical analyses for the different design cases. Subtask 3.1 covered management activities, and Subtask 3.5 covered reporting.

Conceptual designs were developed for several coal gasification facilities based on the fluidized bed U-GAS® gasifier. Subtask 3.2 developed two base case designs for industrial combined heat and power facilities using Southeastern Ohio coal that will be located at an upstate New York location. One base case design used an air-blown gasifier, and the other used an oxygen-blown gasifier in order to evaluate their relative economics. Subtask 3.3 developed an advanced design for an air-blown gasification combined heat and power facility based on the Subtask 3.2 design. The air-blown case was chosen since it was less costly and had a better return on investment than the oxygen-blown gasifier case. Under appropriate conditions, this study showed a combined heat and power air-blown gasification facility could be an attractive option for upgrading or expanding the utilities area of industrial facilities.

Subtask 3.4 developed a base case design for a large lignite-fueled IGCC power plant that uses the advanced GE 7FB combustion turbine to be located at a generic North Dakota site. This plant uses low-level waste heat to dry the lignite that otherwise would be rejected to the atmosphere. Although this base case plant design is economically attractive, further enhancements should be investigated. Furthermore, since this is an oxygen-blown facility, it has the potential for capture and sequestration of CO_2.

The third objective for Task 3 was accomplished by having NETL personnel working closely with Nexant and Gas Technology Institute personnel during execution of this project.

Technology development will be the key to the long-term commercialization of gasification technologies. This will be important to the integration of this environmentally superior solid fuel technology into the existing mix of power plants and industrial facilities. As a result of this study, several areas have been identified in which research and development will further advance gasification technology. Such areas include improved system availability, development of warm-gas clean up technologies, and improved subsystem designs.

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Industrial facilities in the United States are facing stricter environmental regulations in the near future. Currently many industrial and large commercial boiler facilities meet emissions requirements by switching to fuel oil or natural gas to avoid the expense of installing post combustion emission controls. However, the increasing expense and price volatility of using these premium fuels has placed a financial burden on U.S. industry. As environmental rules tighten, industry will be forced to choose between continued expenditures for either 1) emission controls on coal boilers, 2) switching to more costly premium fuels, or 3) shutdowns of non-competitive facilities.

Coal gasification should be considered as another option for operators of industrial facilities. Gasification has the potential to reduce emissions, increase efficiency, and reduce operating costs compared to conventional steam boilers. Furthermore, the use of combined heat and power (CHP) at industrial facilities using coal can contribute to a significant increase in distributed generation (DG), resulting in improved local power grid security.

As part of an ongoing effort of the U.S. Department of Energy (DOE) to investigate the feasibility of gasification on a broader level, Nexant, Inc. was contracted to perform a comprehensive study to provide a set of gasification alternatives for consideration by the DOE. Nexant completed the first two tasks (Tasks 1 and 2) of the *Gasification Plant Cost and Performance Optimization Study* for the DOE's National Energy Technology Laboratory (NETL) in 2003. These first tasks were based on the E-GAS™ gasification technology (now owned by ConocoPhillips). NETL expanded this effort in Task 3 to evaluate Gas Technology Institute's (GTI) fluidized bed U-GAS® gasifier for two distinct applications:

- At an existing industrial site in upstate New York
- At a stand-alone lignite-fueled IGCC power plant in North Dakota

The first application evaluated the use of the U-GAS® gasifier at a plant that is considering replacement of outdated steam boilers. The cases assumed a system incorporating a fluidized bed U-GAS® gasifier coupled with two General Electric (GE) combustion gas turbines and heat recovery steam generators (HRSGs) to co-produce power and high pressure (2.76 MPa, 400 psig) steam at a specific industrial complex in upstate New York. It is hoped that the choice of this location will provide insight into typical retrofit issues similar to what many industrial complexes may encounter in the future.

The second application evaluated the U-GAS® gasifier at a large lignite-fueled IGCC power plant located in North Dakota. The U-GAS® unit, a dry feed fluidized bed gasifier, may be better suited to gasify high moisture content lignite for power applications than a slurry fed process. Unlike the first application, this task assumed that the facility would be a grass-roots, stand-alone power plant consisting of an oxygen-blown gasification train producing sufficient syngas to fully load a single GE 7FB combustion turbine. The plant will be fueled by North Dakota lignite and located at a generic North Dakota site.

THE BENEFITS OF COAL GASIFICATION

Gasification plants can provide industry with a viable alternative to conventional steam boilers. Gasification offers several advantages for a long-term solution. First, coal is an abundant, low-priced energy source that is expected to have a stable low price over the foreseeable future and can be conveniently stored to avoid fuel supply disruptions. Second, integrated gasification combined-cycle (IGCC) systems have higher thermal efficiencies than steam boilers, which reduce both fuel costs and the amount of carbon dioxide generated.

Pollution reduction also is simplified in gasification systems. Sulfur removal is easier, since the sulfur is removed from the syngas stream where it is more concentrated relative to post-combustion flue gas. NOx reduction is accomplished by the use of steam or nitrogen dilution in the gas turbine to reduce thermal NOx production. Mercury and heavy metals removal from syngas has been demonstrated by adsorption on sulfur-impregnated carbon. Furthermore, the potential exists for the capture and sequestration of carbon dioxide, particularly from oxygen-blown gasification systems.

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

In general, lignite contains a substantial amount of moisture, ranging from 30 to 70%. Consequently, the mined lignite has a low heating value (HHV basis), ranging from 14.65 to 19.31 MJ/kg (6,300 to 8,300 Btu/lb). When compared to bituminous and subbituminous coals, lignite is less expensive on a Btu basis, yet is less efficient to use because of the significant energy required to dry the fuel. As a result, lignite is generally consumed relatively close to the mine because of the higher shipping cost on a contained energy basis. Opportunities may exist to use lignite as a cost effective feed for IGCC power plants sited in lignite producing regions.

DESIGN INFORMATION AND RESULTS

As structured, this task had three basic objectives:

- Subtasks 3.2 and 3.3 investigated the first objective, which was to develop IGCC designs for the production of power and steam at an industrial facility in upstate New York (here industrial scale is considered to be less than 100 MW).

- Subtask 3.4 investigated the second objective, which was to examine the application of the U-GAS® gasifier for a stand-alone lignite-fueled IGCC power plant in North Dakota.

- The third objective was to train NETL employees in the methods of process design and systems analysis. This was accomplished by having these individuals work closely with Nexant and GTI personnel during execution of this project.

Conceptual designs were developed for coal gasification facilities using the fluidized bed U-GAS® gasifier. GTI's U-GAS® fluidized bed gasifier technology was chosen for this study for the following reasons:

- Fluidized bed technology is versatile and capable of gasifying a wide range of fuels including lignite.

- Fluidized bed technology can be operated in either the air-blown or oxygen-blown mode. This provides owners with an option to select technology that best meets their needs for efficiency or for process criteria.

- The scale of equipment is ideally suited for fluidized bed technology.

- The technology is ready for commercial deployment.

- Older cost studies examining this technology needed to be updated to better compare this technology with other gasification options.

The results of these analyses are presented in Tables ES-1a and b. Subtask 3.2 developed two base case industrial designs for combined heat and power production using Southeastern Ohio coal. The first design used an air-blown gasifier, and the other used an oxygen-blown gasifier. For operating flexibility, both designs contained two parallel gasification trains and two parallel combustion turbines, as requested by the industrial facility.

Subtask 3.3 developed an advanced design for an air-blown gasification facility based on the Subtask 3.2 air-blown case. This design used a single U-GAS® fluidized bed gasifier to power two GE 10 combustion turbines. It was assumed that by 2015, GTI will have had more experience with the U-GAS® technology and be confident in the turndown operations of the unit, such that it can provide the operating flexibility required. This case also contained other improvements to the Subtask 3.2 design, such as Stamet solids feeding system, a combined ash handling system, metallic candle filters, a venturi scrubber, a LO-CAT® sulfur recovery process, and improved heat integration and heat recovery.

Subtask 3.4 developed a base case design for an oxygen-blown, lignite-fueled IGCC power plant that uses the advanced GE 7FB combustion turbine, located at a generic North Dakota site.

Table ES-1a
Summary of the Task 3 Design Cases
(SI units)

	Subtask 3.2 Air-Blown Case	Subtask 3.2 Oxygen-Blown Case	Subtask 3.3 Alternate Air-Blown Case	Subtask 3.4 Lignite IGCC Power Plant
Design Inputs				
Coal Feed, moisture-free metric tpd	313.6	293.7	313.6	2,320.5
Coal Feed, moisture-free kg/s	3.630	3.399	3.630	26.858
Natural Gas, MW	1.5	2.1	0.0	2.6
Design Outputs				
Export Power, MW	21.7	23.3	21.3	251
Export Steam (2.76 MPa / 561K), kg/s	12.8	3.4	13.3	0.0
Sulfur, kg/s	0.113	0.109	0.113	0.199
Ash, kg/s	0.264	0.185	0.343	2.654
Equivalent Availability, %	85.7	82.6	84.7	87.3
Cold Gas Efficiency, % (HHV basis)	79.3	83.1	79.3	84.0
Net CHP Efficacy, % (HHV basis)[1]	49.0	29.1	49.7	
Net Electrical Efficiency, % (HHV basis)[2]				36.5
Plant Cost (2nd quarter 2004)				
Plant EPC Cost, M$[1]	90.0	100.2	82.1	410.5
Plant EPC Cost, $/kW[2]	3,090	4,057	2,755	1,635
Per Energy Input, k$/MW	784.5	899.4	715.5	596.1
Per Energy Output, k$/MW	1,601.0	3,095.8	1,438.6	1,631.7
Return on Investment, %[3]	5.9	<0	8.4	19.4
Power Selling Price for a 12% ROI, ¢/kWh[4]	9.02	11.8	8.51	4.7
Steam Selling Price for a 12% ROI, $/metric ton[5]	15.93	>40	12.77	NA

1. Net CHP efficacy is defined as (net electrical energy plus exteranlly exchanged heat) divided by total heating value energy of all direct and indirect input fuels
2. Net electrical efficiency is defined as net electrical energy divided by total heating value energy of all input fuels
3. EPC cost is on second quarter 2004 dollars at the plant site. Contingency, taxes, fees, and owners' costs are excluded.
4. Subtasks 3.2 and 3.3 are based on converting the export steam to power at an average steam turbine efficiency.
5. Subtasks 3.2 and 3.3 are based on 8.0 ¢/kWh and 13.23 $/metric ton of steam. Subtask 3.4 is based on a 6.08 ¢/kWh power price.
6. Subtasks 3.2 and 3.3 are based on a 13.23 $/metric ton steam price.
7. Subtasks 3.2 and 3.3 are based on a 8.0 ¢/kWh power price.

Table ES-1b
Summary of the Task 3 Design Cases
(English units)

	Subtask 3.2 Air-Blown Case	Subtask 3.2 Oxygen-Blown Case	Subtask 3.3 Alternate Air-Blown Case	Subtask 3.4 Lignite IGCC Power Plant
Design Inputs				
Coal Feed, moisture-free short tpd	345.7	323.8	345.7	2,558
Coal Feed, moisture-free lb/hr	28,810	26,980	28,810	213,160
Natural Gas, MBtu/hr	5.1	7.3	0.0	8.9
Design Outputs				
Export Power, MW	21.7	23.3	21.3	251.0
Export Steam (400 psig / 550°F), klb/hr	101.72	26.75	105.34	0
Sulfur, lb/hr	899	863	899	1,577
Ash, lb/hr	2,097	1,465	2,719	21,063
Equivalent Availability, %	85.7	82.6	84.7	87.3
Cold Gas Efficiency, % (HHV basis)	79.3	83.1	79.3	84.0
Net CHP Efficacy, % (HHV basis)[1]	49.0	29.1	49.7	
Net Electrical Efficiency, % (HHV basis)[2]				36.5
Plant Cost (2nd quarter 2004)				
Plant EPC Cost, M$[3]	90.0	100.2	82.1	410.5
Plant EPC Cost, $/kW[4]	3,090	4,057	2,755	1,635
Per Energy Input, k$/MBtu/hr	229.9	263.6	209.7	174.7
Per Energy Output, k$/MBtu/hr	469.2	907.3	421.6	478.2
Return on Investment, %[5]	5.9	<0	8.4	19.4
Power Selling Price for a 12% ROI, ¢/kWh[6]	9.02	11.8	8.51	4.7
Steam Selling Price for a 12% ROI, $/ short ton[7]	17.56	>40	14.08	NA

1. Net CHP efficacy is defined as (net electrical energy plus exteranlly exchanged heat) divided by total heating value energy of all direct and indirect input fuels
2. Net electrical efficiency is defined as net electrical energy divided by total heating value energy of all input fuels
3. EPC cost is on second quarter 2004 dollars at the plant site. Contingency, taxes, fees, and owners' costs are excluded.
4. Subtasks 3.2 and 3.3 are based on converting the export steam to power at an average steam turbine efficiency.
5. Subtasks 3.2 and 3.3 are based on 8.0 ¢/kWh and 12 $/short ton of steam. Subtask 3.4 is based on a 6.08 ¢/kWh power price.
6. Subtasks 3.2 and 3.3 are based on a 12 $/short ton steam price.
7. Subtasks 3.2 and 3.3 are based on a 8.0 ¢/kWh power price.

The Subtask 3.2 air-blown design exported 21.7 MW of power and 12.8 kg/s (101.7 klb/hr) of steam (2.76 MPa / 561K or 400 psig / 550°F) to the industrial facility. It required a capital investment of 90 M$ (million US dollars, 2nd quarter 2004). Based on the current value that the industrial facility pays for power and steam, the plant has an expected return on investment of 5.9%. It is encouraging to note that gasification designs of this scale can produce positive financial results even without considering credits for environmental improvements over standard coal boilers.

The Subtask 3.2 oxygen-blown design used the same design philosophy as the air-blown case. This comparison between air- and oxygen-blown cases was done to determine the most economic design for gasification at the industrial scale. While the oxygen-blown case produced about 7% more power, steam output was only 25% of the air-blown case due to higher parasitic energy requirements. It required over 10 M$ (2^{nd} quarter 2004) in additional capital investment, and, subsequently, did not have a positive return on investment. Due to the poor economics of the oxygen-blown case, it was eliminated from further consideration. This does not mean that oxygen-blown processes are never competitive, but at this capacity and for this specific site, the air-blown design was more economic.

Subtask 3.3 developed an advanced design for an industrial air-blown facility. Many of the improvements to the Subtask 3.2 design are currently not commercially proven for syngas applications, but are expected to be available for a 2015 startup. Improvements included enhanced sulfur removal, coal feeding, and particulate removal technologies. While the design has similar performance characteristics as Subtask 3.2 (2% less power and 3.5% more steam production), it had an 8 M$ (2^{nd} quarter 2004) lower capital investment. Based on the current value that the industrial facility pays for power and steam, the plant has an expected return on investment 8.4%, up from 5.9% for Subtask 3.2. This case is a good demonstration of the benefits that technological advancement can bring to gasification designs.

The return on investment for the proposed industrial cases would be higher if credits for the significant reductions in SO_2 and NOx from the gasification process and clean-up equipment are included. These credits could amount to more than 1 $M/yr for each pollutant, based on reductions relative to conventional coal boilers, and values for SO_2 and NOx of 287 $/metric ton (260 $/short ton)[1] and 3,031 $/metric ton (2,750 $/short ton)[2], respectively.

The positive return on investment (8.4%) for Subtask 3.3 shows that an industrial application for gasification with CHP can be economically viable in certain situations depending on site specific criteria, such as environmental credits, electricity and steam prices, and project return hurdle rates. Once environmental credits were included, the plant performance met the study goals set by the owner of the industrial facility at the beginning of this study. Although these goals were met, the industrial facility later decided not to pursue the project. These original project goals were:

- Reduce premium fuel usage
- Increase the use of coal
- Reduce emissions
- Reduce purchased power
- Have a payback of less than ten years

1 U.S. EPA Clean Air Markets, Allowance Trading, Chicago Board of Trade, 3/2004 Spot Auction Results, Clearing Price (lowest price at which a successful bid was made).

2 The Ozone Transport Region NOx Allowance Market, NOx emission allowance market under the Ozone Transport Commission Budget Rule in the Northeastern States, range from $1,000 to $4,500 per ton (1999-2002). Average used.

During the course of the study, the industrial facility's position changed concerning the payback period. The ultimate payback period desired was one more in-line with other industrial projects, as opposed to those typically used for utility applications. While the facility has decided not to pursue the project at this time, it should be made clear that this is not due to the projected technical or financial performance as a utility project. The study results demonstrate that industrial coal gasification can be a viable utility option for power and steam generation using actual industrial conditions.

The Subtask 3.4 oxygen-blown lignite-fueled IGCC power plant contains two U-GAS® fluidized bed gasifiers (one operating and one spare) to power one GE 7FB combustion turbine. An oxygen-blown design was selected for this application to allow for the possible future capture and sequestration of CO_2. The plant consumes 2,320 metric tpd (2,558 short tpd) of moisture-free coal and exports 251 MW of power. The lignite is dried from 32.2% moisture to 20% moisture before being fed to the gasifier. Low-level waste heat that otherwise would be rejected to the atmosphere is the primary heat source for drying the lignite. Use of this drying technique allows the plant to have a net electrical efficiency of 36.5% on a HHV basis. The estimated cost of the facility is 410 M$ (2nd quarter 2004) or about 1,635 $/kW of export power. Based on a power selling price of 6.08 ¢/kWh, the plant has an expected return on investment of 19.4%. These promising financial results show that lignite is not at a major disadvantage versus other coal types for gasification, and that greater consideration of its potential should be given for regional applications.

All designs met the emissions targets established by the DOE in their Clean Coal Technology Roadmap for 2010. These targets are:

- Sulfur > 99% removal
- NOx < 0.05 lb/MBtu
- Particulates < 0.005 lb/MBtu
- Mercury > 90% removal

Results of a sensitivity analysis showed that the capital investment, electricity tariff (price), and plant availability are the most sensitive financial parameters for all cases. Modifying the capital investment estimates by +/-15% changes the project return on investment by roughly 4 percentage points in all cases. Electricity tariff and availability had two to three times the impact on the project economics when compared to the other financial model inputs outside of total cost. Project developers should focus on obtaining the best possible estimates for these key items when estimating the financial return.

RECOMMENDATIONS

All the subtask designs are not fully optimized, and further improvements are possible. However, these designs demonstrate the potential of the GTI U-GAS® gasifier in these or similar applications. Improved technology currently is being developed in DOE's Gasification R&D Program in several areas, such as sulfur, mercury, particulate removal, and air separation, which may be able to reduce plant costs further and improve the efficiency.

As a result of this study, a list of potential enhancements has been identified that should provide additional cost savings as some of the improvements are researched, developed, and implemented. These enhancements, such as improved coal feed systems, development of warm-gas clean up technologies, improved fuel drying methods, and improved subsystem designs, could improve system performance and availability.

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

- A site specific analysis for siting of a lignite-fueled U-GAS® facility in North Dakota. The strong economics (19.4% return on investment) of this case shows great potential for wider use of this abundant, low-cost fuel source.

- Demonstration of warm gas clean-up technologies for sulfur and mercury removal

- Development of an R&D program to address critical issues such as availability and turbine performance

- Determination of the optimum moisture content for the coal feed to the gasifier and development of more efficient methods for coal drying

- Studies of the physical characteristics and properties of applicable coals to better predict gasifier performance

Efforts should be made to seek out other industrial facilities with suitable conditions for potential small gasifier-based industrial CHP project implementation

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.[1] These tasks analyzed 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's (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. Fluidized bed technology is versatile, capable of gasifying a wide range of fuels including lignite, and can be operated in either the air-blown or oxygen-blown mode. In addition, the scale of equipment considered herein is ideally suited for fluidized bed technology.

This report describes Task 3 of the *Gasification Plant Cost and Performance Optimization Study* and focuses on *Gasification Alternatives for Industrial Applications*. The initial 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 3.3 developed an alternate design for the Subtask 3.2 case that is lower cost and has a higher return on investment. 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. The scope for Subtask 3.4 changed during the project from a lignite-fueled industrial facility, to a larger, stand-alone power plant due to the desire by the DOE to design an economically viable oxygen-blown lignite case. Subtask 3.1 covered management activities.

Another objective of Task 3 was to train DOE employees in the methods of process design and system analysis. These individuals worked closely with the Nexant and Gas Technology Institute personnel in the execution of this task and the reporting of the results.

This is the Final Report for Task 3. The main body of this report briefly summarizes the three design subtasks and presents the principal results. The details are contained in the three Appendices:

- Appendix A is the Topical Report for Subtask 3.2, Preliminary Design for the Eastern Coal Case. This case is the base case design for a combined heat and power (CHP) facility that is located at an industrial site in upstate New York. The

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

plant uses two GTI U-GAS® gasifiers to power two GE 10 combustion turbines. Two designs were considered: an air-blown case and an oxygen-blown case.

- Appendix B is the Topical Report for Subtask 3.3, Alternate Design for the Eastern Coal Case. This is an alternate design for the air-blown case for Subtask 3.2 that improved the performance and reduced the cost. It also is a CHP facility that is located at an industrial site in upstate New York. This plant uses a single GTI U-GAS® gasifier to power the same two GE 10 combustion turbines as in Subtask 3.2. Design improvements resulted in a lower cost design that has a higher return on investment than the base case.
- Appendix C is the Topical Report for Subtask 3.4, Lignite-Fueled IGCC Power Plant. This plant is located at a generic site in North Dakota. It uses a single U-GAS® gasifier to power a single GE 7FB combustion turbine.

This final report is divided into the following nine sections.

Section	Title
1	Executive Summary
2	Introduction
3	Methodology
4	Industrial Combined Heat and Power Plants
5	Lignite-Fueled IGCC Power Plant
6	Summary, Conclusions and Recommendations
7	References
8	Acknowledgements
9	List of Acronyms and Abbreviations

Section 1 is the Executive Summary, providing a brief review of all project results.

Section 2, the Introduction, outlines the scope of the project and gives an overview of this final report.

Section 3 describes the methodology used in developing the plant designs and financial analyses.

Section 4 summarizes the designs and results for the three combined heat and power plant cases at the industrial facility located in upstate New York.

Section 5 summarizes the design and results for the lignite-fueled IGCC power plant located in North Dakota.

Section 6 summarizes the work that was done in Task 3, lists the major conclusions, and provides recommendations for further systems analysis and R&D efforts.

Section 7 contains a list of pertinent references.

Section 8 acknowledges the contributions of the individuals who contributed their talents to this project

Section 9 contains a list of acronyms and abbreviations used in this report.

Because this report describes plant designs that are based on proprietary information, some key details are omitted. However, the report contains sufficient information to allow the reader to assess the performance of GTI's gasification section design for each subtask. Basic heat and material balance information can be found in the block flow diagrams and the tables. This information was taken from detailed heat and material balances developed by the project team for each subtask. This information can be used to check the overall mass, carbon, and energy balances for the gasification plant and the power block, and possibly adapt these designs to new cases. However, the project team, particularly GTI, would prefer to generate project specific mass and energy balances under a secrecy agreement. Such an agreement will allow GTI to provide additional details and to share confidential information.

This section describes the methodology, calculation procedures, and assumptions that were used to develop the conceptual plant designs in this task. More details on each of the following items are contained in the appendices.

3.1 GASIFICATION MODELING

The gasification blocks (Unit 200 through Unit 500) were modeled and designed by Gas Technology Institute (GTI) using their proprietary computer modeling program. GTI's U-GAS® technology consists of a fluidized bed process that can operate over a wide range of temperatures ranging from 1172 to 1366K (1650°F to 2000°F) and pressures ranging from 0.1 to 7.1 MPa (1 to 70 atmospheres). The oxidant, steam and fuel enter the gasifier vessel from the bottom. Syngas leaves the top of the gasifier and then passes through two cyclone separators that recycle unburned carbon particles and flyash back to the gasifier. Agglomerated ash is removed from the bottom of the vessel, cooled, and sent to storage for disposal or sale.

An empirical model was developed for the gasification block in ASPEN Plus® based on the principle of restricted chemical equilibrium. The gasifier conditions were set at those defined by the GTI design.[1] Eight independent chemical reactions were specified, with the chemical equilibrium restricted by varying the temperature approach of each reaction to match the specified yields. This technique is reasonable since the reactor is not completely homogenous, and the various reactions have different rates of approach to equilibrium. Different parameters were required for each set of design conditions. Although this approach is not completely rigorous, it is a reasonable approach for predicting performance at conditions that are similar to the design case based on the limited amount of available information.

The specific modeling parameters for the three gasification reactor designs are given in the appropriate appendix.

3.2 DOWNSTREAM PROCESS MODELING

ASPEN Plus® was used to model the downstream process units in the syngas cleanup system and the sour water stripper to develop material and energy balances. In addition, modeling was done with sufficient detail to size (and subsequently cost) various pieces of process equipment. For example, the COS hydrolysis reactor was modeled by an RYIELD process block, with the design and cost of the unit provided by an outside supplier. Other units, such as the reactor preheater and reactor effluent cooler train, were modeled in sufficient detail to predict the required duties and equipment sizes.

The Peng Robinson–Boston Mathias (PR-BM) property method set was used for syngas, which contains some hydrocarbons. However, in the low-temperature heat

1 Aspen Technology, Inc., Ten Canal Park, Cambridge, MA 02141-2201. Versions 11.1.1 and 12.1 were used.

recovery system where water is condensed, the electrolyte non-random-two-liquid (ElectrolyteNRTL) property method set was used in order to predict the amounts of ammonia (NH_3), hydrogen sulfide (H_2S), and carbon dioxide (CO_2) that are dissolved in the condensate. The ElectrolyteNRTL property method also was used for the simulation of the sour water stripper to develop a conservative design for this unit.

Details of the downstream process modeling for each design case are given in the appropriate appendix. Process flow diagrams (PFDs), stream flowrates, and process compositions also are given in the appendices.

3.3 POWER BLOCK MODELING

The combustion turbines, HRSG (Heat Recovery Steam Generator), and steam turbine (for Subtask 3.4 only) were modeled using the GateCycle process simulation modeling software.[2] The GateCycle program contains default parameters for most combustion turbines when fired by natural gas. The default parameters for both the GE 10 and GE 7FB turbines were modified for syngas from information supplied by General Electric. The cleaned syngas composition that was generated by the ASPEN Plus® gasifier model was used in modeling the gas turbines.

When modeling the HRSG, the stack temperature was maintained above the acid dew point temperature (~389K, 240°F) so that condensation and corrosion would not occur within the system. In addition, sufficient 5.1 MPa (50 psig) superheated steam was generated to satisfy the internal steam demands of the upstream processing equipment. The balance of the steam generation was superheated high-pressure steam, some of which was consumed by the gasifier. In Subtasks 3.2 and 3.3, the industrial combined heat and power cases, 40.5 MPa (400 psig) superheated steam at about 561K (550°F) was supplied to the industrial facility.

In the Subtask 3.4 Lignite-Fueled IGCC Power Plant, the high-pressure steam (at 101 MPa at 839K, 1,000 psig at 1050°F) went directly to the steam turbine. 50.7 MPa (500 psig) steam at 727K (850°F) was extracted from the high pressure section of the steam turbine. A portion of this 50.7 MPa steam was desuperheated to 561K (550°F) and sent to the gasifier. The remainder was reheated to 839K (1050°F) in the HRSG and sent to the low-pressure section of the steam turbine. The GateCycle program also was used to model the reheat steam turbine.

3.4 COST ESTIMATION

For these evaluations, all investment costs are for the second quarter 2004 either at the upstate New York location (Subtasks 3.2 and 3.3) or for North Dakota (Subtask 3.4). Labor rates and productivities associated with these sites also were used.

The plant consists of the following process blocks and subsystems:

[2] GateCycle for Windows, Version 5.52.0.r, The General Electric Company, GE Energy, 1631 Bentley Parkway South, Minden, NV 89423.

- Unit 100: Coal Preparation – Handling, Sizing and Drying
- Unit 150: Air Separation Unit
- Unit 200: Solids Feeding System
- Unit 300: Gasification
- Unit 400: Fines Separation
- Unit 500: Ash Handling
- Unit 600: High Temperature Heat Recovery
- Unit 650: Particulate Removal (Subtask 3.3 and 3.4 only)
- Unit 700: Syngas Scrubber, COS Hydrolysis Reactor, Low Temperature Heat Recovery and Mercury Removal
- Unit 800: Sulfur Removal and Recovery, Sour Water Stripper (SWS)
- Unit 900: Power Block including a combustion turbine (CT) with a heat recovery steam generator (HSRG), and for Subtask 3.4 a single steam turbine.
- Unit 1000: Utilities (e.g., instrument and plant air, cooling water systems, firewater system) and other offsites (e.g., flare, DCS, plant roads, buildings, chemical storage)

The material and energy balances, along with the PFDs, establish the operating and design conditions for the individual pieces of equipment. The equipment was then sized and materials selected to provide a 20-year life. The Raymond Professional Group (Chicago, Illinois) provided the equipment list and sizing for Unit 100, coal handling and drying. Equipment sizing for Units 200 through 500 (with the exception of the Stamet pump) was prepared by GTI. The designs for most of the equipment in Units 600 through 1000 (excluding the COS hydrolysis, acid gas removal and sulfur recovery systems) were prepared by Nexant and NETL using the material and energy balances produced by ASPEN Plus® and GateCycle as the basis. An equipment list for each design is provided in the appropriate appendix.

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 costs to determine the total erected cost for each individual piece of equipment. This method is well founded both 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 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 piece of equipment, 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 was developed based on previous estimates for similar units. 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 several systems, for example the coal handling and drying unit (from the Raymond Professional Group), ASU (from Air Products), gas turbine (from General Electric), HRSG (from Vogt Power), and mercury removal (from Calgon Carbon)

Specific details for each case are contained in the appropriate appendix.

3.5 AVAILABILITY ANALYSIS

Common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period all are dependent on the annual plant performance, which defines the project cash flow. Although the design capacity is the major factor influencing the annual production, other influential factors 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.

The effect of sparing (back-up equipment or parallel trains of reduced capacity) can have a significant effect on the capacity factor (availability) of a plant depending upon the amount of spare equipment or parallel trains that are present. Sparing is most effective in increasing the overall plant availability when those portions of the plant with the lowest on-stream factors are replicated. Because operability is key to the design of any gasification facility, sparing plays an important role in the design development to provide optimum on-stream capacity while also attempting to maintain economic viability.

The availability of the gasification block was calculated based on information supplied by GTI from an analysis for the various components in the gasification block. They estimated that each gasification train would be out of service for unscheduled outages for approximately 9 days per year.

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.[3] For this analysis, most operations of the Task 3 facility, exclusive of the gasification block, are fundamentally similar to those of the Wabash River Repowering Project. This availability information formed the basis for the availability analysis. Subsequently, data presented at the 2002 Gasification Technologies Council conference showed further improvements in the on-stream performance of the Wabash River Gasification Repowering Project.[4] In addition to the gasification block, two other blocks, 1) coal preparation and handling and 2) mercury removal are not represented in the Wabash River final report. Availability estimates for those operations are based on the conceptual design, and are not based on actual operating experience. Additionally, availability estimates for the combustion turbine are based on the GE 7FA advanced combustion turbine design used at Wabash River. The turbine used for Subtasks 3.2 and 3.3, the GE 10, is not currently available for use with coal-derived syngas. Therefore, although demonstrated on-stream performance is not available for a syngas application, it was assumed to be the same as that of the GE 7FA turbine used at Wabash River.

Based on the above availability estimates and data, analyses were calculated using the Electric Power Research Institute (EPRI) recommended procedure.[5] This procedure calculates availabilities based only on two plant states: operating at design capacity or not operating. For a single train plant with all the units in a series configuration (i.e., no redundancy), the overall plant availability simply is the product all the individual unit availabilities. For multiple trains (or for plant sections with spare units), the EPRI report presents mathematical formulae based on a probabilistic approach for predicting the availability of all trains or combinations thereof, such as 1 of 2, 2 of 3, 1 of 3, etc. Appropriate combinations of these formulas are used to represent plants with some sections containing multiple trains or spare equipment, and other sections being single trains.

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

3 Global Energy, Inc., "Wabash River Coal Gasification Repowering Project – Final Report," September 2000.

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

5 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.

3.6 COMMODITY PRICING

The initial basis for the commodity prices used in the gasification model came from information provided by U.S. government agencies. This includes data from the DOE's Energy Information Administration (EIA) Annual Energy Outlook 2004[6] for commercial electricity values, natural gas, and coal, and the U.S. Geological Survey (USGS) for sulfur.[7] The steam value was calculated using natural gas as the marginal fuel for steam production, while the gasifier bottoms value was estimated using previous values from Nexant gasification studies. Each value was normalized where necessary to reflect current nominal value, using an appropriate escalation rate. Table 3.1 below lists the major assumptions for commodity prices.

Table 3.1 Basic Economic Parameters

	Subtasks 3.2 and 3.3		Subtask 3.4	
Description:	Industrial Combined Heat and Power		Lignite-Fueled IGCC Power Plant	
Location:	Upstate New York		North Dakota	
Feeds	**Price**	**Escalation, %/yr**	**Price**	**Escalation, %/yr**
Coal, \$/metric ton (\$/short ton)*	38.14 (34.60)	2.0	10.24 (9.29)	2.0
Natural Gas, HHV, \$/MW (\$/MBtu)**	1.37 (4.68)	4.0	1.76 (6.0)	4.0
Products				
Electric Power, cents/kWh***	8.0	3.0	6.082	3.0
Steam, \$/metric ton (\$/short ton)	13.23 (12.0)	3.0	NA	NA
Sulfur, \$/metric ton (\$/short ton)	29.23 (26.52)	3.0	29.23 (26.52)	3.0
Gasifier Bottoms, \$/metric ton (\$/short ton)	11.02 (10.0)	3.0	11.02 (10.0)	3.0

* As received coal price. The southeastern Ohio coal that is used at the upstate New York location contains 8.4% moisture, and the North Dakota lignite contains 32.24% moisture.

** The small natural gas usage is considered part of the operating and maintenance cost

*** The electric power price at the upstate New York plant is the price that the facility pays for delivered power, and the power price at North Dakota is the wholesale price.

For the CHP cases, the project team reviewed the initial estimates for the commodity prices, and made modifications to both the electricity and steam values to better reflect the actual costs currently incurred by at the upstate New York facility. The electric power value is that of the marginal supplier to the industrial facility. Sulfur, gasifier bottoms, natural gas, and coal values were left unchanged from the 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 are basically 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

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

7 Joyce Ober, US Geological Survey, Mineral Commodity Summaries, January 2004.

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 have 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. This number may be relevant to future tasks, if co-firing with natural gas is used, or as a comparison with other industrial power producing alternatives.

In keeping with previous Nexant studies and expectations of oversupply in the coal industry, the escalation rate was kept at 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 took a conservative approach 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 11 $/metric ton. This assumption was tested in the sensitivity analysis due to the volatile nature of this price, including negative value inputs.

3.7 FINANCIAL ANALYSIS

The reported ROIs (returns on investments) and NPVs (net present values) were calculated by the Nexant-developed IGCC Financial Model Version 3.01.[8] This version of the model was developed in May 2002 specifically for NETL under a task order from NETL on-site support contractor E^2S. The model has been used in previous gasification studies, and has undergone critical scrutiny by NETL and other technical experts. It is a robust discounted cash flow model that considers 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 1) NETL's "Quality Guidelines for Energy System Studies"[9], 2) an industry study analyzing the potential for gasification in the U.S. refining market[10], and 3) previous gasification optimization studies performed by Nexant, namely Tasks 1 and 2 of the "Gasification Plant Cost and Performance Optimization" study (DOE Contract number DE-AC26-99FT40342) for

[8] Nexant, Inc., 101 Second Street, San Francisco, CA, May 2002.
[9] McGurl, Gilbert V. et al, "Quality Guidelines for Energy System Studies", November 24, 2003.
[10] Gray, D. and Tomlinson, G., "Potential of Gasification in the U.S. Refining Industry", DOE Contract DE-AC22-95PC95054, June 2000.

NETL. A few of the major financial assumptions and some of the areas that were explored via sensitivity analysis are listed below:

- + 30/-15% accuracy assumed for this phase of the analysis
- A 15% project contingency applied across the entire plant with the exception of the gasifier block. For the gasifier block, a separate 25% process contingency was used to reflect the higher uncertainty in the cost estimate for this area.
- Scheduled annual downtimes of 14-21 days based on 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 for the Subtask 3.2 and 3.3 plants, and 42 months for Subtask 3.4. A construction schedule for each facility is given in the appropriate appendix.
- 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 balances provided by GTI and verified by Nexant/DOE, along with the subsequent design work by Nexant and NETL, set the values for items such as power output, steam production, sulfur produced, and quantity of gasifier bottoms. The plant EPC costs used for the model analysis were 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. Appropriate scale-up factors used in previous gasification projects allowed any equipment not reflecting installed cost to be properly estimated.

This section describes the three combined heat and power (CHP) plant designs that were developed for the industrial facility located in upstate New York. Listed below are the cases that were considered:

1. Subtask 3.2 developed two base case designs: an air-blown plant design and an oxygen-blown plant design. Both designs were developed to compare them for application at a typical industrial facility. When this project was started, it was unclear as to which design would be most attractive. This subtask focused on operability and proven technology as the key design criteria.

2. Subtask 3.3 developed an alternate air-blown plant design for 2015 that contained several improvements to the Subtask 3.2 base case air-blown design. This design has a higher return on investment and efficiency, along with a lower cost than the base case design. Unlike Subtask 3.2, this subtask focused on optimization efforts and used design learnings from the Subtask 3.2 base case.

The details of each of the three designs listed above are contained in the appendices.

The following five items summarize the general design basis of each of the three cases:

- The plant will be located at an industrial facility in upstate New York.

- Southeast Ohio coal with 8.4% moisture will be used. It will be dried to 5% for gasification.

- Two GE 10 gas turbines @ ~12.5 MW each (total = 25 MW) are utilized to offset power purchases from the local power supplier.

- The design is to maximize co-generation of steam from the gas turbines and HRSGs. Some of this will be used internally and the rest can be used for additional power generation, heating, and/or cooling within the industrial facility.

- Steam (2.76 MPa/561K, 400 psig/550°F) is exported to the industrial site.

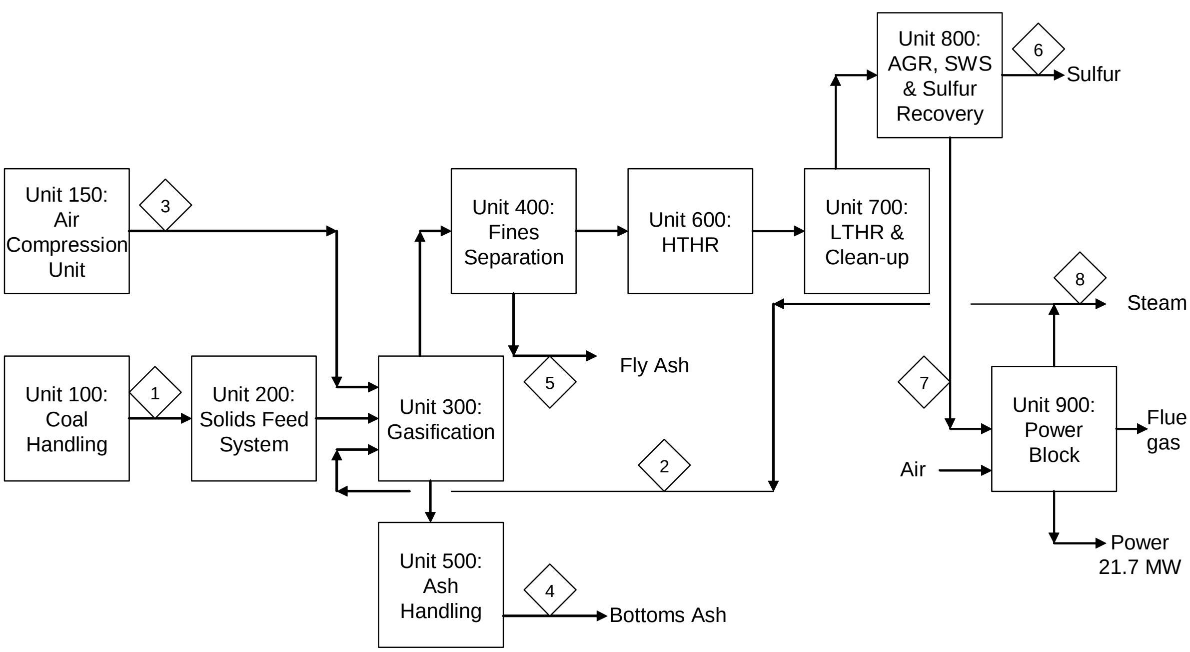
4.1 SUBTASK 3.2 AIR-BLOWN DESIGN

Figure 4.1 is a schematic flow diagram and overall material balance for the Subtask 3.2 air-blown design.

Southeast Ohio coal is delivered to the site by rail. Unit 100 dries it to less than 5% surface moisture and crushes it so that no more than 2% is greater than 6.4 mm (¼-inch) and no more than 10% is less than 100 mesh. It is then sent to the feed hoppers in Unit 200.

The prepared coal, at a rate of 3.81 kg/s (30,250 lb/hr), is fed to the bottom of the two parallel GTI fluidized bed U-GAS® gasifiers in Unit 300 through a series of weight and lock hoppers where it is mixed with compressed air from Unit 150 and steam. Syngas leaves the top of each gasifier and then passes through a series of two cyclone separators (Unit 400) that recycle unburned carbon particles and fly ash back to the gasifier. Unit 500 removes agglomerated ash from the bottom of the gasifier vessel, cools it, and sends it to storage for disposal.

Figure 4.1 Overall Heat and Material Balance for the Subtask 3.2 Air-Blown Design



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, 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)
Temperature, K (F)	294 (70)	561 (550)	533 (500)	1283 (1850)	1283 (1850)	NA	322 (120)	561 (550)
Flow Rate, kg/s (lb/hr)	3.8 (30,250)	1.2 (9,653)	12.1 (96,106)	0.2 (1,465)	0.1 (632)	0.1 (899)	15.5 (123,098)	12.8 (101,700)

Unit 600, the high-temperature heat recovery (HTHR) unit, cools the syngas to about 589K (600°F) by producing 2.93 MPa (425 psig) saturated steam. The cooled syngas then goes to unit 700, the low-temperature heat recovery (LTHR) unit.

In Unit 700, the syngas is cooled to about 402K (265°F) in an impingement scrubber column, which also removes any residual particulates from the syngas. The syngas then is reheated to prevent any condensation in the COS hydrolysis reactor where the carbonyl sulfide (COS) is converted to hydrogen sulfide (H_2S). Finally, the syngas is cooled to 316K (110°F) in a series of heat exchangers before cleanup.

Unit 800 cleans the syngas. Mercury is removed by adsorption on sulfur-impregnated carbon. H_2S is removed from the syngas by scrubbing with an amine solution. The cleaned syngas is then sent to the power block. Sulfur is recovered in a Claus system followed by a SCOT off-gas treating process to minimize sulfur emissions.

The cleaned syngas then is sent to the power block, Unit 900. The power block consists of two parallel GE 10 combustion turbines, each with a heat recovery steam generator (HRSG). The gas turbines produce about 15 MW of power each, some of which is consumed within the facility. The HRSGs produce both 0.34 MPa/450K (50 psig/350°F) superheated steam and 2.76 MPa/561K (400 psig/550°F) superheated steam. All of the 0.34 MPa (50 psig) steam and some of the 2.76 MPa (400 psig) steam is consumed internally. The plant exports 21.7 MW of power and 12.8 kg/s (101,700 lb/hr) of 2.76 MPa/561K (400 psig/550°F) superheated steam to the industrial facility.

4.2 SUBTASK 3.2 OXYGEN-BLOWN DESIGN

Figure 4.2 is a schematic flow diagram and overall material balance for the Subtask 3.2 oxygen-blown design. On the surface the design appears very similar to the previous air-blown design. However, there are several differences which are outlined below.

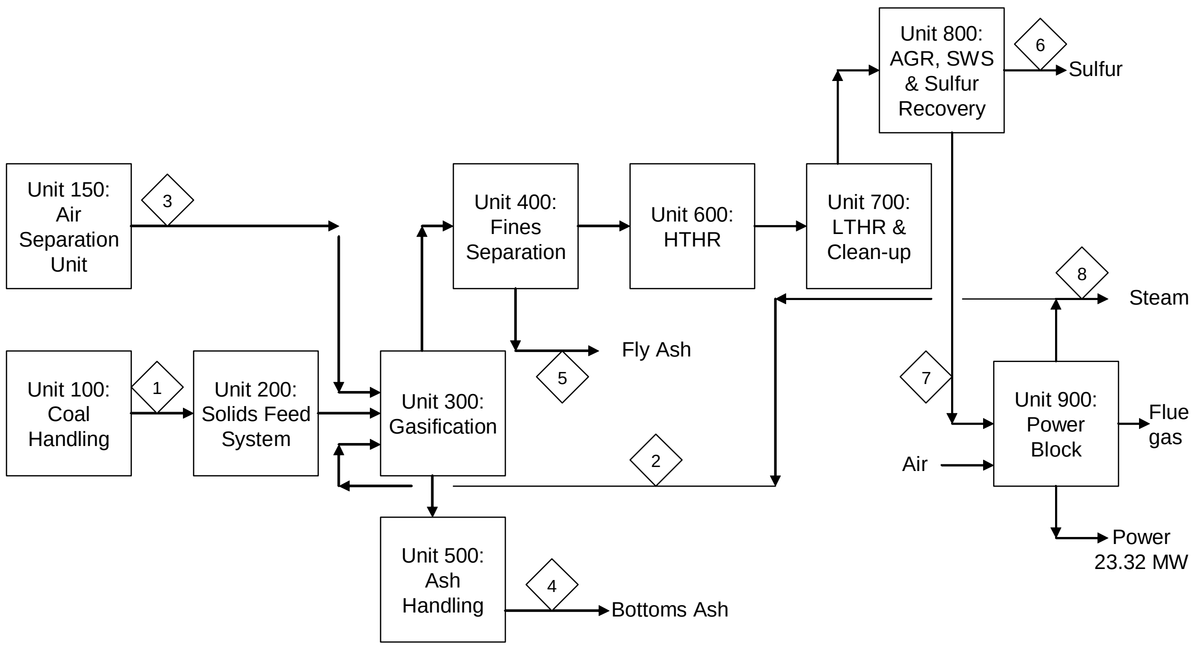
Unit 150, an Air Products' PRISM® APack™ air separation unit supplies a total of 2.48 kg/s (19,685 lb/hr) of 95% oxygen to the gasifiers via two parallel compressors (one associated with each gasifier for turndown purposes). Unit 150 also contains a nitrogen compressor to compress the nitrogen extracted from the air and sends it to the combustion turbines to reduce NOx formation.

Southeast Ohio coal is dried to 5% moisture and fed to the gasifiers at a rate of 3.58 kg/s (28,400 lb/hr). This is less coal than that required by the previous air-blown case because in the air-blown case, more energy is consumed during gasification to heat the nitrogen in the incoming oxidant stream.

Generally, units 600 through 800, which are downstream of the gasifier, are smaller than in the air-blown case due to lower volumetric flow of nitrogen in the syngas.

The two GE 10 combustion turbines in the power block each produce slightly less power than those in the air-blown case (14.9 MW vs. 15.0 MW) because of the lower mass flow. The plant exports 23.32 MW of power and 3.38 kg/s (26,800 lb/hr) of 2.76 MPa/561K (400 psig/550°F) superheated steam to the industrial facility.

Figure 4.2 Overall Material Balance for the Subtask 3.2 Oxygen-Blown Design



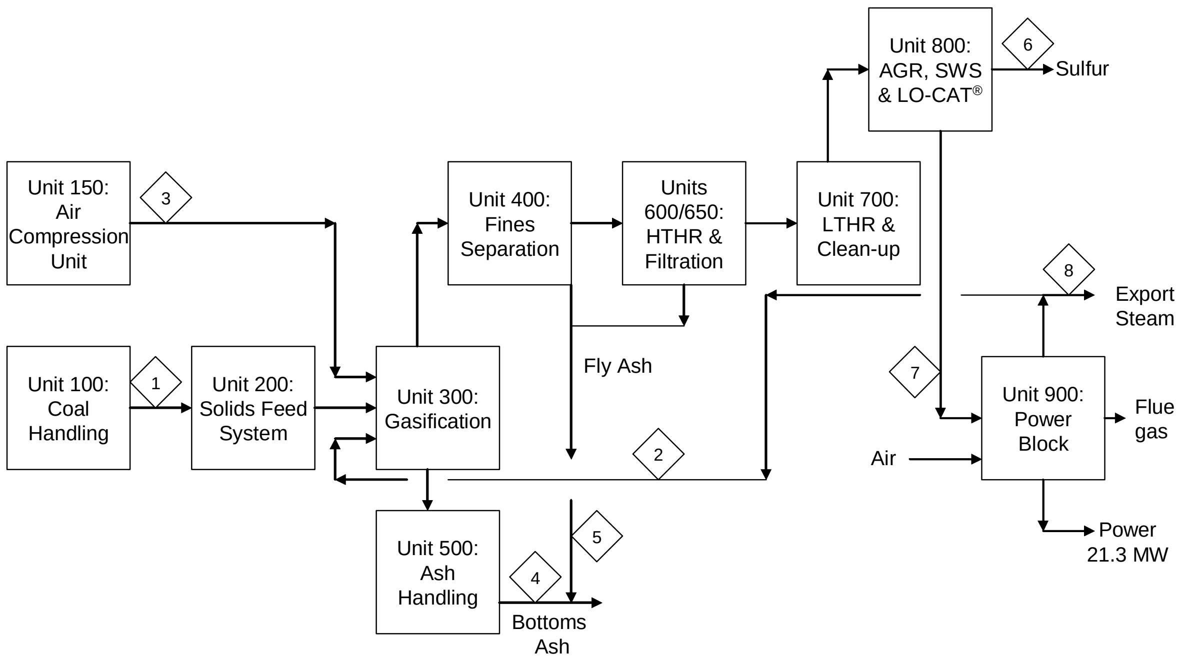
Stream Number	1	2	3	4	5	6	7	8
Description	**Coal to Gasifier**	**Steam to Gasifier**	**Oxidant to Gasifier**	**Bottom Ash from Gasifier**	**Fly Ash from Cyclones**	**Sulfur Product**	**Clean Syngas to Gas Turbine**	**Export Steam Production**
Pressure, MPa (psig)	0.10 (14.7)	2.90 (420)	2.86 (415)	0.24 (34.7)	0.24 (34.7)	NA	2.03 (295)	2.76 (400)
Temperature, K (F)	294 (70)	561 (550)	533 (500)	1283 (1850)	1283 (1850)	NA	322 (120)	561 (550)
Flow Rate, kg/s (lb/hr)	3.6 (28,400)	3.8 (30,208)	2.5 (19,685)	0.1 (872)	0.1 (594)	0.1 (863)	6.4 (50,881)	3.4 (26,800)

4.3 SUBTASK 3.3 ALTERNATE AIR-BLOWN DESIGN

The Subtask 3.3 Alternate Air-Blown Design was developed from the Subtask 3.2 Air-Blown Design by considering additional ideas generated during the Value Improving Practices (VIP) sessions for improving performance, reducing investment, and reducing operating costs. The start-up for this case is 2015, 7 years later than Subtask 3.2, allowing technologies that are expected to be commercial by that time to be used. The result is a case that is less costly, more efficient, and has a higher ROI than the Subtask 3.2 design. Modifications were made to improve the economics of the design by reducing the cost of facility. Figure 4.3 is a schematic flow diagram and material balance for the Subtask 3.3 Alternate Air-Blown Design.

A number of significant improvements were made to the Subtask 3.2 design in developing the Subtask 3.3 case. Some of these, such as the Stamet solids pump, have not been tested for the desired process conditions, but are expected to be commercially proven by the 2015 plant start up. These changes are discussed in detail in Appendix B. The following section describes the changes on a unit by unit basis.

Figure 4.3 Overall Heat and Material Balance for the Subtask 3.3 Alternate Air-Blown Design



ım 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**
;ure, 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)
ɔerature, 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)

Units 200 through 500, the gasification block, were redesigned to contain a single 100% capacity gasifier instead of two smaller 50% ones. A trade-off study for a spare gasifier showed that the return on investment will be lower with the spare gasifier and therefore, it was not included in the final design. Economy of scale has a direct impact on these results. The cost of the spare gasifier is a significant portion of the total investment for this small capacity plant, whereas the larger lignite-fueled power plant can justify a spare gasifier (see section 5.1).

In Unit 200, a pair of 100% Stamet solids pumps replaced the gasifier feed lockhopper and screw system. These Stamet pumps are less expensive than the lockhopper and screw systems that they replaced, but consume considerably more power. A financial analysis showed the economic benefit is marginal, but demonstrated performance of these pumps at lower pressures indicates that they should enhance the overall plant availability.

The bottoms and fly ash handling systems in Units 400 and 500 have been combined to simplify the design and remove extra pieces of equipment. The hot fly ash from the third stage cyclone now is mixed with the hot bottom ash and fed to a common cooling screw and lockhopper system. The cooled combined ash is sent to a common storage vessel for sale or disposal. Previously, each system had separate cooling screws, lockhopper systems, and storage vessels. Furthermore, the heated BFW from the cooling screws was returned to the industrial facility, but in this case, the BFW is consumed within the HRSG.

Unit 650, a metallic candle filter particulate removal system, has been added following Unit 600, the high temperature heat recovery (HTHR) steam generator. This system removes the remaining particulates from the syngas as a dry material and mixes it with the cooled bottoms ash and fly ash from the third stage cyclones for disposal. Previously, this material was removed in the syngas scrubber column and discarded with the wash water. Now it is recovered as a dry material that can be sold.

Ceramic candle filters that were located upstream of the HTHR steam generator also were considered. Although these filters would simplify the design of the HTHR steam generator, they were rejected because they were not economic.

Unit 700, the low-temperature heat recovery (LTHR) and syngas clean-up section, was modified to add an additional heat recovery step and replaced the impingement scrubber with a venturi scrubber followed by a smaller wash column. The additional heat exchanger preheats BFW going to the steam generator. The venturi scrubber system now requires less water to scrub and cool the syngas because 1) the syngas now is cooler, and 2) the syngas is particulate free since upstream candle filters have removed the particulates.

Unit 800, the acid gas removal (AGR), sour water stripper (SWS), and sulfur recovery area now contains a LO-CAT® process instead of a Claus plant with a SCOT tail gas treatment system. This change was made due to the superior economics of the LO-CAT® system, and the expectation that the system will have greater experience operating on syngas streams by 2015. The H_2S rich stream from the amine unit now goes to the LO-CAT® unit instead of the Claus plant. Furthermore, the use of the LO-CAT® process eliminates the need for any natural gas that would normally be consumed in the Claus plant.

Furthermore, the wastewater treatment area became simpler because the sour water from the syngas scrubber now is essentially solids-free. The sour water stripper is smaller due to less water used in the venturi scrubber.

Unit 900, the power block, was modified to reduce the amount of heat rejected to the atmosphere by enlarging the BFW coil in the HRSG and using some of this BFW to heat the syngas going to the combustion turbine. After heating the BFW, the cooled BFW is returned to the HRSG.

The Subtask 3.3 alternate air-blown gasifier consumes 3.81 kg/a (30,250 lb/hr) of Southeast Ohio coal (with a 5% moisture content) and exports 21.33 MW of electric power and 13.3 kg/s (105.34 klb/hr) of superheated 2.76 MPa/561K (400 psig/550°F) steam to the adjacent industrial facility.

4.4 COMPARISON OF THE SUBTASK 3.2 AND 3.3 COMBINED HEAT AND POWER DESIGNS

Table 4.1 compares the combined heat and power plant designs for the Subtask 3.2 Air-Blown Design, the Subtask 3.2 Oxygen-Blown Design, and the Subtask 3.3 Alternate Air-Blown Design.

Table 4.1a Comparison of the Subtask 3.2 and Subtask 3.3 Plant Designs

	Subtask 3.2 Air-Blown Design	Subtask 3.2 Oxygen-Blown Design	Subtask 3.3 Alternate Air-Blown Design
Design Inputs			
Coal Feed, moisture-free metric tpd	313.6	293.7	313.6
Coal Feed, moisture-free kg/s	3.630	3.399	3.630
Fuel (Natural Gas), MW	1.5	2.1	0
Makeup Water Input from the Industrial Facility			
Boiler Feed Water, m^3/s	0.031	0.030	0.026
Quench Water, m^3/s	0.002	0.004	0
Cooling Tower Makeup Water, m^3/s	0.003	0.004	0.004
Design Outputs			
Export Power, MW	21.7	23.3	21.3
Export Steam (400 psig, 550°F), kg/s	12.8	3.4	13.3
Sulfur, kg/s	0.113	0.109	0.113
Ash, kg/s	0.264	0.185	0.343
Condensate (to industrial facility), kg/s	7.67	8.25	6.85
EPC Cost, M$*	90.0	100.2	82.1
Plant EPC Cost, $/kW**	3,090	4,057	2,755
Plant Energy Input, k$/MW	784.5	899.4	715.5
Plant Energy Output, k$/MW	1,601.0	3,095.8	1.438.6
Equivalent Availability, %	85.7	82.6	84.7
Return on Investment, %***	5.9	<0	8.4
Cold Gas Efficiency, % (HHV basis)	79.3	83.1	79.3
Net CHP Efficacy, % (HHV basis)†	49.0	29.1	49.7

* 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 13.23 $/metric ton (12 $/short ton) of steam

† Net CHP efficacy is defined as the net electrical energy plus external exchanged heat divided by the total heating value energy of all direct and indirect input fuels

As seen in the table, the Subtask 3.2 oxygen-blown design produces the most export power from the least amount of coal. However, this case exports 75% less steam and has the highest EPC cost. The Subtask 3.2 Air-Blown Design produces slightly more export power and less export steam than the Subtask 3.3 Alternate Air-Blown Design from the same amount of feed coal. Since the Subtask 3.3 Air-Blown Design is by far the least expensive of the three designs, it produced highest overall return on investment (ROI). Further breakdowns of the EPC costs for all three designs are given in the appendices.

Table 4.1b Comparison of the Subtask 3.2 and Subtask 3.3 Plant Designs
(English units)

	Subtask 3.2 Air-Blown Design	**Subtask 3.2 Oxygen-Blown Design**	**Subtask 3.3 Alternate Air-Blown Design**
Design Inputs			
Coal Feed, moisture-free short tpd	345.7	323.8	345.7
Coal Feed, moisture-free lb/hr	28,810	26,980	28,810
Fuel (Natural Gas), MBtu/hr	5.1	7.3	0
Makeup Water Input from the Industrial Facility			
Boiler Feed Water, gpm	495	473	418
Quench Water, gpm	30	70	0
Cooling Tower Makeup Water, gpm	53	72	58
Design Outputs			
Export Power, MW	21.7	23.3	21.3
Export Steam (400 psig, 550°F), Mlb/hr	101.72	26.75	105.34
Sulfur, lb/hr	899	863	899
Ash, lb/hr	2,097	1,465	2,719
Condensate (to industrial facility), Mlb/hr	60.9	65.5	54.4
EPC Cost, M$*	90.0	100.2	82.1
Plant EPC Cost, $/kW**	3,090	4,057	2,755
Plant Energy Input, k$/MBtu/hr	229,872	263,587	209,695
Plant Energy Output, k$/MBtu/hr	469,209	907,280	421,585
Equivalent Availability, %	85.7	82.6	84.7
Return on Investment, %***	5.9	<0	8.4
Cold Gas Efficiency, % (HHV basis)	79.3	83.1	79.3
Net CHP Efficacy, % (HHV basis) †	49.0	29.1	49.7

* 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 $/short ton of steam

† Net CHP efficacy is defined as the net electrical energy plus external exchanged heat divided by the total heating value energy of all direct and indirect input fuels

Table 4.2 compares the financial results for the three cases. With an 8.0 cents/kWh (80 $/MWh) export power price and a steam price of 13.23 $/metric ton (12 $/short ton), the Subtask 3.3 Alternate Air-Blown Design has the highest ROI of 8.4% and the highest NPV at a 10% discount rate. It also has the lowest export power price required to give a 12% ROI with a fixed steam price, and the lowest steam price required to give a 12% ROI with a fixed export power price.

Table 4.2 Financial Comparison of the Subtask 3.2 and Subtask 3.3 Plant Designs

Designs	Subtask 3.2 Air-Blown Design	Subtask 3.2 Oxygen-Blown Design	Subtask 3.3 Alternate Air-Blown Design
Return on Investment (ROI), %*	5.9	<0	8.4
Net Present Value (NPV) at 10% Discount Rate, M$	-14.6	-48.6	-5.2
Number of Years to Payback	17	>20	14
Electricity Selling Price for a 12% ROI, cents/kWh**	9.02	11.8	8.51
Steam Selling Price for 12% ROI, $/metric ton ($/short ton)***	19.36 (17.56)	>40 (>40)	15.52 (14.08)

* With an export power price of 8.0 cents/kWh and a steam price of 13.23 $/metric ton (12.0 $/short ton)

** With a steam price of 13.23 $/metric ton (12.0 $/short ton)

*** With an export power price of 8.0 cents/kWh

Figures 4.4 and 4.5 show the expected ROI and NPV at a 10% discount rate versus the electricity tariff for the three Subtask 3.2 and Subtask 3.3 plants. The reference power price of 80 $/MWh is indicated by an arrow on the abscissa. Both of these figures dramatically show how much more economic the two air-blown cases are compared to the Subtask 3.2 oxygen-blown case. Furthermore, these figures demonstrate that the improvements made to the Subtask 3.3 Alternate Air-Blown Design significantly improved both the ROI and NPV of this case compared to the Subtask 3.2 Air-Blown Design. The Subtask 3.2 Oxygen-Blown Design is significantly inferior to either air-blown case for the CHP applications analyzed for this report.

Based on the above financial comparison, the Subtask 3.3 Alternate Air-Blown Design is economically superior to the other designs. Hence, the remainder of the discussion in the section will be confined to the Subtask 3.3 Alternate Air-Blown Design.

Figure 4.4 Return on Investment vs. Electricity Tariff for the Subtask 3.2 and Subtask 3.3 Plant Designs

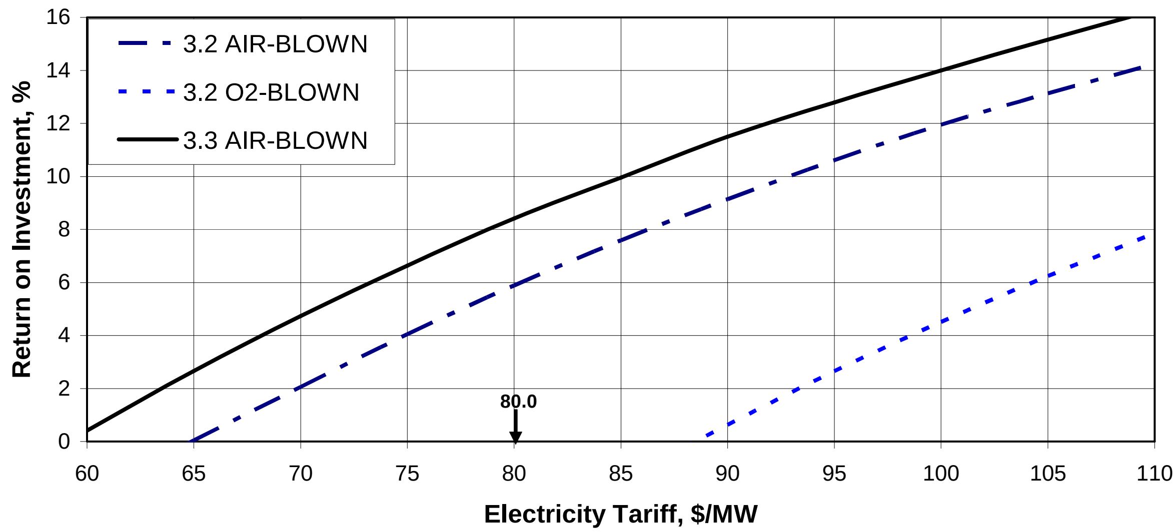


Figure 4.5 Net Present Value at a 10% Discount Rate vs. Electricity Tariff for the Subtask 3.2 and Subtask 3.3 Plant Designs

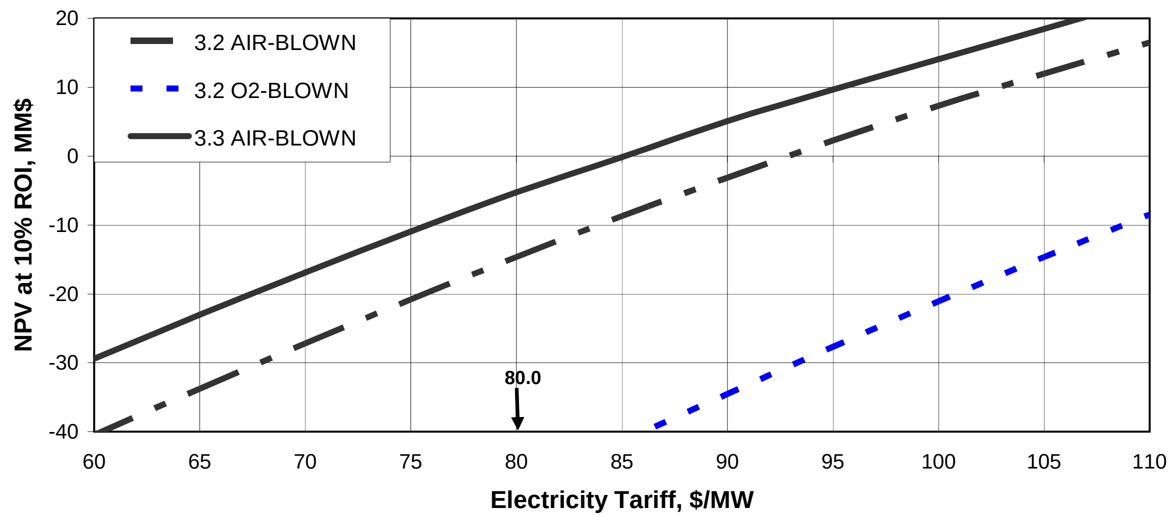
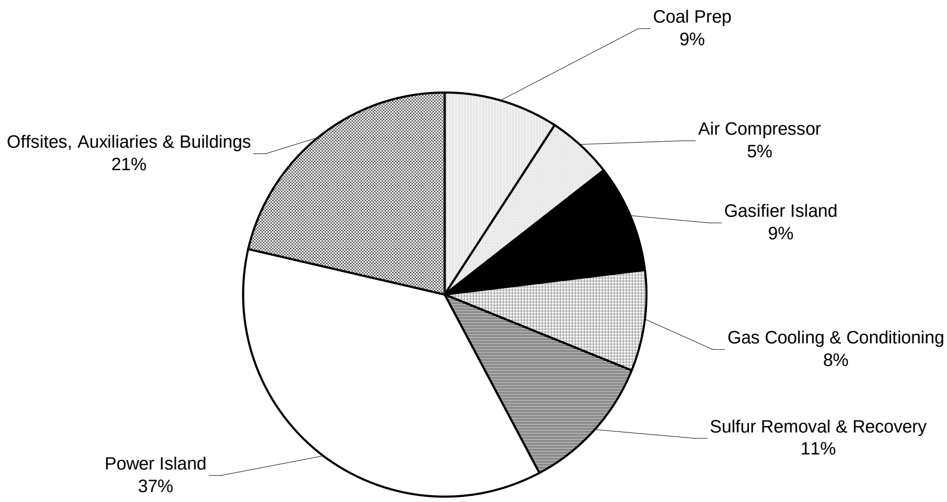


Figure 4.6 presents the EPC cost distribution of the Alternate Air-Blown Case. The EPC cost is estimated to be 82.1 MM$ on a second quarter 2004 basis. The investment is adjusted for labor rates and productivity in upstate New York.

Figure 4.6 Capital Cost for Alternate Air-Blown Case
(82.1 million US$)



4.5 SENSITIVITY OF THE SUBTASK 3.3 ALTERNATE AIR-BLOWN DESIGN TO THE FINANCIAL PARAMETERS

Table 4.3 shows the rate of return, NPV, payback year, and required electricity and steam selling prices to obtain a 12% ROI with the other values fixed. There are two major products from this facility, electricity and steam, and the pricing of both must be considered when determining the suitability of this project. Besides the base case, a "Low" and "High" estimate is shown reflecting the current cost accuracy assumption of -15%/+30%.

Figure 4.7 shows the sensitivities of the Subtask 3.3 Alternate Air-Blown Design to plus and minus 10% changes in the various economic parameters. A 10% change in the electricity tariff from the base price of 8.0 cents/kWh has the most influence on the NPV. A 10% change in the equivalent availability (annual average on-stream time) from the base value of 84.9% has the next largest effect. The consequence of 10% changes in the steam price (from a base price of 13.23 $/metric ton), loan interest rate (from a base rate of 8%), project life (from a base life of 20 years), total O&M costs (from a base value of 5%), and coal price (from a delivered price of 38.14 $/metric ton) are next in importance. A 10% change in the percent debt (loan amount from a base amount of 66%), income tax rate (from a base rate of 40%), owners contingency (from a base value of 18.8%) and coal escalation (from a base value of 2%) are less important. Finally, changes of 10% in either the ash or sulfur prices have almost an insignificant effect on the project NPV.

Table 4.3 Sensitivity of the Subtask 3.3 Alternate Air-Blown Design to the EPC Cost

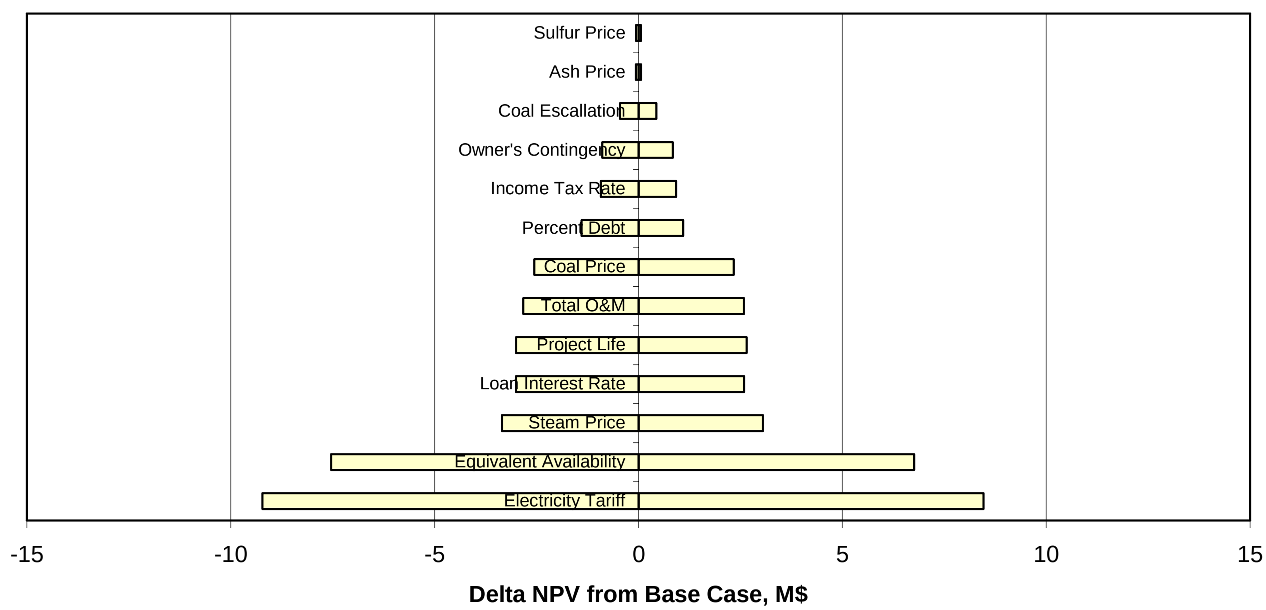
	Subtask 3.3 Air-Blown Design	EPC Cost Reduced by 15%	EPC Cost Increased by 30%
EPC Cost, M$	82.1	69.8	106.7
Return on Investment (ROI), %*	8.4	12.9	0.15
Net Present Value (NPV) at 10% Discount Rate, M$	-5.2	8.6	-38.8
Number of Years to Payback	14	9	>20
Electricity Selling Price for a 12% ROI, cents/kWh**	8.51	7.11	11.31
Steam Selling Price for a 12% ROI, $/metric ton, ($/short ton)***	15.52 (14.08)	9.27 (8.41)	28.00 (25.40)

* Export power price of 8.0 cents/kWh and a steam price of 13.23 $/metric ton (12.0 $/short ton)

** With a steam price of 13.23 $/metric ton (12.0 $/short ton)

*** With an export power price of 8.0 cents/kWh

Figure 4.7 Sensitivities of a +/-10% Change in Selected Inputs on Project NPV for Subtask 3.3 Alternate Air-Blown Design



Figures 4.8 and 4.9 show the expected ROI and NPV versus the equivalent availability for the Subtask 3.3 Alternate Air-Blown Design. An arrow in the abscissa shows the base case design equivalent availability of 84.9%. This equivalent availability is the annual average availability including scheduled shutdowns for the facility. Long downtimes throughout the life of the project will have a negative affect on the overall

project economics at a given project life. Scheduled operating hours also will have a similar negative impact since they are related to availability.

Figure 4.8 Return on Investment vs. Equivalent Availability for the Subtask 3.3 Alternate Air-Blown Design

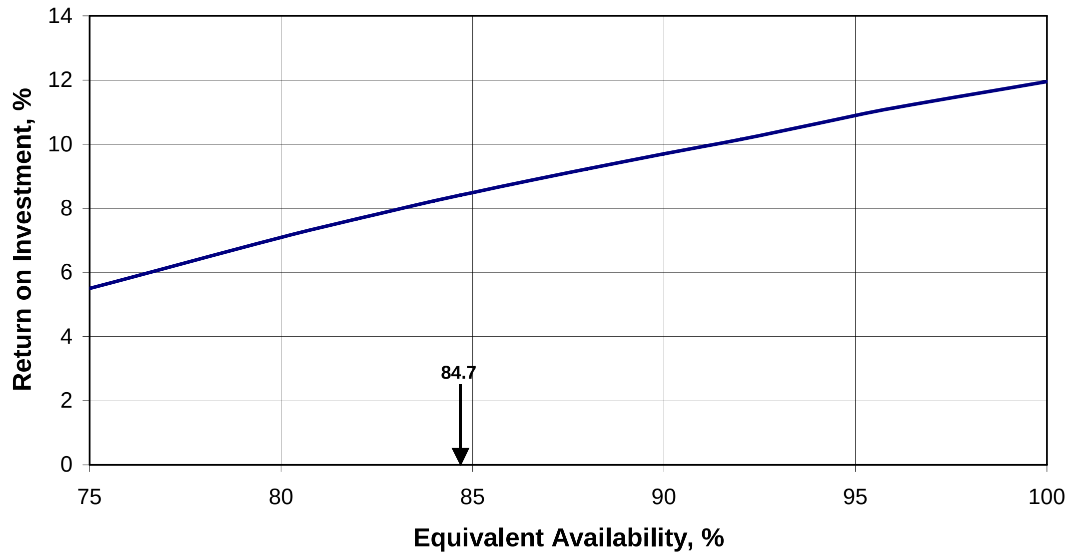
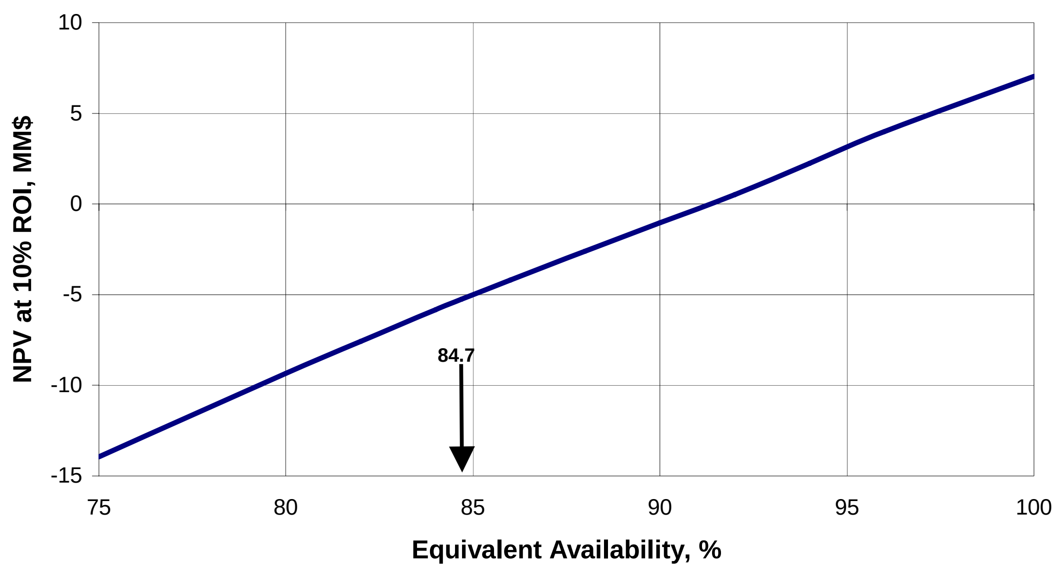


Figure 4.9 NPV at a 10% Discount Rate vs. Equivalent Availability for the Subtask 3.3 Alternate Air-Blown Design



4.6 SUMMARY OF THE SUBTASK 3.2 AND 3.3 PLANT DESIGNS

Three cases were developed for the Industrial Heat and Power scenario using the GTI fluidized bed U-GAS® gasifier to be integrated with an industrial facility in upstate New York. These cases were designed to maximize export steam production (2.76 MPa/561K (400 psig/550°F) superheated steam) while fully loading two GE 10 combustion turbines.

In this application, both air-blown gasifier designs had better economics than the oxygen-blown gasifier design because of the lower investment required and higher net CHP efficacy achieved. The Subtask 3.3 Alternate Air-Blown Design included several improvements to the base case Subtask 3.2 Air-Blown Design which reduced the cost, improved the efficiency, and increased the return on investment. These improvements included:

- Single train gasification island and HTHR
- Stamet solids feeding system
- Combined ash handling system
- Metallic candle filters
- Venturi scrubber for chlorides and light hydrocarbon removal
- LO-CAT sulfur recovery process
- Improved heat integration/heat recovery

The parameters that have the greatest impact on the overall project finances were capital investment, availability, and electricity tariff value. All other parameters, while important to a complete picture of a facility's financial potential, will not have the impact of these items.

These cases showed that a small industrial combined heat and power plant can be integrated economically into an industrial facility under the right circumstances. On a per unit of output basis, small plants are more expensive than larger ones because they do not take advantage of the economy of scale. However, industrial facilities, such as the one considered in this study, are developed over time as the facilities are expanded, and generally contain multiple power generating units to provide a more reliable source of power relative to a single large unit.

In addition, no credit has been taken in the economics for the significant reductions in SO_2 and NOx from these designs relative to conventional industrial coal boilers. Based on input from potential industrial users of these facilities, it was estimated that these credits could amount to as much as 1 M$/yr.

The design for Subtask 3.3 met the original financial goals set by the industrial facility at the beginning of the study when the environmental credits were added. These goals were to reduce premium fuel usage, increase the use of coal, reduce emissions, reduce purchased power and have a payback of less than ten years. However, during the course of the study, the industrial facility's position changed concerning the payback period from that of a utility to one more in-line with other industrial projects. Consequently, they have decided not to implement the project at this time.

The positive return (8.4%) for Subtask 3.3 shows that an industrial application for gasification with CHP can be feasible in certain situations depending on site specific criteria, such as environmental credits, electricity and steam prices, and project hurdle rates.

This section describes the base case design for a lignite-fueled IGCC power plant that was developed in Subtask 3.4. This design consists of a single oxygen-blown gasification train using Gas Technology Institute's fluidized bed U-GAS® gasifier producing sufficient syngas to fully load a single GE 7FB combustion turbine. The plant is fueled by North Dakota lignite and will be located at a generic North Dakota site. At design conditions, the plant consumes 2,320 metric tpd (2,558 short tpd) of moisture-free lignite and produces about 251 MW of export power. Sulfur and ash are the only byproducts. Unlike the air-blown case, this design produces a concentrated stream of CO_2, making CO_2 capture and sequestration more economic than a syngas stream diluted with nitrogen. This design represents a case focused on commercially proven technologies. The details of the design are contained in Appendix C.

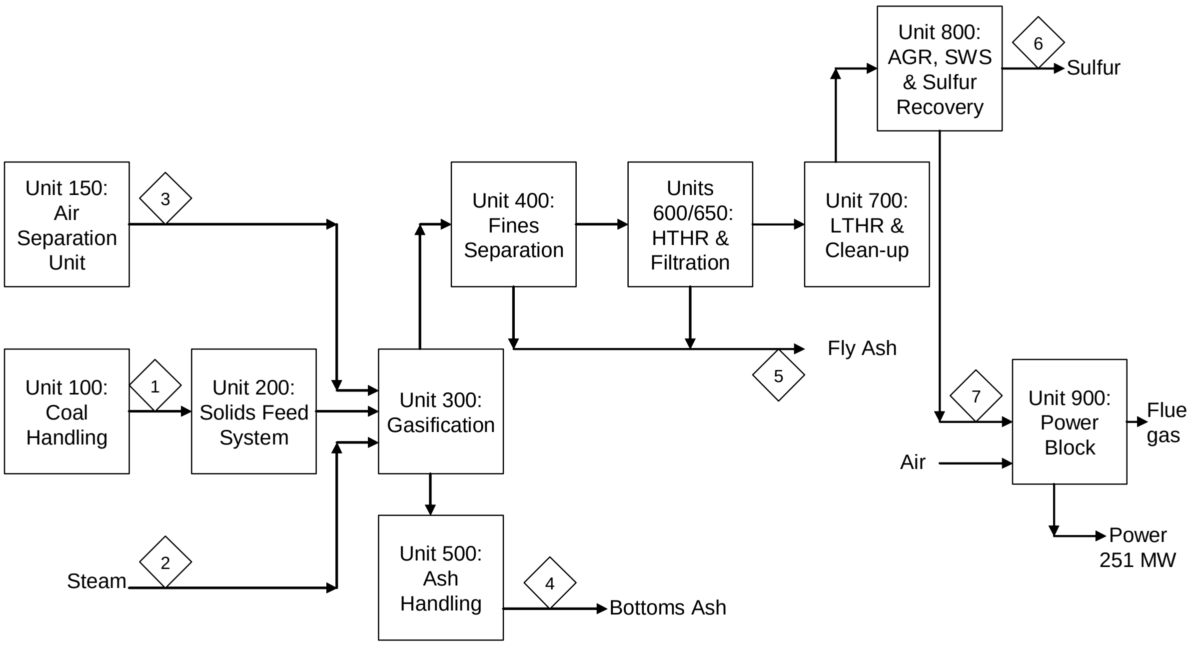
The design that is described in this section is the second of the two designs that were developed for the lignite-fueled IGCC power plant. The first preliminary design was rejected because it had an unacceptably low electrical efficiency. This electrical efficiency was low for several reasons. First, the lignite feed to the gasifiers was dried to 10% moisture content using syngas as the fuel for the driers. In the current design, it is dried to only 20% moisture, and the primary heat source for drying is low-level heat that otherwise would be rejected to the atmosphere either by air coolers or at the cooling tower. Secondly, the preliminary design used two smaller GE 6FA combustion turbines (~69 MW ISO conditions), which are significantly less efficient than the new GE 7FB turbine used in the new design. This turbine is currently under development, and is expected to produce about 211 MW of power from syngas. Finally, other changes were made to increase the efficiency of the internal plant heat exchange.

5.1 DESCRIPTION OF THE LIGNITE-FUELED IGCC POWER PLANT DESIGN

Figure 5.1 is a simplified block flow diagram and material balance of the lignite-fueled facility. The complete material balance is contained in Appendix C.

At design conditions, Unit 100, the coal handling, drying and sizing area, processes 3,425 metric tpd (3,775 short tpd) of wet lignite (32.24% moisture). Unit 100 is sized to process 3,868 metric tpd (4,263 short tpd) of wet lignite containing 40% moisture. Lignite is delivered to the site by rail. Unit 100 contains two 100% Heyl and Patterson crushers (one spare) and four 33.3% fluidized bed drier units (one spare). Each section includes a spare unit to account for periods of maintenance. The as-received lignite is crushed to a top size of 6.4 mm (¼ inch) with no more than 10% less than 100 mesh. The sized material is dried in the fluidized bed driers to 20% moisture by passing heated air through it. The air is heated in a series of heat exchangers using, in succession, cooling water return to the cooling tower, stripped water from the sour water stripper, sour water from the syngas scrubber, and finally by 0.34 MPa (50 psig) steam. The dried lignite is then transferred via a screw feeder to a 24-hour storage silo where it is kept under nitrogen to prevent spontaneous combustion until it is fed to one of the two gasifiers (one operating and one spare).

Figure 5.1 Simplified Block Flow Diagram and Material Balance



Stream Number	1	2	3	4	5	6	7
Description	**Coal to Gasifier**	**Steam to Gasifier**	**Oxidant to Gasifier**	**Bottom Ash**	**Fly Ash**	**Sulfur Product**	**Clean Syngas to Gas Turbine**
Pressure, MPa (psig)	0.10 (14.7)	3.45 (500)	3.45 (500)	0.10 (14.7)	0.10 (14.7)	NA	2.59 (375)
Temperature, K (F)	294 (70)	561 (550)	583 (590)	1144 (1600)	1144 (1600)	NA	316 (110)
Flow Rate, kg/s (lb/hr)	33.6 (266,450)	7.9 (62,968)	16.1 (127,462)	2.3 (18,399)	0.7 (5,329)	0.2 (1,577)	43.6 (345,865)

Air Products designed Unit 150, the cryogenic air separation unit with nitrogen compression. This unit delivers 1,388 metric tpd (1,530 short tpd) of a 95% pure oxygen stream (1,326 metric tpd of contained oxygen) at about 305K (90°F) and 3.45 MPa (500 psia). It also provides 4,498 metric tpd (4,958 short tpd) of a 99.3% purity nitrogen stream at 377K (219°F) and 3.45 MPa (500 psia). Unit 150 also contains storage for 227 metric tpd (250 short tpd) of nitrogen for internal use.

The gasification island (Units 200-500) consists of the coal lockhopper feed system, two gasifiers (one operating and one spare), startup heaters, dust cyclones, and ash and dust removal systems. The spare gasifier was justified based on a payout of 4.8 years and because reliable operation for a base load power generation facility is an important feature to an operator. Details of this trade-off study are provided in Appendix C. The dried coal is fed to one of the two gasifiers by a lockhopper system. Each gasifier has two 100% lockhopper systems to provide complete redundancy in case of an outage. Oxygen and saturated 3.45 MPa (500 psig) steam are mixed and fed to the gasifier. After passing through two cyclones, which return the removed particles back to the gasifier vessel, the syngas leaves the gasifier at about 1144K (1600°F).

The Unit 400 dust removal system removes fly ash from the bottom of the third-stage cyclones via a lockhopper system. The ash first enters a refractory lined lockhopper, and then is sent through a cooling screw to reduce the temperature to 533K (500°F). The cooled ash enters a surge hopper, where it is transported via a pneumatic system to a day tank from which it either can be disposed or sold. The Unit 500 bottom ash removal system functions in a similar manner to remove the bottom ash from the gasifier vessel. One area of improvement that should be considered is combining these systems as was done in Subtask 3.3 (see Appendix B).

Syngas leaving the third-stage cyclone goes to Unit 600, the high-temperature heat recovery unit, and then to Unit 650, syngas filtration. In Unit 600, the syngas is cooled to 616K (650°F) in a fired tube heat exchanger that generates 6.90 MPa (1,000 psig) saturated steam. The remaining particulates are removed from the syngas by metallic candle filters in Unit 650. The fly ash that is collected from the filters is sent to the second lockhopper in Unit 400 where it is mixed with the cooled fly ash from the third-stage cyclone.

In Unit 700, the syngas is cooled to about 533K (500°F) in a heat exchanger that generates 3.45 MPa (500 psig) steam. It then is cooled to about 401K (263°F) in an impingement scrubber column, which also removes any residual particulates from the syngas. The syngas then is reheated by a few degrees to prevent any condensation in the COS hydrolysis reactor where the COS is converted to H_2S. Finally, the syngas is cooled to 316K (110°F) in a series of heat exchangers before cleanup.

Unit 800 cleans the syngas. Mercury is removed by adsorption on sulfur-impregnated carbon. H_2S is removed from the syngas by scrubbing with an amine solution. The cleaned syngas is then sent to the power block. Sulfur is recovered in a Claus system followed by a SCOT off-gas treating process to minimize sulfur emissions. Other sulfur removal and recovery processes (e.g., CrystaSulf) were considered and are discussed in Appendix C.

Unit 900, the power block consists of a single GE 7FB combustion turbine followed by a heat recovery steam generator (HRSG). Before entering the turbine, the syngas is heated with hot BFW. The combustion turbine produces about 211 MW of power. The HRSG produces 6.90 MPa (1,000 psig) superheated steam, 3.45 MPa (500 psig) superheated steam, and 50 psig steam for internal plant use. The 6.90 MPa (1,000 psig) and 3.45 MPa (500 psig) superheated steams are sent to a two-pressure steam turbine that generates about 91 MW of electrical power. The facility has an internal parasitic power consumption of about 51 MW, reducing the net export power production from the facility to 251 MW. The electrical efficiency of this liqnite-fueled IGCC power plant is 36.5% on an HHV basis.

The estimated EPC cost of the grass roots facility (including the spare gasification train) is 410.5 M$ (second quarter 2004 dollars), or about 1,635 $/kW of design export power.

Table 5.1 summarizes the major input and output streams along with some key operating parameters of the Subtask 3.4 Lignite-Fueled IGCC Power Plant. This design represents a case focused on commercially proven technologies. Additional analysis should be performed to consider other technologies and design improvements that may be able to reduce the plant cost further.

Table 5.1 Overall Plant Summary of the Subtask 3.4 Lignite-Fueled IGCC Power Plant

	Lignite-Fueled IGCC Power Plant
Design Inputs	
Lignite Feed, moisture-free metric tpd (short tpd)	2,320 (2,558)
Lignite Feed, moisture-free kg/s (lb/hr)	26.858 (213,160)
Fuel (Natural Gas), MW (MBtu/hr)*	2.6 (8.93)
Makeup Water, m^3/s (gpm)	0.121 (1,920)
Design Outputs	
Export Power, MW	251.0
Sulfur, kg/s (lb/hr)	0.199 (1,557)
Ash, kg/s (lb/hr)	2.654 (23,729)
EPC Cost, M$**	410.5
Plant EPC Cost, $/kW	1,635
Plant Energy Input, k$/MW (k$/MBtu/hr)	596.1 (174.7)
Plant Energy Output, k$/MW (k$/MBtu/hr)	1,631.7 (478.2)
Cold Gas Efficiency, % (HHV basis)	84.0
Net Electrical Efficiency, % (HHV basis)***	36.5

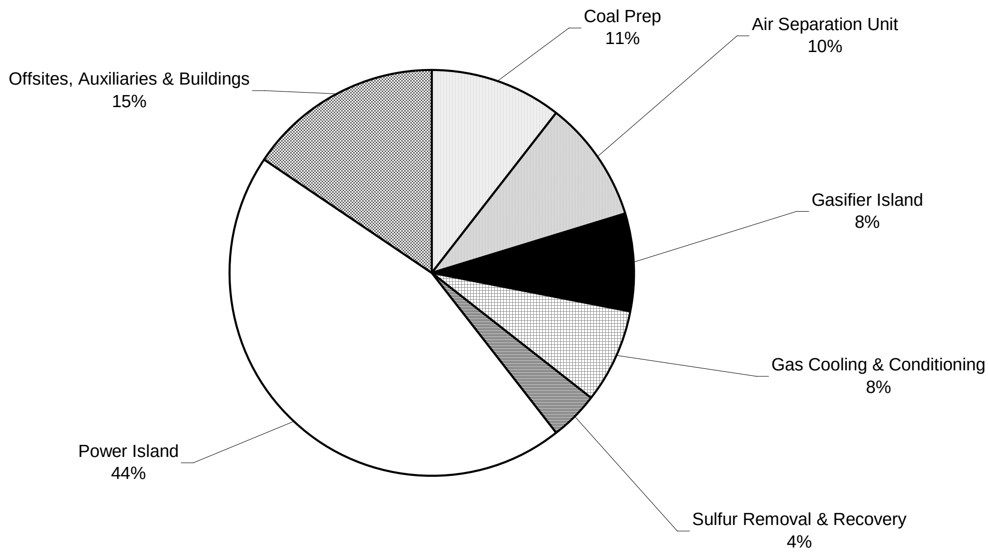
* This small amount of natural gas is used in the Claus plant, and since this usage is so small, it is being considered as an operating cost under the catalyst and chemical classification and not as a plant feed.

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

*** Net electrical efficiency is defined as the net electrical energy divided by the total heating value energy of the input fuel

Figure 5.2 presents the EPC cost distribution of the Lignite-Fueled IGCC Plant (including a second spare gasifier). The EPC cost of the facility is 410.5 M$ on a second quarter 2004 basis. The investment is adjusted for labor rates and productivity in North Dakota.

Figure 5.2 Capital Cost for Lignite-Fueled IGCC Power Plant
(410.5 million US$)



5.2 FINANCIAL RESULTS FOR THE IGCC POWER PLANT DESIGN

The lignite-fueled IGCC power plant has an expected return on investment (ROI) of 19.4%, with a net present value (NPV) of 175.6 M$ at a 10% discount rate over a 20-year project life based on an expected power selling price of 6.08 cents/kWh. As expected, Subtasks 3.2 and 3.3, the nominal 25 MW subbituminous industrial gasification facilities, have higher installed costs (2,700-3,100 $/kW) because of the economy of scale disadvantage. However, studies of larger IGCC designs (450 MW) have been able to capture even greater economy of scale benefits, with installed costs of 1,300 to 1,650 $/kW[1]. The installed cost of this case is approaching the cost of the large IGCC facilities by taking advantage of a greater economy of scale. The results point to the possibility that a larger design may be able to reduce installed costs further.

Table 5.2 outlines the rate of return (ROI), NPV, payback year, and required electricity selling price to obtain a 12% ROI with all other entries fixed. The ash and sulfur produced in the plant accounts for all additional revenue beyond electricity tariffs.

The investment cost accuracy for this study ranges is -15% to +30%. Taking this range into account, the likely ROI for this facility would be between 23.6% to 12.8 %.

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

Table 5.2 Financial Cost Summary for Lignite-Fueled IGCC Power Plant

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

The results reported in Table 5.2 do not include any credits for the environmental benefits gained from the use of IGCC technology. In order to properly compare this design versus other power generation technologies using lignite, a life cycle analysis also should be performed. Quantification of the environmental differences will provide a more level playing field by which alternate technologies can be evaluated. A project developer must consider alternative compliance costs to meet new emission rules versus the cost of the IGCC plant.

Figure 5.3 shows the impacts of selected variables on the NPV at a discount rate of 10%. In all of the cases, the input parameter is varied by ±10%, and the NPV changes from the base case are shown. 10% changes were used to give a common ground by which all variables were evaluated. However, the range of realistic possibilities for each variable could differ significantly. For example, 10% changes in the availability or income tax rate 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 electricity tariff has the greatest impact on the plant net present value; increasing it by 10% increases the net present value by more than 60 M$. In this case, “electricity tariff” refers to the sales price of the electricity that the plant generates. This variable also was the most significant in Subtasks 3.2 and 3.3. The significance is more pronounced in this design since, unlike Subtasks 3.2 and 3.3, there is no steam export. Also very significant is the availability (annual average on-stream time). By reducing the availability by 10%, the net present value is reduced by more than 45 M$. All other variables associated with the amount of time the plant is operating (availability, operating hours, and plant life) also have a significant impact on plant economics.

The remainder of the input variables impacted the plant economics to a significantly lower extent. It is interesting to note that the interest rate, amount of debt financing, and the plant fixed O&M cost have a greater impact on the economics than in Subtasks 3.2 and 3.3. This is due in large part to the higher EPC cost of Subtask 3.4. Changes in these variables will impact the early cash flow to a greater extent than in the industrial gasification case. Income tax rate also has a greater impact than in Subtasks 3.2 and 3.3 due to the positive cash flows throughout the operational life of the project. Coal prices could change fairly significantly without changing the overall economics to a great extent. If the coal price is increased to 13.23 $/metric ton (a nearly 30% rise), the NPV is only decreased by 15.3 M$, a 0.8% change in the return on investment.

Figure 5.3 Comparisons of a +/-10% change in selected inputs on Project NPV
(Discount Rate = 10%)

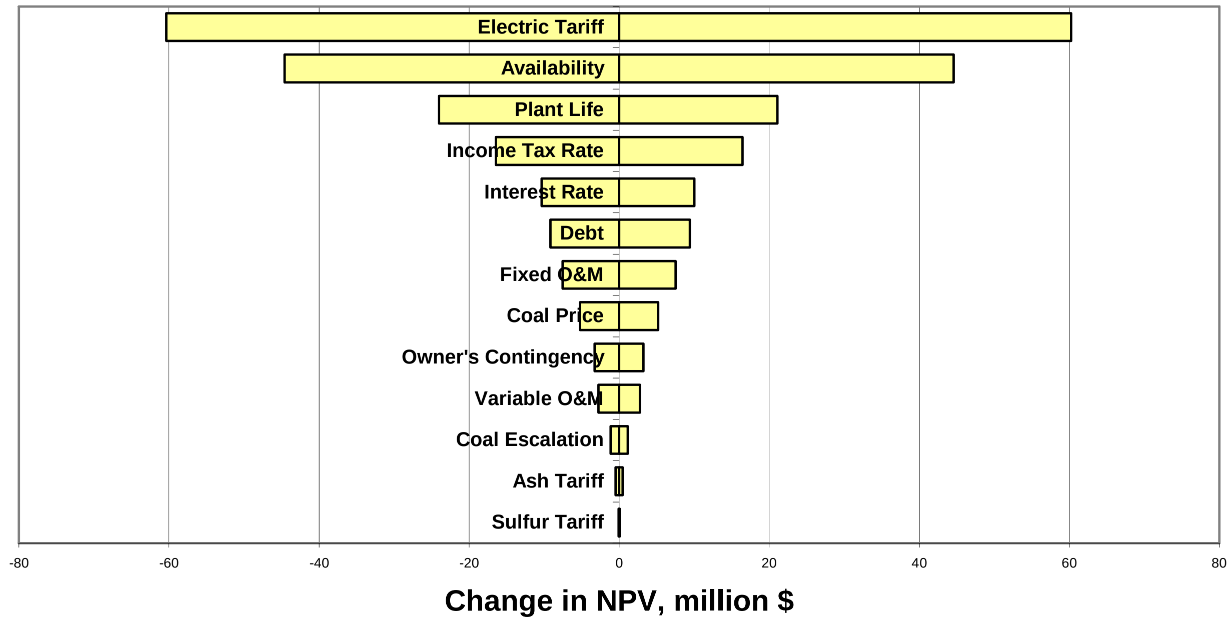


Figure 5.4 shows the relationship between the electricity tariff and ROI assuming a 10% discount rate. The plant relies most heavily on the electricity tariff for the economic outcome due to electricity being the main product. Even with the relatively low electricity prices that exist in North Dakota, the plant still demonstrates positive economics. If the electric price used for upstate New York for Subtasks 3.2 and 3.3 of nearly 8 cents/kWh were applied here, the plant would have a return of over 27%. Regardless of the tariff value assumed, any electricity market could obtain positive returns with this facility, all other plant inputs being equal.

Figure 5.4 Effect of Electricity Tariff on Investment Return

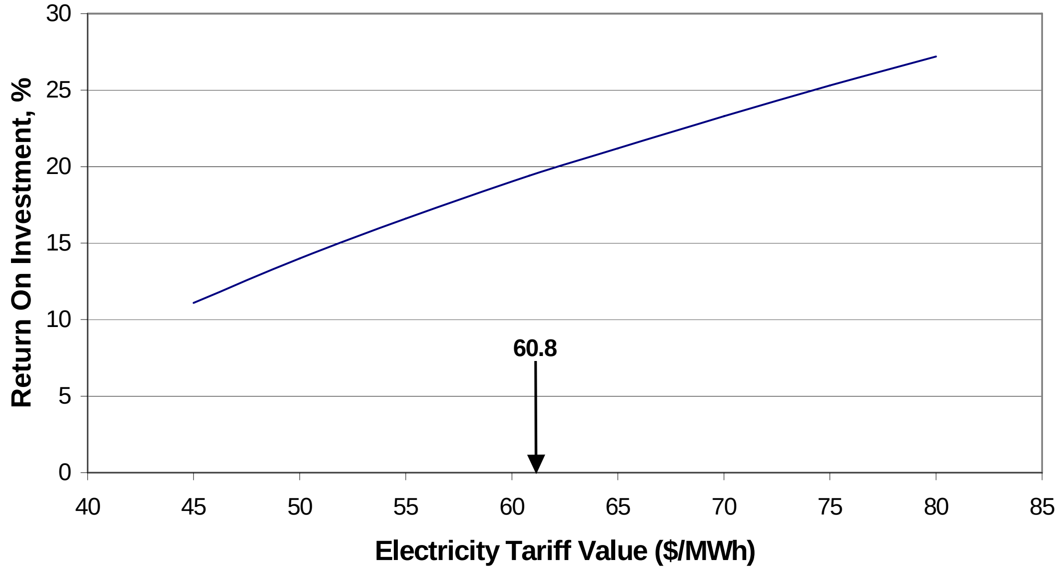
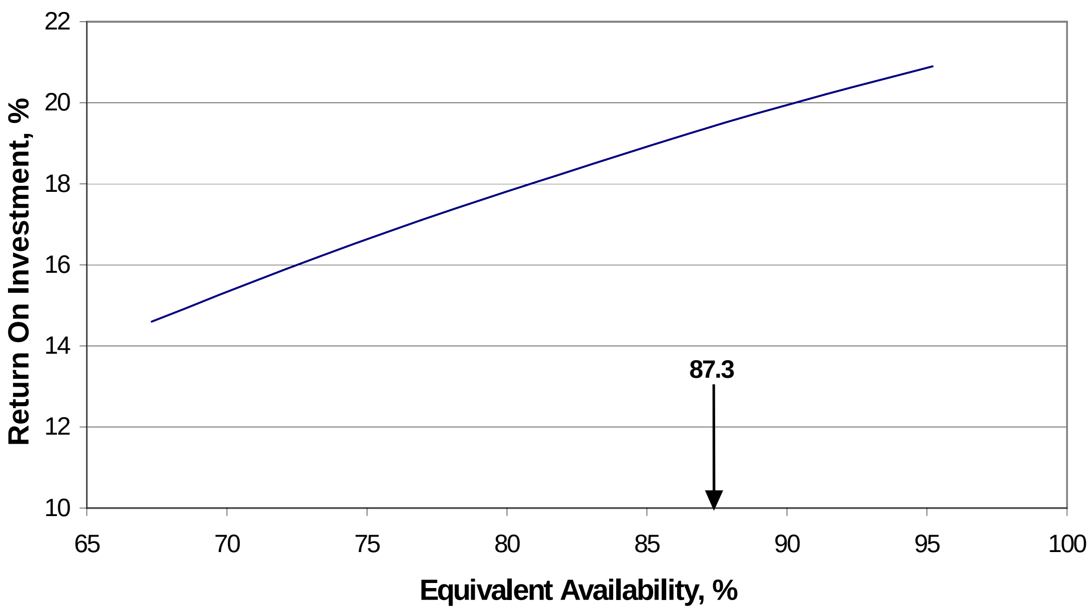


Figure 5.5 shows the relationship that varying the equivalent availability (annual average on-stream time) has on the ROI.

Figure 5.5 Effect of Availability on Investment Return

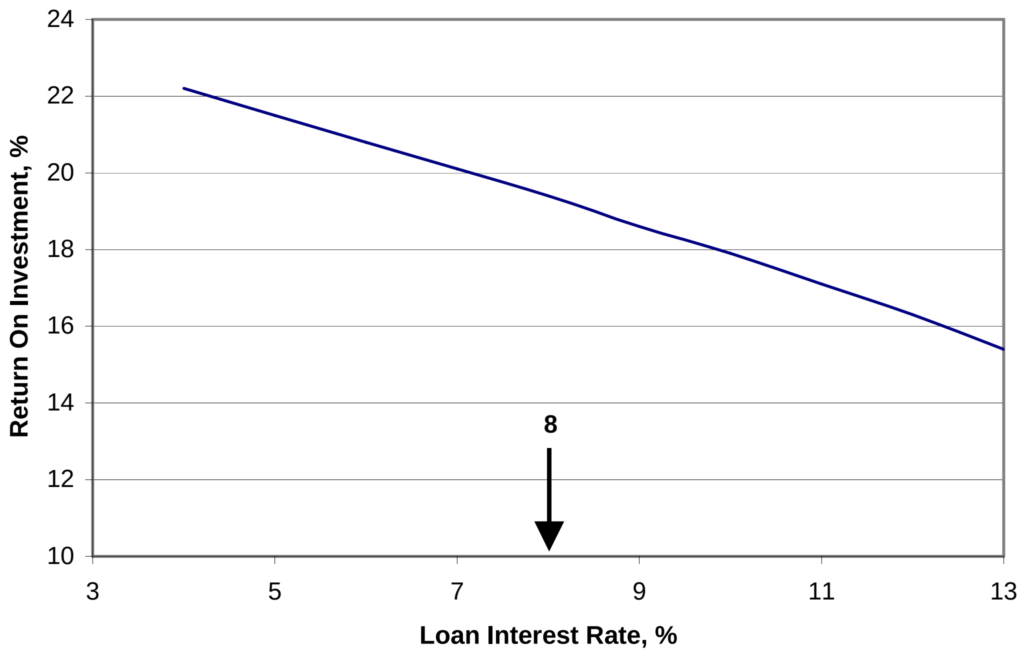


The impact that availability has on plant economics comes as little surprise. Reliable operation is very important to assure that the cost of project development and construction can be recovered. Long downtimes throughout the life of the project will significantly hurt overall project economics given a 20-year project life. The impact of availability on overall plant economics is similar to that of Subtasks 3.2 and 3.3. Since a plant of this size should be built to have a lifetime longer than the 20 years used in the model, greater consideration of plant life should be made during the project development phases. Figure 5.6 makes this point more clearly. A certain economic life is required in order to pay off the debt incurred during project construction. Once this debt has been paid and construction costs recouped, the steady cash flow will lead to a stable rate of return.

The interest rate for debt financing plays a larger role in this case than in Subtasks 3.2 and 3.3. Interest payments will be significantly higher than in the industrial gasification case. However, interest rate variations do not have a relatively greater significance than either availability or electricity price, as shown in Figure 5.7.

Figure 5.6 Effect of Plant Life on Investment Return

Figure 5.7 Effect of Interest Rate on Investment Return



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

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

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

5.3 SUMMARY OF THE SUBTASK 3.4 LIGNITE-FUELED IGCC POWER PLANT DESIGN

A preliminary financially viable design for a lignite-fueled IGCC power plant has been developed. This design consists of a single oxygen-blown gasification train using Gas Technology Institute's fluidized bed U-GAS® gasifier producing sufficient syngas to fully load a single GE 7FB combustion turbine. An oxygen-blown design was selected to allow for the possible future application of CO_2 capture and sequestration. It is believed that a concentrated stream of CO_2 will make CO_2 capture and sequestration more economic than that for a stream diluted with nitrogen as in an air-blown design. The plant is fueled by North Dakota lignite and will be located at an unspecified generic North Dakota site. At design conditions, the plant consumes 2,320 metric tpd (2,558 short tpd) of moisture-free lignite and produces about 251 MW of export power. Sulfur and ash are the only byproducts.

In this design, the lignite is dried from 32.2% moisture to 20% moisture before being fed to the gasifier. Waste low-level heat that otherwise would be rejected to the atmosphere either by air coolers or at the cooling tower is the heat source for drying the lignite. Use of this drying technique allows the plant to have an electrical efficiency to power of about 36.5% on a HHV basis. The estimated cost of the facility is 410 M$ or about 1,635 $/kW of export power. Based on an expected power selling price of 6.08 cents/kWh, the plant has an expected return on investment of 19.4%, with a net present value (NPV) of 175.6 M$ at a 10% discount rate over a 20-year project life.

As expected, Subtasks 3.2 and 3.3, the 25 MW subbituminous industrial gasification facilities, have higher installed costs (2,700-3,100 $/kW) because of the economy of scale disadvantage. However, studies of larger IGCC designs (450 MW) have been able to capture even greater economy of scale benefits, with installed costs of 1,300 to 1,650 $/kW[2]. The installed cost of this case is approaching the cost of the large IGCC facilities by taking advantage of a greater economy of scale. The results point to the possibility that a larger design may be able to reduce installed costs further.

Results of a sensitivity analysis point to capital investment, availability, and electricity tariff as the most sensitive parameters.

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

6.1 SUMMARY

Conceptual designs were developed for several coal gasification facilities using Gas Technology Institute's U-GAS® fluidized bed gasifier. Subtask 3.2 developed two base case designs for industrial combined heat and power facilities (CHP) fed with Southeastern Ohio coal and that will be located at an upstate New York location. One base case design used an air-blown gasifier, and the other used an oxygen-blown gasifier. When this project started, it was unclear as to which mode of gasifier operation would be more economic for a CHP application. The evaluation showed that the air-blown case was less costly and more efficient. Subtask 3.3 developed an advanced design for an air-blown gasification combined heat and power facility based on the Subtask 3.2 air-blown case to reduce its cost and improve its return on investment. Subtask 3.4 developed a base case design for a large lignite-fueled IGCC power plant that uses a GE 7FB combustion turbine that will be located at a generic North Dakota site.

The Subtask 3.2 air-blown industrial CHP facility uses two parallel GTI fluidized bed gasifiers to power two GE 10 combustion turbines. Twin gasifiers and turbines were used at the request of the industrial facility to provide a wide range of operating flexibility. The plant consumes 313.6 metric tpd (345.7 short tpd) of moisture-free coal, while exporting 21.7 MW of power and 12.8 kg/s (101.7 klb/hr) of steam (561K/2,76 MPa, 550°F/400 psig) to the industrial facility. Small amounts of byproduct sulfur and ash also are produced. The plant has a net CHP efficacy of 49.0% (HHV) which includes the energy exported in the steam. It requires a capital investment of 90 M$ (million US dollars, 2nd quarter 2004). Based on the marginal values for power purchases and steam production, the plant has an expected return on investment of 5.9%.

Subtask 3.2 also investigated the economics and performance of an oxygen-blown facility with a similar coal feed rate. The oxygen-blown unit also uses two parallel GTI fluidized bed gasifiers to power two GE 10 combustion turbines. The plant consumes 293.7 metric tpd (323.8 short tpd) of moisture-free coal, while exporting 23.3 MW of power, but only 3.4 kg/s (26.8 klb/hr) of steam (561K/2,76 MPa, 550°F/400 psig) to the industrial facility. The lower steam export rate is largely due to the internal consumption by the oxygen facility. Slightly smaller amounts of byproduct sulfur and ash also are produced. The plant has a net CHP efficacy of 29.1% (HHV), which includes the energy exported in the steam. It requires a capital investment of 100 M$ (2nd quarter 2004). Based on the marginal values for power purchases and steam production, the plant has a negative return on investment. Consequently, this oxygen-blown case was eliminated from further consideration.

The Subtask 3.3 alternate air-blown industrial combined heat and power facility uses a single GTI fluidized bed gasifier to power two GE 10 combustion turbines. This case is an improved version of the original Subtask 3.2 Air-Blown Design. The plant consumes

313.6 metric tpd (345.7 short tpd) of moisture-free coal and exports 21.3 MW of power and 13.3 kg/s (105.3 klb/hr) of steam (561K/2,76 MPa, 550°F/400 psig), similar to the basis for the air-blown case in Subtask 3.2. Compared to Subtask 3.2, the plant has a higher net CHP efficacy (49.7%, HHV) and a lower capital investment of 82 M$, (2nd quarter 2004). This decrease in cost is partially due to the use of advanced technologies (e.g., Stamet solids pump, advanced gas clean-up systems, and more integrated heat recovery). Other advantages include economy of scale, lower capital investment, an improved sulfur recovery system, and improved heat integration. The plant has an expected return on investment 8.4%.

Compared to the original Subtask 3.2 air-blown base case, the Subtask 3.3 alternate case exports about 2% less power and 3.5% more steam to the industrial facility, costs about 9% less, has a higher net CHP efficacy, and has a higher return on investment.

The Subtask 3.4 oxygen-blown lignite-fueled IGCC power plant uses two GTI fluidized bed gasifiers (one operating and one spare) to power one GE 7FB combustion turbine. An oxygen-blown design was selected for this application to allow for the possible future capture and sequestration of CO_2. The plant consumes 2,320 metric tpd (2,558 short tpd) of moisture-free coal and exports 251 MW of power. It also produces 0.2 kg/s (1,577 lb/hr) of byproduct sulfur and 2.65 kg/s (21,063 lb/hr) of ash. The lignite is dried from 32.2% moisture to 20% moisture before being fed to the gasifier. Waste low-level heat that otherwise would be rejected to the atmosphere either by air coolers or at the cooling tower is the primary heat source for drying the lignite. Use of this drying technique allows the plant to have an electrical efficiency to power of about 36.5% on a HHV basis. The estimated cost of the facility is 410 M$ (2nd quarter 2004) or about 1,635 $/kW of export power. Based on the average electricity tariff in North Dakota, the plant has an expected return on investment of 19.4%, with a net present value (NPV) of 175.6 M$ at a 10% discount rate over a 20 year project life. “Electricity tariff” here refers to the sales value for the electricity that the plant generates.

6.2 CONCLUSIONS

This study has shown that the cost for electric power and steam produced from an industrial size CHP gasification facility can be competitive with energy generated from premium fuels or purchased power at retail rates.

The objective of Subtask 3.2 was to develop conceptual base case designs that would provide reliable long-term operation. These designs were developed using proven subsystems at the expense of efficiency, with limited heat integration to promote operability and reliability; this decision was based on the premise that reliability of the utility system in an industrial facility is a major requirement.

Subtask 3.3 developed an alternate air-blown design that improved the Subtask 3.2 air-blown base case design. It has a higher net CHP efficacy, is less costly, and has a higher return on investment. This design uses the Stamet solids pump, (currently under development for this type of service), metallic filters to remove particulates from the syngas, a LO-CAT® sulfur recovery system, and contains improved heat integration. Further improvements still are possible. New technologies are being developed that are expected to further reduce plant costs and improve operations. These technologies should be investigated before a design can be finalized.

Subtask 3.4 developed a financially viable design for a power only plant operating with a gasification combined cycle. This study shows that an IGCC plant employing the U-GAS® gasification technology can be cost competitive with other means of power generation. This plant uses waste heat that normally would be rejected to the atmosphere to dry the lignite from 32.2% to 20% moisture. Future optimization of the plant should further reduce the cost and improve the economics. Besides cost reduction, one area that optimization could significantly improve is the availability.

In all cases, the results of sensitivity analyses show that capital investment, availability, and electrical tariff are the most sensitive financial parameters that are under the control of the plant designer.

Furthermore, a list of potential enhancements has been identified that should provide additional cost savings as some of the improvements are researched, developed and implemented. These are:

- Improved sulfur removal methods including warm sulfur removal
- Warm mercury removal systems
- Improved particulate removal systems to reduce capital costs and improve efficiency
- Improved fuel drying methods

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

- Developing improved sulfur and mercury removal technologies
- 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 811K, 1000°F)

- Determining the combustion turbine output and emissions capabilities on syngas
- Determining the characteristics of the ash associated with the char

The Subtask 3.3 and Subtask 3.4 designs should meet the emissions targets established by the DOE in their roadmap for 2010 for SOx, NOx, particulates, and mercury.

Another objective of Task 3 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 designs.

6.3 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 higher temperature contaminant removal technologies. One example involves the integration of the gas turbine and ASU, which could both reduce compression costs 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 422K, 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

- The optimum moisture content of the gasifier feed for solids-fed gasifiers needs to be established. The energy required to dry the gasifier feed must be compared against the energy required to evaporate the moisture in the gasifier (and its subsequent recovery in the downstream cooling system) to determine the optimum moisture content of the feed.

- The physical characteristics and properties of the coal must be studied further in order to better predict gasification system performance. These include:
    - o Characterization of the crushing and drying performance of the design coal feedstock
    - o Determination of the gasification reactivity of the desired feedstock.
    - o Determination of the ash characteristics associated with the char
    - o Characterization of the particulate properties
    - o 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 811K, 1000°F).
    - o During a visit to a gasification facility in China it was noted that at temperatures above 811K (1000°F) the cyclone efficiency drops off sharply. Emtrol, a domestic company that is a world leader in cyclone design, confirmed this loss of efficiency.

The pertinent references to previous work are in footnotes at appropriate places in this report. However, they are repeated here, along with other related references, in alphabetical order for convenience of the reader. References to the original reports for the Task 3 subtasks are not shown since updated versions of these reports are included as appendices to this report.

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°C	degrees Celsius
°F	degrees Fahrenheit
$	United Stated dollars
$/kW	United States dollars per kilowatt
¢/kWh	United States cents per kilowatt
$/kscf	United States dollars per thousand standard cubic feet
$/MBtu	United States dollars per million British thermal units
$/MW-hr	United States dollars per megawatt hour
$/ton	United States dollars per short ton
%	percent
%/yr	percent per year
AEP	American Electric Power
AGFS	Acid Gas removal - Formulated Solvents
AGR	acid gas removal
ASU	air separation unit
atm	atmosphere(s)
BACT	best available control technology
BFW	boiler feed water
BHP	brake horse power
BOP	balance of plant
Btu	British thermal unit(s)
Btu/scf	British thermal units per standard cubic foot
CEM	continuous emission monitoring
Cu ft	cubic feet
CHP	combined heat and power
CH_4	methane
CO	carbon monoxide
CO_2	carbon dioxide
COS	carbonyl sulfide
CT	combustion turbine
CW	cooling water
d	day
DCF	discounted cash flow
DCS	distributed digital control system
DGA	Diglycolamine
DLE	dry low emissions
DOE	Department of Energy
EDC	IGT's Energy Development Center

E-GAS™	name of the gasification technology at Wabash River
EIA	Energy Information Administration
EPA	Environmental Protection Agency
EPC	engineering, procurement and construction
EPRI	Electric Power Research Institute
FOAK	first-of-a-kind
fpm	feet per minute
F-T	Fischer-Tropsch
ft, ft^2, ft^3	foot (feet), square feet, cubic feet
FT	fluid temperature
gal	gallon(s)
GE	General Electric
gpm	gallons per minute
GT	gas turbine
GTC	Gasification Technologies Council
GTG	gas turbine generator
GTI	Gas Technology Institute
H_2O	water
H_2S	hydrogen sulfide
Hg	mercury
HHV	higher heating value
HP	high pressure
hp	horse power
hr	hour(s)
HRSG	heat recovery steam generator
HT	hemispherical temperature
HTHR	high temperature heat recovery unit
HV	high voltage
IGCC	integrated gasification combined-cycle
in, in^2, in^3	inches, square inches, cubic inches
IGT	Institute of Gas Technology
IP	intermediate pressure
IRR	internal rate of return
ISBL	inside battery limits
ISO	International Organization for Standardization (conditions: 15°C, 60% relative humidity, at sea level altitude)
IT	initial deformation temperature
K	kelvin
kg	kilogram
kg/s	kilogram per second

klb	thousands of pounds
klb/hr	thousands of pounds per hour
KO	knock out
kscf	thousands of standard cubic feet
kscf/hr	thousands of standard cubic feet per hour
kW	kilowatt
kW-hr	kilowatt-hour
kV	kilovolt
lb	pound(s)
lbmol	pound mole(s)
lb/hr	pounds per hour
lb/MBtu	pounds per million British thermal units
lb/MW-hr	pounds per megawatt hour
LP	low pressure
L/V	liquid/vapor
LHV	lower heating value
LTHR	low temperature heat recovery
m, m^2, m^3	meter, square meter, cubic meter
M$	thousands of United States dollars
MCC	motor control center
MDEA	methyldiethylamine
MDT	mean down time
min	minute(s)
Mlb	millions of pounds
M	million(s)
M$	millions of United States dollars
MBtu	millions of British thermal units
MP	medium pressure
MTBF	mean time between failures
MW	megawatts
MW-hr	megawatt-hours
NETL	National Energy Technology Laboratory
NFPA	National Fire Protection Association
NGCC	natural gas combined cycle
NH_3	ammonia
Nm^3	normal meters cubed
NOx	nitrogen oxides
NPV	net present value
O&M	operating and maintenance

OSBL	outside battery limits
P&IDs	piping and instrument drawing
PC	Pulverized Coal
PDQ$®	Price and Delivery Quoting Service for Chemical Process Equipment
PFD	process flow diagram
pH	a measure of acidity
ppm	parts per million
ppmv	parts per million by volume
ppmw	parts per million by weight
ppmvd	parts per million by volume dry
PSA	pressure swing adsorption
psi	pounds per square inch
psia	pounds per square inch absolute
psig	pounds per square inch gauge
ROI	return on investment
ROM	run of mine
RPG	Raymond Professional Group
S/C	subcontract
scf	standard cubic foot (feet) at 60ºF and 1 atmosphere
scfm	standard cubic feet per minute
scfh	standard cubic feet per hour
SCOHS	selective catalytic oxidation of hydrogen sulfide
SCR	selective catalytic reduction
SCOT	Shell Claus Off-gas Treating
SH	superheated
SO_2	sulfur dioxide
SOx	sulfur oxides
ST	steam turbine
ST	softening temperature
STG	steam turbine generator
SRU	sulfur recovery unit
SWS	sour water stripper
TBtu	trillions of British thermal units
tpd	short tons per day
tph	short tons per hour
TTO	tailgas thermal oxidation
U-GAS®	name of the GTI gasification technology
USGC	United States Gulf Coast
VIPs	value improving practices

VOC	volatile organic compounds
vol	volume
wt	weight
WWT	waste water treatment
yr	year
ZLD	zero liquid discharge
ZnO	zinc oxide
ZnS	zinc sulfide
ΔT	temperature difference
µg	micro gram

Appendix A

Task 3 Gasification Alternatives for Industrial Applications

DOE Contract No
DE-AC26-99FT40342
Subtask 3.2 Preliminary Design for Eastern Coal Case

Prepared For:

United States Department of Energy / National Energy Technology Laboratory

In association with:

Task 3 Gasification Alternatives for Industrial Applications

Subtask 3.2
Preliminary Design for Eastern Coal Case

DOE Contract
No. DE-AC26-99FT40342

August 2004

Prepared For:

United States Department of Energy / National Energy Technology Laboratory

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There are many industrial facilities in the United States that generate a significant portion, if not all, of their internal power consumption from coal. At times, some of these facilities also export power to the grid helping to stabilize the power grid. During the next few years, these facilities are facing more restrictive environmental regulations. As a result, many of these facilities will switch their fuel source from coal to either natural gas or oil and experience higher and more volatile fuel prices. This document is the first in a series of three topical reports that explore coal gasification alternatives for industry. These options will provide for a broader examination for determining long term industrial strategies for meeting energy requirements and complying with future emission constraints.

Examination of national energy use data for industry reveals that the use of coal to produce electricity and steam at industrial facilities has dropped by about 50% over the past 30 years (re. Figure 1.1).[1] This reduction has been countered by an increased reliance on natural gas during the 1990s. During the past few years, the higher prices for natural gas have resulted in a decline in total energy use by industry as a whole. Across the country there are over 1,000 industrial facilities that currently use coal. As discussed in the body of the report key drivers for fuel choice are price and emission issues. As the price for natural gas increases with inflation (4%) compared with coal (2%) over the coming years (see Figure 1.2) the need for coal based technologies that comply with emission criteria will be more attractive from an economic perspective.[2] The capability to use the existing coal based infrastructure to increase the use of coal using clean coal technologies can be readily absorbed by industry and conserve substantial quantities of natural gas for other uses in the economy.

This first subtask was developed with input from a large industrial complex in upstate New York. Two conceptual designs are prepared for industrial-scale IGCC (Integrated Gasification Combined Cycle) power plants using southeast Ohio coal to supply both electrical power and steam to the industrial facility. A true IGCC power plant combines gasification with a combustion turbine, heat recovery steam generator (HRSG), and a steam turbine. This is a modified IGCC plant in the sense that there is no steam turbine so that the excess steam is exported to the industrial facility (to be used in existing steam turbines to provide additional electricity); i.e., the combined heat and power (CHP) concept.

The conceptual designs for the facility uses the U-GAS® fluidized bed gasifier coupled to two gas turbines providing a 20-25 MW CHP facility. The U-GAS® technology is provided by Gas Technology Institute (GTI) located in Des Plaines, Illinois. The gas turbines used for the study are the GE 10 engines provided by General Electric. GTI developed the design and cost information for the gasification block, and Nexant developed that information for the remainder of the facility.

[1] EIA, July 2004 Monthly Energy Review

[2] Energy Review 2002, DOE/EIA-0384(2002) (Washington, DC, October 2003)

Figure 1.1 Historical Fuel Consumption

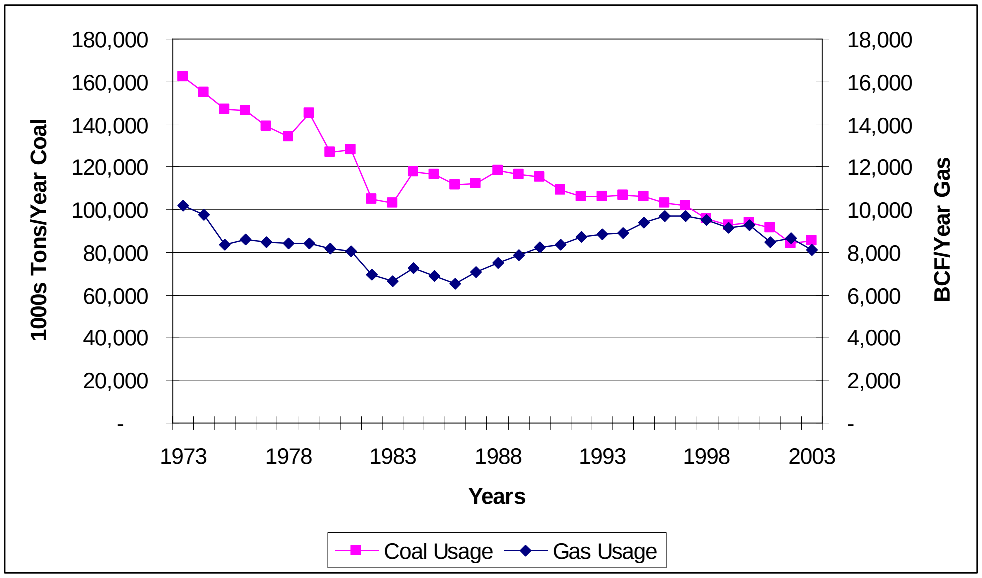
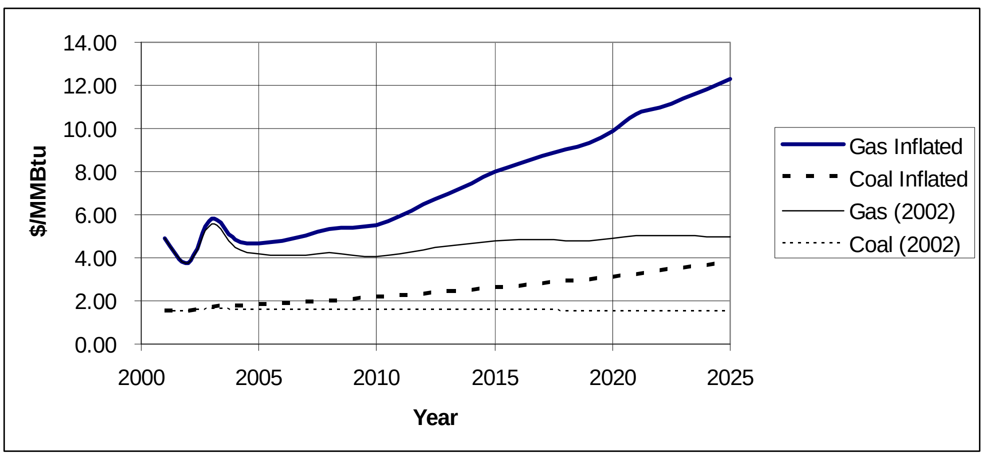


Figure 1.2 Forecast Fuel Prices



The objective of this subtask (Subtask 3.2) was to develop conceptual base case designs that would provide reliable, long-term operation. This is because this concept for an industrial-scale facility has not been demonstrated in a sustained long-term performance at the commercial level. Thus, these designs were 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. Furthermore, although not needed to provide the required capacity, two parallel, half-size gasification trains are used to provide the operating flexibility requested by the industrial client and typical of most industrial facilities. Subtask 3.3 will optimize the design for the Southeast Ohio coal case. Subtask 3.4 will develop a base case design for North Dakota lignite.

Two conceptual designs were developed; one is based on an air-blown gasifier, and the other uses an oxygen-blown gasifier. Table 1.1 summarizes these two designs.

Table 1.1 Overall Plant Summary

	Air-Blown Case	Oxygen-Blown Case
Design Inputs		
Coal Feed, moisture-free tpd	345.7	323.8
Coal Feed, moisture-free lb/hr	28,810	26,980
Fuel (Natural Gas), MBtu/hr	5.1	7.3
Makeup Water Input from the Industrial Facility		
Boiler Feed Water, gpm	495	473
Quench Water, gpm	30	70
Cooling Tower Makeup Water, gpm	53	72
Design Outputs		
Export Power, MW	21.7	23.3
Export Steam (400 psig, 550°F), Mlb/hr	101.72	26.75
Sulfur, lb/hr	899	863
Ash, lb/hr	2,097	1,465
Condensate (to industrial facility), Mlb/hr	60.9	65.5
EPC Cost, M$*	90.0	100.2
Plant EPC Cost, $/kW**	3,090	4,057
Plant Energy Input, k$/MBtu/hr	229.9	263.6
Plant Energy Output, k$/MBtu/hr	469.2	907.3
Equivalent Availability, %	85.7	82.6
Return on Investment, %***	5.9	<0
Cold Gas Efficiency, % (HHV basis)	79.3	83.1
Net CHP Efficacy, % (HHV basis)	49.0	29.1

* 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

The air-blown design exports about 1.6 MW less power. However, it is less costly because it does not require the costly air separation unit (ASU or oxygen plant) even though it requires larger processing equipment to handle the inert nitrogen that is contained in the syngas. Although the oxygen-blown case has smaller processing equipment, it exports less steam than the air-blown case because it consumes more steam to control the gasifier temperature, and consequently, requires more water (both boiler feed water and quench water). In the oxygen-blown case, steam is also used to reduce the NOx emissions from the gas turbine. Since the costs of the processing factors tend to compensate, the investment cost difference between the two designs is about that of the cost of the air separation unit.

For an air-blown facility with EPC (engineering, procurement and construction) costs of 90.0 M$ and a project life of 20 years, the return on investment (ROI) is expected to be 5.9%, with a net present value (NPV) of -14.6 M$ given a 10% discount factor. Table 1.2 outlines the rate of return, NPV, payback year, and required electricity and steam selling prices to obtain a 12% ROI with other entries fixed. It is important to keep in mind that there are two major products from this facility, electricity and steam, and both values must be considered when determining the suitability of this project. Besides the base case, a "high" and "low" estimate is listed reflecting the current cost accuracy assumption of +30/-15%.

Table 1.2 Air-Blown Financial Cost Summary

Cases

	Base	**Low -15% EPC**	**High +30% EPC**
ROI (%)*	5.9	10.7	2.5
NPV @ 10% Discount Rate (M$),	-14.6	2.26	-33.3
Number of Years to Payback	17	14	>20
Electricity Selling Price for 12% ROI (cents/kWh)**	9.02	8.4	11.6
Steam Selling Price for 12% ROI ($/ton)***	17.56	13.8	27.5

* With an export power price of 8.0 cents/kWh and a steam price of 12 $/ton

** With a steam price of 12 $/ton

*** With an export power price of 8.0 cents/kWh

For an oxygen-blown facility with EPC costs of 99.8 M$ and a project life of 20 years, the return on investment (ROI) is expected to be less than zero, with a net present value (NPV) of -48.6 M$ given a 10% discount factor. Table 1.3 outlines the rate of return, NPV, payback year, and required electricity and steam selling prices to obtain a 12% ROI with other entries fixed. "High" and "low" estimates are listed as well to reflect the current cost accuracy assumption of +30/-15%.

Table 1.3 Oxygen-Blown Financial Cost Summary

	Cases			
	Base	**Low -15% EPC**	**High +30% EPC**	**Air-Blown Base Case**
ROI (%)*	<0	1.5	<0	5.9
NPV @10% Discount Rate, (M$)	-48.6	-26.9	-70.3	-14.6
Payback Year	>20	>20	>20	17
Electricity Selling Price for 12% ROI (¢/kWh)**	11.8	10.8	14.2	9.02
Steam Selling Price for 12% ROI ($/ton)***	>40	61	>100	17.56

* With an export power price of 8.0 cents/kWh and a steam price of 12 $/ton

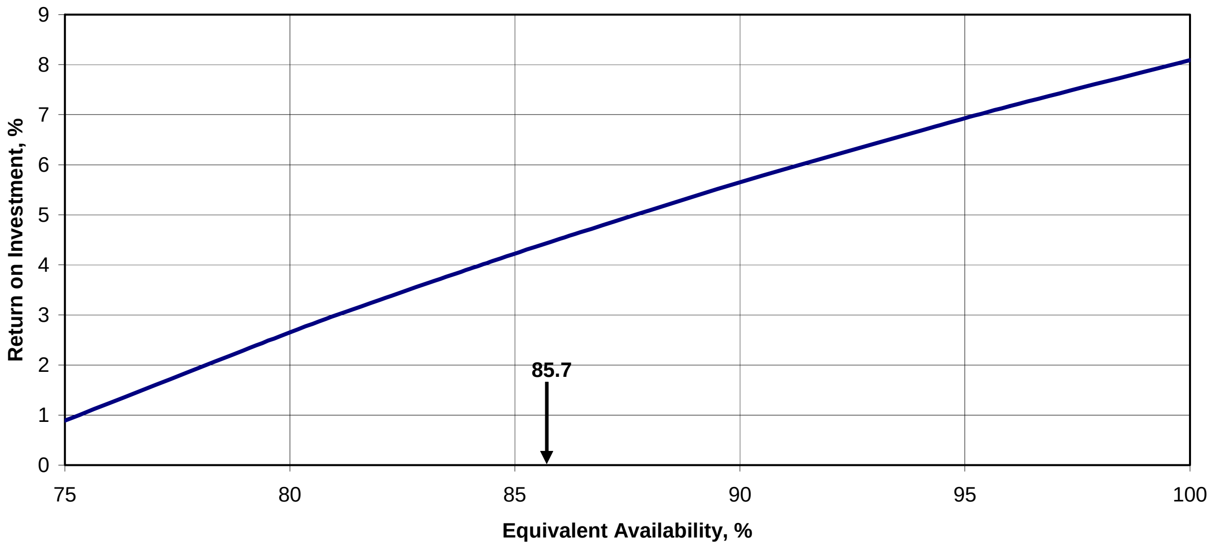
** With a steam price of 12 $/ton

*** With an export power price of 8.0 cents/kWh

The return on investment (ROI) for the air-blown case is higher than for the oxygen-blown case due mostly to the lower investment and higher steam export. Plant net CHP efficacy is 49% for the air-blown case and 29% for the oxygen-blown case.

The two inputs that had the greatest impact on overall project finances were guaranteed availability and the electricity tariff level. Total operating hours and electricity escalation, because of their direct relationship to availability and electricity value, also were found to have a strong financial impact. Figure 1.3 shows the relationship that varying the guaranteed availability has on ROI.

Figure 1.3 Effect of Availability on Air-Blown Project ROI

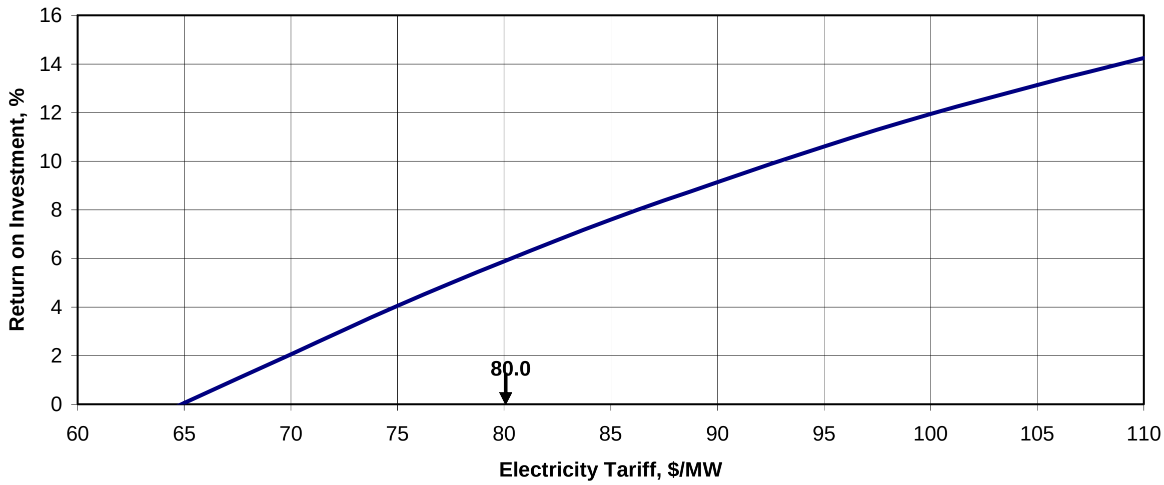


The impact that availability has on the plant economics comes as little surprise. Reliable operation is very important to assure that the cost of project development and construction can be recovered. Long downtimes throughout the life of the project will significantly hurt overall project economics given a 20-year project life. While gasification economics continue to show promise, project developers should consider the operating hours required for the facility to be economically justified.

Figure 1.4 shows the relationship between the electricity tariff value and ROI assuming a steam value of 12 $/ton.

Figure 1.4 Effect of Electricity Tariff on Air-Blown Project ROI

(Steam Value = 12 $/ton)



Because there has been no change in the financial assumptions made between the air-blown and oxygen-blown cases, the parameters found to be most sensitive in the oxygen-blown case are the same as in the air-blown.

This study has shown that:

- A ROI of 5.9% 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). The cost elements developed by this study should be useful as building blocks for developing reasonable cost estimates for plants of varying size within the 5-100 MW size range defined by this study.

- 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.

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

- As a result of this study, a list of potential enhancements has been identified that should provide additional cost savings as some of the improvements are researched, developed and implemented, such as:

- o Economy of scale (i.e., single train gasifier island)
- o The Stamet "solids" feeding system
- o A combined bottom and fly ash handling system
- o Candle filters for the removal of solid particles
- o A venturi scrubber in place of the impingement scrubber to reduce water consumption and capital investment
- o Improved heat integration
- o Simplified sour water stripper
- o Improved sulfur removal methods including warm sulfur removal (e.g., LO-CAT® system)
- o Warm mercury removal systems
- o Improved particulate removal systems

Subtask 3.3 employed a number of these improvements in the alternate design for the Air-Blown Eastern Coal Case.

- As a result of this study, a list of R&D needs have been identified including:
  - o Studying improved coal drying techniques
  - o Investigating the effect that the coal moisture content has on the U-GAS® gasifier operation
  - o Updating the database for gasification reactivity of the desired coal
  - o Characterizing the particulate properties
  - o Characterizing the hydrocarbon content of the syngas to confirm the sour water stripper design and effluent water treatment facilities
  - o Investigating cyclone performance at high temperatures (greater than 1000°F)
  - o Determining the combustion turbine performance capabilities for the desired engine(s) (both output and emissions)
  - o Further study of the ash characteristics associated with the char
- Based on the simulations prepared for this study the design should meet emission targets established by the DOE in their roadmap for 2010 (re. Section 5.3).

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

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, increasing the overall efficiency.

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

- Although it is known that reducing the moisture content of the coal feed going to the gasifier is more efficient than evaporating the moisture in the gasifier, it has not been established that 5% is the optimum moisture content of the gasifier feed. 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. recently completed the first two parts of the *Gasification Plant Cost and Performance Optimization Study* for the U.S. Department of Energy (DOE)/National Energy Technology Laboratory (NETL). This study focused on the use of the E-GAS^{TM} gasifier. These tasks used the E-GAS^{TM} 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.

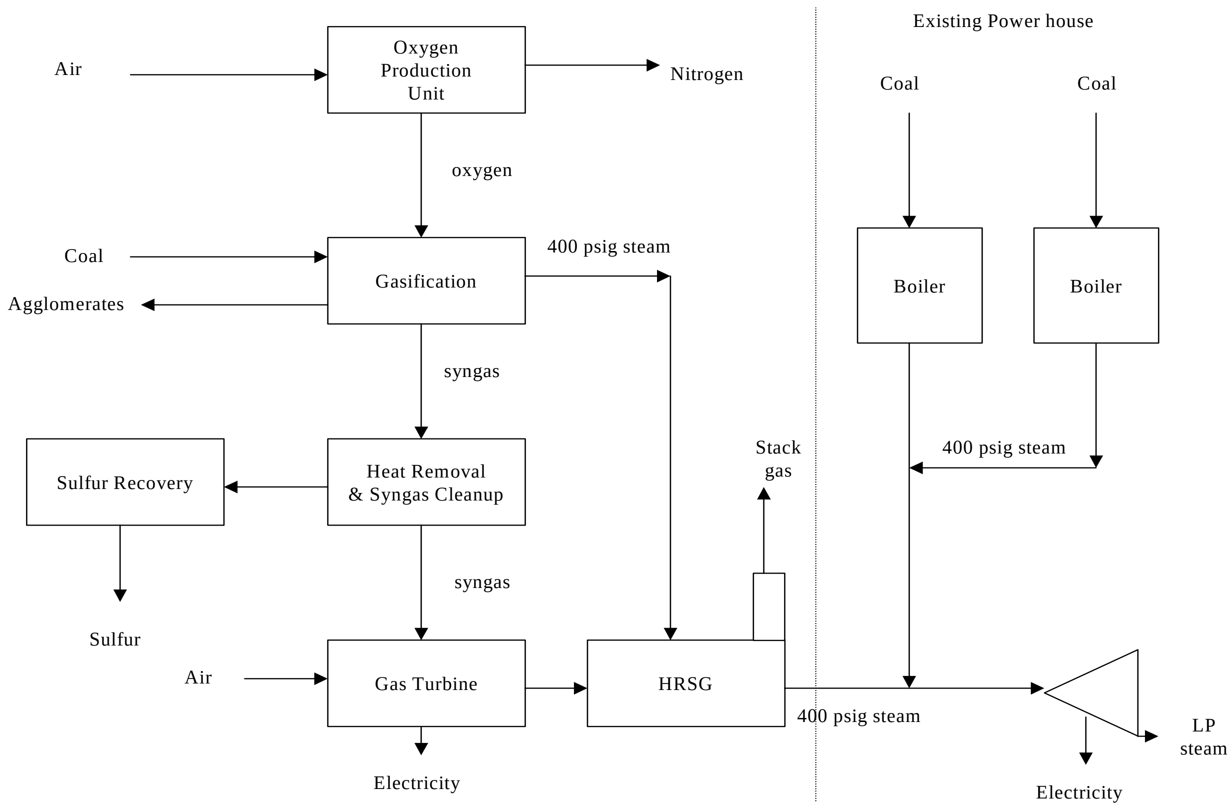
This project is defined as Task 3 of the *Gasification Plant Cost and Performance Optimization Study* and focuses on Gasification Alternative for Industrial Applications. This report encompasses the work performed on the first of three subtasks. In this report Subtask 3.2 provides a base case design for a facility where either air-blown or oxygen-blown gasification are considered. 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. 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.)

Subtask 3.2 consists of a preliminary design for a gasification plant at an upstate New York location processing Eastern coal. Coal from Southeast Ohio was selected as a low cost fuel to form the basis for the study. 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 will be 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,
- The nominal 400 psig/550°F pressure steam from the HRSG will be 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 current project discussed herein is modeled after an actual industrial facility, which is considering a similar project and provides a prototypical model for the study. Figure 2.1 illustrates how such a gasification plant might be integrated to replace a portion of the existing steam generation capability at an existing facility. The gasification plant is shown at the left side of the diagram, which consists of two steam boilers providing a total of about 360,000 lb/hr of nominal 400 psig steam for several steam turbine generators producing a total of 25 MW.

Figure 2.1 Block Flow Diagram



For the specific facility studied, there are other steam boilers and steam turbine generators within the existing powerhouse complex. The medium pressure (260 psig) steam leaving the steam turbine is used elsewhere throughout the complex. The contemplated project likely may replace two steam boilers with a fluidized bed U-GAS® gasifier processing about 350 tpd of coal. After cleanup the syngas is combusted in a GE 10 (or similar sized) gas turbine coupled to a heat recovery steam generator (HRSG).

This subtask proposes an IGCC design for upgrading the industrial power and steam facility. It begins with coal grinding and includes the gasification island, syngas cooling and cleanup, two GE 10 (or similar sized) gas turbines, and a heat recovery steam generator (HRSG). It assumes that existing facilities for coal receiving and storage, make-up water treating, and wastewater discharge can be used.

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

3.1 STUDY OBJECTIVES

The objective of this study is to investigate Gasification Alternatives for Industrial Applications. This is the first of three topical reports defined as subtasks under this contract with the DOE. This first topical report presents the capital and operating costs for a preliminary design of an industrial-size, Integrated Gasification Combined Cycle (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. This subtask developed preliminary design(s) for upgrading the industrial IGCC power and steam facility based on a set of criteria that can be applied to a wide cross section of industrial facilities across the United States.

Industrial facilities in the United States are facing stricter environmental regulations 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 premium fuels has placed a financial burden on the U.S. industry. Using coal as the fuel source at an industrial site gives the owner the knowledge that he will have low fuel costs that will be relatively constant and predictable. Furthermore, there are abundant coal resources (over 240 years supply at current usage rates) in the United States compared to limited amounts of oil and natural gas. As environmental rules tighten, industry will be forced to choose between continued expenditures: 1) for emission controls on coal boilers; 2) fuel switching to more costly premium fuels; or 3) shutdown of non-competitive facilities.

IGCC gasification plants can provide industry with a viable alternative. Gasification offers several advantages towards a long-term solution. First, coal is an abundant, low-priced energy source that is expected to have a stable low price over the foreseeable future and can be conveniently stored to avoid fuel supply disruptions. Secondly, IGCC systems have higher thermal efficiencies than steam boilers, which reduce fuel costs by reducing the amount of coal that is consumed to produce a given amount of power, and simultaneously, reduce the amount of carbon dioxide that is generated by burning this coal.

Pollution reduction also is simplified in an IGCC system. 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 dilution 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.

This report examines Gas Technology Institute's (GTI's) fluidized bed U-GAS® gasifiers coupled with two GE 10 (or similar sized) gas turbines and heat recovery steam generators (HRSGs) to co-produce power and high pressure (400 psig) steam at a specific industrial complex. Use of this site provides insight into typical retrofit issues, which are similar to many industrial complexes that would consider IGCC in the future. The steam from the IGCC process is integrated for use at various locations throughout the complex.

One of the objectives of this task is to enhance NETL's capabilities to perform system analysis. In order to accomplish this objective 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 appropriate approach commensurate to the level of results needed
- Used spreadsheets, ASPEN, and other software models to analyze a system or concept
- Determined economic and technical feasibility (developing cost estimates, project financing, mass and energy balances, etc.)
- Conducted sensitivity analyses to determine the primary variables that affect cost and/or performance
- Evaluated trade-offs for optimization
- Developed summary tables, flowcharts, written documentation, and presentation materials that effectively report the project objective, approach, and results

3.2 BACKGROUND

3.2.1 Introduction

The mission of NETL's Strategic Center for Coal is to ensure the availability of abundant low cost, domestic coal-based energy (including hydrogen) to fuel economic prosperity and strengthen energy security by developing advanced technologies and improving scientific knowledge.

The United States relies on fossil fuels for about 85 percent of the energy it consumes and forecasts indicate U.S. reliance on these fuels could exceed 87 percent by 2025.

DOE's fossil energy activities are designed to ensure that the economic benefits from moderately priced fossil fuels and a strong domestic industry (that creates export-related jobs) are compatible with the public's expectation for exceptional environmental quality and reduced energy security risks.

These activities include fostering the development of energy systems and practices that will provide current and future generations with energy that is clean, efficient, reliable, and reasonably priced.

The Strategic Center for Coal focuses virtually and exclusively on supporting the President's top initiatives for energy security, clean air, climate change, and coal research. The Center's programs: 1) support the development of lower cost, more effective pollution control technologies embodied in the President's Coal Research Initiative; 2) expand the nation's technological options for reducing greenhouse gases either by increasing power plant efficiencies or by capturing and isolating these gases from the atmosphere as called for by the President's Global Climate Change Initiative; and 3) measurably add to the nation's energy security by providing a longer-term alternative to imported oil, such as hydrogen produced from coal as conceived in the Hydrogen Initiative.

Together with the Coal Utilization Research Council and Electric Power Research Institute, the Department has developed a Clean Coal Technology Roadmap that outlines a clear path forward that establishes performance targets and time frames for advanced coal technologies research, development, and demonstration. The performance targets are shown in Table 3.1.

The Strategic Center for Coal is also focused on the development of a new generation of electric power generating "platforms" employing advanced coal gasification, coal-capable turbines, and novel combustion concepts that will form the core of the zero-emission coal plant of the future.

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 of this contract 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

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

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, and by reference should be considered a part of the final report. [3,4]

Table 3.1 Performance Targets$^{(*)}$

	Reference Plant$^{(**)}$	2010	2020
Air Emissions			
Sulfur (SO_2) removal	98%	99%	>99%
NOx, lb/10^6 Btu	0.15	0.05 $^{(***)}$	<0.01
particulate matter, lb/10^6 Btu	0.01	0.005 $^{(†)}$	0.002
Mercury (Hg) removal	$^{(††)}$	90%	95%
By-Product Utilization	30%$^{(†††)}$	50%$^{(‡)}$	100%
Plant Efficiency (HHV)$^{(‡‡)}$	40%	45-50%	50-60%
Availability$^{(‡‡‡)}$	>80%	>85%	>90%
Plant Capital Cost$^{(‡‡)}$, $/kW	1000-1300	900-1000	800-900
Cost of Electricity$^{(ƚ)}$, cents/kWh	3.5	3.0-3.2	<3.0

* Targets are without carbon capture and sequestration and reflect current cooling tower technology for water use

** Reference plant has performance typical of today's technology; improved performance achievable with cost/efficiency tradeoffs.

*** For existing plants, reduced cost for achieving <0.01 lb/10^6 Btu using combustion control by 25% compared to SCR by 2010; same cost reduction for 0.15 lb/10^6 Btu by 2005.

† Achieve targets for existing plants in 2010: 99.99% capture of 0.1-10 micron particles

†† Some mercury reduction is being achieved as a co-benefit with existing environmental control technologies

††† Represents average for existing plant locations

‡ Target represents technically achievable for new or existing plants; economics are site specific

‡‡ Range reflects performance projected for different plant technologies that will achieve environmental performance and energy cost targets

‡‡‡ Percent of time capable of generating power (ref. North American Electric Reliability Council)

ƚ Bus-bar cost-of-electricity in today's dollars; reference plant based on $1,000/kW capital cost, $1.20/$10^6$ Btu coal cost

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."[9] This task was

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[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/coal/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 No. DE-AC26-99FT40342, "Gasification Plant Cost and Performance Optimization"

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

designed to develop smaller gasification plants for industrial applications using GTI's U-GAS® fluidized bed gasifier. Task 3 is divided into three technical subtasks. Subtask 3.2 investigates a brownfield design modeled after the requirements of a specific industrial site in upper New York state that will co-produce both power and steam. Both air and oxygen-blown gasification systems were considered. Subtask 3.3 will generalize and optimize the Subtask 3.2 plant design. Subtask 3.4 will develop a design of a nominal 200 MW power plant that will use North Dakota Lignite.

This document is the Topical Report for Subtask 3.2.

3.2.2 Contract Overview

The objectives of this Gasification Plant Cost and Performance Optimization contract were to examine the current state-of-the-art of coal gasification and to develop designs that would reduce the cost of power generated by IGCC plants by reducing capital and operating costs, increasing efficiency, and making them less polluting. The original contract contained two major tasks:

- Task 1 developed cases using both coal and petroleum coke feedstocks. Task 1 included nine individual subtasks, which are described below. A primary aspect of the study was Subtasks 1.2 and 1.3, which considered co-producing hydrogen and steam as part of a market entry strategy for lowering the technical risk and the capital and operating costs of future coal gasification plants. A secondary benefit was to provide baseline cases from which the Department of Energy can measure future progress towards achieving their goals.

- Task 2 considered the coproduction of power and liquid transportation fuel precursors by the gasification of either petroleum coke or coal. Task 2 had three subtasks.

3.2.2.1 Task 1

The primary objective of Task 1 was to develop optimized engineering designs and cost estimates for five Integrated Gasification Combined Cycle (IGCC) plant configurations fueled by either coal or petroleum coke. Starting from the as-built design, operation, and cost information from the commercially proven Wabash River Coal Gasification Repowering Project, the following eleven cases were developed:

- Wabash River Greenfield Plant (Subtask 1.1)

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

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

[9] Contract modification November 21, 2003

- Non-optimized Petroleum Coke IGCC Coproduction Plant (Subtask 1.2)
- Optimized Petroleum Coke IGCC Coproduction Plants that produce hydrogen and industrial-grade steam in addition to electric power (Subtasks 1.3 and 1.3 Next Plant – four cases)
- A future Advanced Coal IGCC Power Plant producing only power using a next generation gas turbine (Subtask 1.4)
- Single-train Coal and Coke IGCC Power Plants (Subtask 1.5 – two cases) A Nominal 1,000 MW Coal IGCC Power Plant (Subtask 1.6)
- A Coal to Hydrogen Plant (Subtask 1.7)

The left side of Figure I.1 shows the chronological development of the above Task 1 gasification plant designs.

In addition there were two other subtasks. Subtask 1.8 reviewed various warm gas cleanup methods that are applicable to IGCC systems. The Subtask 1.8 cases covered a variety of processes and provided a look at potential future syngas cleanup methods. Subtask 1.9 documented the method and results of the availability calculations for the design subtasks. The results of the Task 1 study have been previously reported in a Topical Report[10].

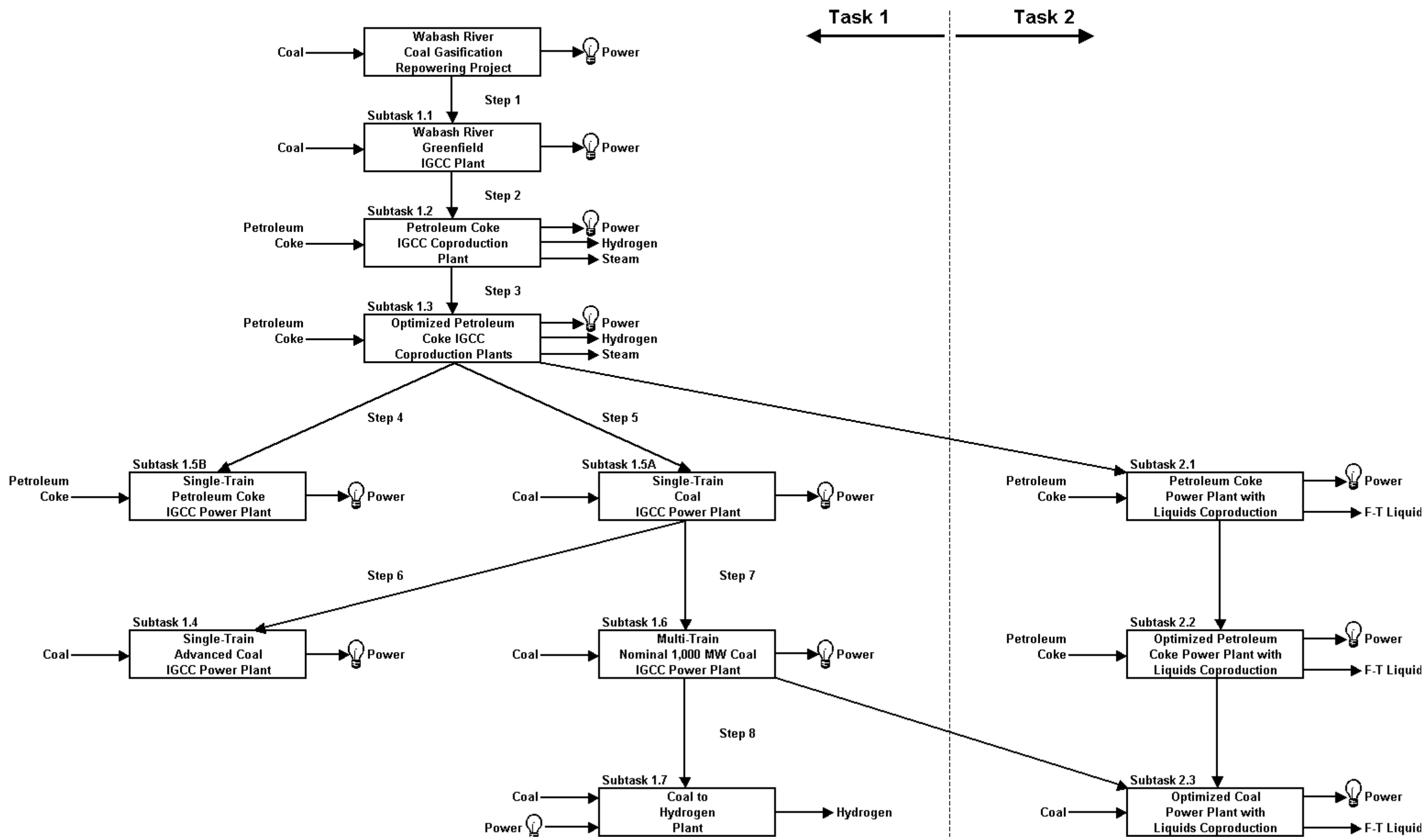
3.2.2.2 Task 2

Task 2 had the objectives of developing optimized designs, cost estimates and economics for petroleum coke gasification power plant with liquids fuel precursors coproduction and a coal gasification power plant with liquids fuel precursors coproduction. Based on the results of Task 1, the following three cases were developed.

- A non-optimized petroleum coke IGCC power plant with liquid fuel precursors coproduction (Subtask 2.1)
- An optimized petroleum coke IGCC power plant with liquid fuel precursors coproduction (Subtask 2.2)
- An optimized coal IGCC power plant with liquid fuel precursors coproduction (Subtask 2.3)

Figure I.1

Schematic Diagram Showing the Chronological Development of the Gasification Plant Designs



The right side of Figure I.1 shows the chronological development of the three Task 2 subtasks and the Task 1 subtasks on which they are based.

The Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant was the basis for the two petroleum coke cases of Subtasks 2.1 and 2.2. The Subtask 1.6 1,000 MW Coal IGCC Power Plant and the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction were the bases for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. Building the Task 2 cases on the previous Task 1 cases provided a common basis for comparison between the cases with and without the coproduction of liquid fuel precursors.

The results of the Task 2 study have been previously reported in a Topical Report.[11]

3.2.3 History of U-GAS® Process

The name U-GAS® was derived from the suitability of the gaseous product as a fuel for utility applications, such as firing in a gas turbine. The process was developed over 40 years ago to support GTI's other gasification efforts to develop substitute natural gas (HYGAS®). The process uses a single fluidized bed reactor where carbon in the coal is reacted with steam to form hydrogen and carbon monoxide. The unique design of the reactor allows for high carbon conversion of many different types of fuel. The technology is especially well suited to fuels with a high ash content. Another feature of U-GAS® is its ability to operate with either air or oxygen as an oxidant.

The first pilot plant testing was begun in 1973 at IGT's[12] Energy Development Center (EDC) with a near atmospheric reactor (60 psia). Numerous tests were conducted using this 24 tpd pilot unit. On the basis of the successful testing in the pilot unit, the technology was selected for commercialization by the Synthetic Fuels Corporation for a demonstration site in Memphis, Tennessee. Memphis Light, Gas and Water Company was the host utility. Foster-Wheeler performed extensive design engineering for this project. IGT conducted testing in the pilot plant to support the design efforts with Kentucky #9 coal. Design data developed from that study were used to support efforts for this project as appropriate.

At the heart of the Memphis Project were 3 gasifiers, each 15-foot in diameter, capable of gasifying 1,000 tpd of coal. The plant was designed to produce 175 million cubic feet per day of medium Btu syngas. Foster-Wheeler performed the detailed engineering and

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

[11] "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.

[12] The Institute of Gas Technology (IGT) is a predecessor company to GTI.

cost estimates for the project. Extensive design engineering was conducted on the scale-up concepts for the plant. The project was cancelled, when the Synthetic Fuels Corporation was disbanded in 1984. The engineering for the facility has proven valuable in providing the basis for ongoing commercialization efforts.

Research at the pilot scale continued with the construction of a high pressure pilot facility at the EDC that was commissioned in 1982 (5 tpd coal feed rate, 510 psia). This facility was operated throughout the 1980s to provide data to support development of U-GAS® for gas turbine applications. These two pilot plants logged over 11,000 operating hours. During this time period 125 test runs were conducted utilizing more than 20 different fuel types including bituminous and subbituminous coal, lignite, peat, oil shale, and metallurgical coke.

The U-GAS® technology was exclusively licensed to Tampella Power Inc. in 1989. GTI designed a gasification pilot plant that was constructed by Tampella Power in Tampere, Finland. This pilot facility was designed to demonstrate the gasification island of an IGCC power plant. It included fuel preparation and hot gas clean-up. This facility was commissioned in 1991 and could maintain a coal feed rate of 30 tpd. The operating pressure was 330 psia. This facility was also designed to gasify biomass. The fuel feed rate with biomass was 100 tons per day. The Tempere pilot plant has logged 3,800 hours of operation. During this time period, 5,900 tons of fuel were processed in 26 test runs. The tested fuels included coal (Polish, Colombian, coke, German lignite) and biomass (wood, paper mill wood waste, forest residue, willow, straw, alfalfa) and mixtures of coal and biomass. The gas generated in the gasification plant was combusted in a heat recovery boiler producing district heat for the city of Tampere.

Commercialization plans for U-GAS® proceeded with the development of the Tom's Creek Clean Coal project that was awarded to Enviropower (a division of Tampella). Enviropower planned to construct a coal-based IGCC facility in Virginia. The design was based on powering a GE 6B engine. This project was not completed for a variety of commercial factors. Again, substantial design engineering was developed for the project that was beneficial to this study.

Enviropower was also awarded a contract for a 75 MW power plant to gasify alfalfa stems. This MINVAP project was to be located in Minnesota. An agricultural cooperative was to be the co-owner of the facility with equity participation by key suppliers. This project was not completed due to various financial requirements imposed upon the participants forcing them to discontinue their equity positions.

Kverner subsequently purchased Tampella and divested their gasification interest to Carbona for commercialization. Carbona and GTI have collaborated on a number of projects. Carbona is currently constructing a 5 MW biomass gasification demonstration project in the Netherlands that is cofunded by the International Energy Agency (IEA) and DOE.

The first commercial scale demonstration of U-GAS® was conducted by Shanghai Coking and Chemical Company in China. A battery of eight gasifiers was built to produce fuel gas for coke ovens, freeing up the higher heating value coke oven product gas for blending into town gas. The use of eight gasifiers provided a high degree of reliability (2 were used as spares) and the wide turndown capability needed to meet the seasonal demands for the fuel gas. The gasifiers operated at low pressure (3 atm) and each unit processed 150 tpd of Chinese bituminous coal and produced 500,000 Nm^3/day of fuel gas. The facility was commissioned during 1995 and began commercial operation. The plant has logged over 70,000 hours of operation, processed over 220,000 tonnes of coal, and produced over 1 billion Nm^3 of fuel gas for commercial use. The longest continuous operation of a gasifier was 3,000 hours. This plant is currently "mothballed" due to a lack of demand for the "town gas" that was supplied from the coke ovens fired with synfuel. This is a result of natural gas becoming available to mainland China. GTI is working with the plant's owners to identify new uses for the fuel gas facility.

In the late 1990s, the EDC was sold. The U-GAS®, HYGAS®, and other pilot plants located at the site were dismantled and sold. Test equipment actively being used for ongoing research projects was relocated to GTI's new location in Des Plaines, Illinois (since 1994).

GTI has recently constructed a unique gasification test platform at its research campus in Des Plaines. The Flex-Fuel Gasification Test Facility is being used to facilitate commercialization of advanced gasification and down-stream end-use technologies. The State of Illinois Department of Commerce and Economic Opportunity and GRI[13] provided financial support for construction of the facility. This test platform will evaluate advanced and innovative gasification processes employing a variety of low-cost, solid carbonaceous fuels. The Flex-Fuel Gasification Test Facility will employ GTI's fluidized bed gasifier as a primary platform for testing coal and a variety of other solid fuels, including biomass. The procurement and construction of this facility in 2003-2004 provides an up-to-date basis for costing a variety of equipment and elements associated with gasification equipment and hardware used to prepare this report.

To date, no other gasifier in the world has been operated on the range of feedstocks or the range of scale as GTI's fluidized bed systems. Its flexibility, scale-up experience and in particular the capability to limit tars and oils in the fuel gas to very low levels makes U-GAS® the ideal choice for the proposed application. Gas Technology Institute's (GTI's) U-GAS® fluidized bed gasifier technology was chosen for this study for the following reasons:

- Fluidized bed technology is versatile and capable of gasifying a wide range of fuels including lignite

[13] The Gas Research Institute (GRI) is a predecessor company to GTI that provides funding for research by combining funds collected from the gas industry.

- Fluidized bed technology can be operated in either the air-blown or oxygen-blown mode. This provides owners with an option to select technology that best meets their needs for efficiency or for process criteria including sequestration of CO_2.
- The scale of equipment is ideally suited for fluidized bed technology
- The technology is ready for commercial deployment
- Older cost studies examining this technology needed to be updated to better compare this technology with other gasification options.

3.3 METHODOLOGY

3.3.1 Basis for Tasks 1 and 2

In 1990, Destec Energy, Inc. of Houston, Texas and PSI Energy, Inc. of Plainfield, Indiana formed the Wabash River Coal Gasification Repowering Project Joint Venture to participate in the Department of Energy's Clean Coal Technology Program by demonstrating the coal gasification repowering of an existing 1950's vintage generating unit. In September 1991, the project was selected by the DOE as a Clean Coal Round IV project to demonstrate the integration of the existing PSI steam turbine generator and auxiliaries, a new combustion turbine, a heat recovery steam generator, and a coal gasification facility to achieve improved efficiency and reduced emissions. In July 1992, a Cooperative Agreement was signed with the DOE. Under terms of this agreement, the Wabash River Coal Gasification Repowering Project Joint Venture developed, constructed and operated the coal gasification combined cycle facility. The DOE provided cost-sharing funds for construction and a three-year demonstration period. Construction was started in July 1993, and commercial operation began in November 1995. The demonstration period was completed in January 2000.[14,15]

The participants jointly developed, separately designed, constructed, owned, and operated the integrated coal gasification combined-cycle power plant to repower the oldest of the six units at PSI's Wabash River Generating Station in West Terre Haute, Indiana. The Destec gasification process is integrated with an existing steam turbine generator using some of the pre-existing coal handling facilities, interconnections, and other auxiliaries. The power block consists of an advanced General Electric MS 7001 FA gas turbine unit that produces 192 MW, a Foster Wheeler HRSG, and a 1953 vintage Westinghouse reheat steam turbine. The steam turbine, which was refurbished as part of the repowering project, produces 104 MW of power. Parasitic power is 34 MW giving a net power output of 262 MW.

Since the initial startup of the Wabash River Repowering Project, many modifications and improvements have been made to the plant to improve plant performance and to

[14] Topical Report No. 20, "The Wabash River Coal Gasification Repowering Project – An Update," U. S. Department of Energy, September 2000.
[15] Global Energy, Inc., "Wabash River Coal Gasification Repowering Project – Final Report," September 2000.

increase availability. The net result of these changes has been a substantial improvement in plant operations. Furthermore, in addition to operation on Illinois coals, the plant has demonstrated successful and reliable operation on petroleum coke.

The design, construction, cost, and operational information obtained from this commercial facility provide the basic information for Tasks 1 and 2 of this project. That is, the sum total of knowledge gained from the plant starting from the initial design through current operations on both coal and petroleum coke have been studied to compile relevant information for this project. Current performance information was analyzed to develop a heat and mass balance model that was the basis for developing models for the subsequent subtasks. As-built cost information provided the cost basis for the cost estimates. Because the cost estimates are based on actual equipment purchases and construction labor use, these cost estimates are more accurate than typical estimates would be for this type of study. Availability and reliability information from the final year of the DOE demonstration period were the basis for the availability analyses.

3.3.2 Basis for Task 3

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 is 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 does not consider applications for a grass-roots plant, but rather as a retrofit situation that uses part of the existing industrial facility's infrastructure.

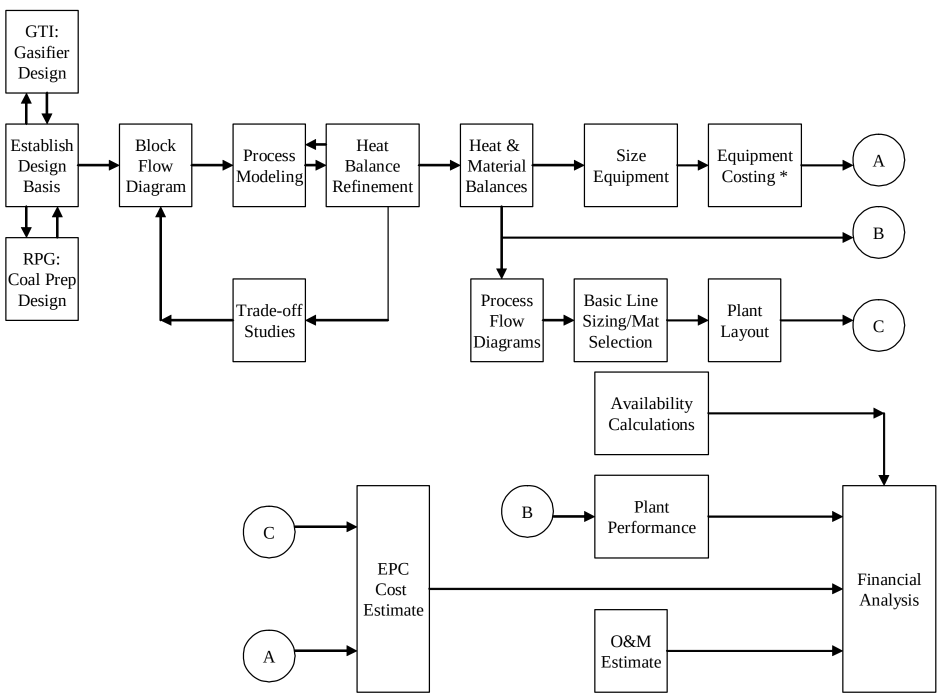
The U-GAS® gasification technology system developed by the Gas Technology Institute was the basis for this project study. This system is based on a non-slagging, fluidized bed gasifier. The total of knowledge gained from previous GTI gasifier designs using this technology on coal has been studied to compile relevant information for this project. A history of the U-GAS® process is provided in Section 3.2.3 of this report.

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 G (the design basis work plans). Based on these design bases, input from GTI's gasifier model and input from various vendors, an elementally balanced process simulation model of the gasifier was developed using Aspen Plus®, a commercially available process simulation program. This is a very detailed process simulation program that simulates the various heat exchange and steam generation steps within the gasification area. Process simulations were also developed for the syngas cooling and cleanup portion of the plant and the sour water stripper. The resulting heat and material balances provided the feed to the GateCycle simulation program for a detailed simulation of the power block. This report and its addendums contain sufficient information for verification of the carbon, slag, sulfur, and heat balances.

PFDs, sized equipment lists, line sizing, and other information necessary to calculate the plant cost were developed based on the model results. 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 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.[16] 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.[17]

Recent data presented at the 2002 Gasification Technologies Council conference by Clifton Keeler show further reliability improvements in the on-stream performance of the Wabash River Repowering Project.[18] 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.

Since the objective of this availability study is to determine the projected annual revenue stream, this study does not differentiate between forced and scheduled outages. In other words, 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. Consequently, the annual availabilities reported in this study will be lower than those from studies which do not consider scheduled outages.

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

3.4.1 Use of Natural Gas

The facility design is based on multiple trains for high reliability. 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. However, 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.

[16] "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.

[17] 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, Pala Alto, CA 94304, August 1985.

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

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.

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.

The Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction uses purchased power to maintain Fischer-Tropsch liquids production during periods when the combustion turbine is unavailable.

Although natural gas is used in most of the Task 1 and Task 2 subtasks to increase production when sufficient syngas is not available, no natural gas is used for this purpose in Subtask 3.2. However, some natural gas is used during startup for heat up, refractory conditioning, etc. This gas usage is considered to be an O&M cost and not a feedstock cost.

3.5 COMMODITY PRICING

The initial basis for the commodity prices into the gasification model came from information provided by U.S. government agencies. This includes data from the DOE's Energy Information Administration (EIA) Annual Energy Outlook 2004[19] for commercial electricity values, natural gas, and coal, and from the U.S. 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 value was normalized where necessary to reflect current nominal value, using a 3% inflation rate. The preliminary model runs were made using these inputs. Table 3.2 below lists the major assumptions for commodity prices. The financial sensitivities (Sections 5 and 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.

Table 3.2 Basic Economic Parameters

Feeds	Price	Escalation (%/yr)
Coal	27.20 $/short ton	2.0
Natural Gas, HHV	4.68 $/MBtu	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

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

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 in Subtask 3.2, the small amount of natural gas that is used is accounted for in the variable O&M costs, making this input insignificant. This number may be relevant to future tasks, if co-firing is used, or as a comparison with other industrial power producing alternatives.

In keeping with previous Nexant studies and expectations of oversupply in the coal industry, the 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 value inputs.

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

In order to develop the appropriate financial assumptions for the 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 1) NETL's "Quality Guidelines for Energy System Studies", 2) an industry study analyzing the potential for gasification in the U.S. refining market, and 3) 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 made can be found in Addendum C of this report. A few of the major assumptions and areas that will be explored via sensitivity analysis are listed below:

- + 30/-15% accuracy assumed for this phase of the analysis
- A 15% project contingency applied across the entire plant with the exception of the gasifier block. For the gasifier block, a separate 25% process contingency was used to reflect the higher uncertainty in this unit's cost estimate.
- Scheduled 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 financial model from the design basis. The material and energy balance provided by GTI and verified by Nexant/DOE, along with the subsequent design work by Nexant and NETL, set the entries for items such as power output, sulfur production, and quantities of ash produced. The plant EPC cost used for the model analysis was determined by establishing installed cost estimates for all major unit operations, off-sites, and balance-of-plant items. The basis for installed costs came from a combination of GTI input for the gasifier block, vendor quotes for major unit operations, process design software, and team expertise for the remaining pieces of equipment. A more rigorous explanation of how these numbers were developed is given in Sections 5.5 and 6.5 (Plants Costs). Appropriate scale-up factors used in previous gasification projects allowed any additional equipment to be properly estimated.

4.1 STUDY BASIS

This study investigates the cost for installation and operation of a combined heat and power (CHP) facility at an industrial site. The goal of the study is to identify alternatives for reducing operating costs and lowering plant emissions associated with power generation. Two cases have been developed – an air-blown and an oxygen-blown case. The location for this facility is at an industrial site in upstate New York.

Design Criteria for Subtask 3.2 is:

- 2 gas turbines @ ~12.5 MW each (total = 25 MW)
- maximize co-generation of steam from the gasifier and HRSG (approximately 130,000 lb/hr for the air-blown case)
- Export steam to industrial site (at 400 psig/550°F)
- Southeast Ohio coal (assume 15% moisture for design, 8.5% 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		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

Environmental performance based on the DOE target emission and performance goals established in their roadmap for 2010 have been used as the basis for emissions targets as follows:

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

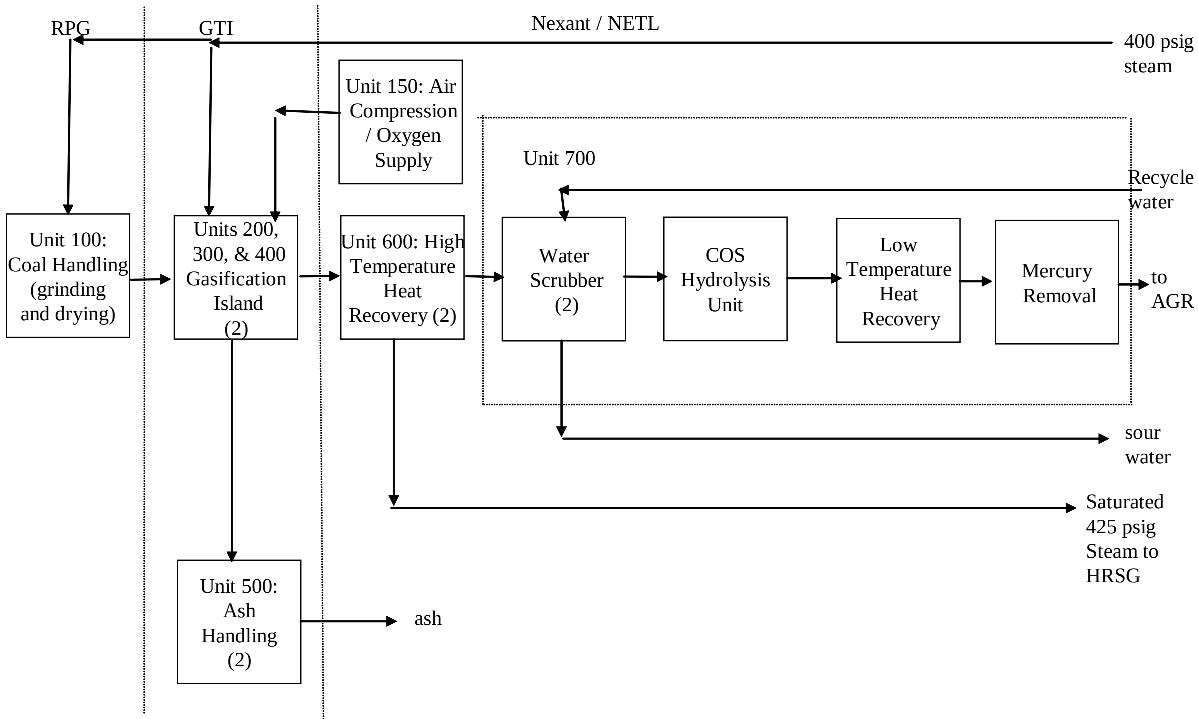
4.1.1 Plant Description

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

- Unit 100: Coal Prep/Handling
- Unit 150: Air Separation or Compression Unit
- Unit 200: Solids Feeding System
- Unit 300: Gasification
- Unit 400: Fines Separation
- Unit 500: Ash Handling
- Unit 600: High Temperature Heat Recovery
- Unit 700: Water Scrubber, COS Hydrolysis Reactor, Low Temperature Heat Recovery and Mercury Removal
- Unit 800: Acid Gas Removal (Amine) Unit, Sour Water Stripper (SWS), Sulfur Plant, Tail Gas Clean-up
- Unit 900: Power Block including the gas turbines (CT) and heat recovery steam generator (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 4.1 is a block flow diagram of the plant. It is in two parts. The first part on page 4-3 shows the syngas generation and processing areas, and the second part on the next page shows the sulfur removal and sulfur recovery areas, sour water stripper, and the power block.

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

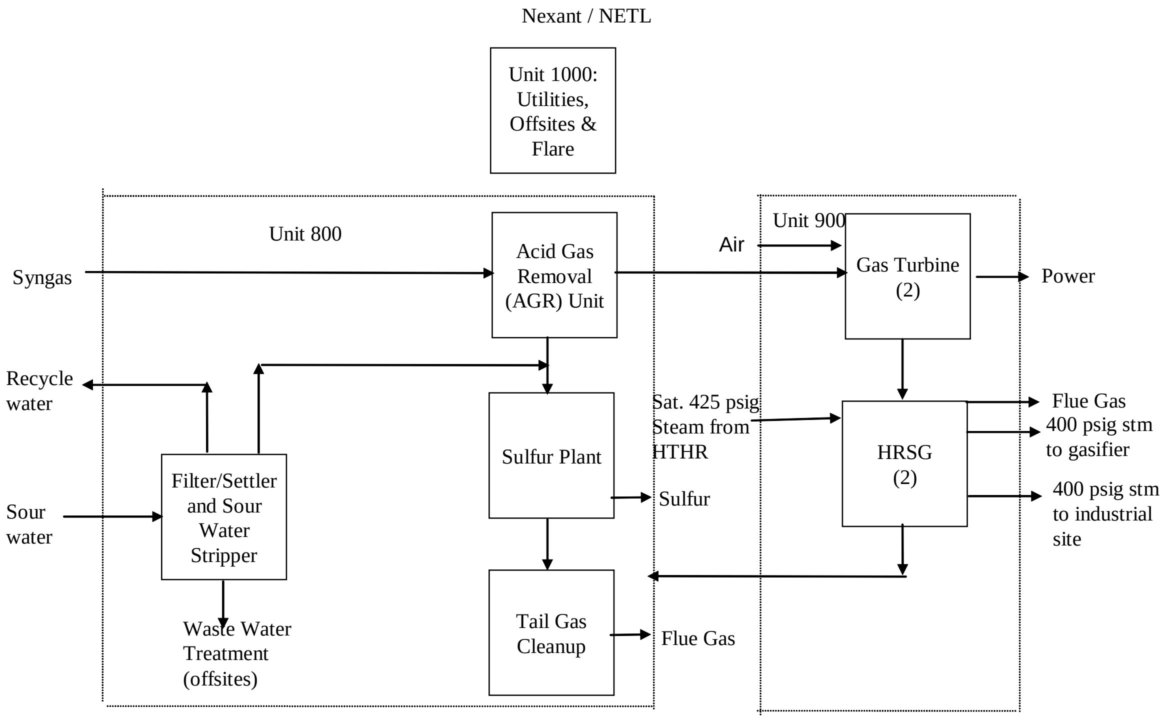


4.1.2 Site Selection

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 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 coal from Southeastern Ohio was chosen as the design fuel representing 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 these fuels. It is anticipated that coal from this area can be delivered to industrial facilities at a cost of about 1.00 to 1.50 $/MBtu. 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 in the past. 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 an agglomerating mode. Agglomeration of the ash 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 topical report is the first in a series of studies of a preferred design(s) for upgrading the industrial IGCC power and steam facility. Each study is based on a set of criteria that can be applied to a wide cross section of industrial facilities across the United States. This first study 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. Subsequent studies will examine a variety of alternatives for optimizing plant costs, using new sulfur removal technologies and examining the use of lignite.

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

- The use of IGCC increases efficiency and reduces operating costs compared to use of premium fuels or continued upgrading of old coal facilities to meet tighter emission standards.

- IGCC is the cleanest means of providing power and steam from coal, thereby reducing emissions from the utility facilities at an industrial site.

- The use 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.

The second objective of this study (i.e., to apply the GTI fluidized bed U-GAS® gasifier technology at a stand-alone lignite fueled IGCC power plant in North Dakota) was achieved with Subtask 3.4.

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 U.S. 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 gasification plants can provide industry with a viable alternative. Gasification offers three advantages as a long-term solution. These relate to lower cost, lower emissions, 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 purchase of energy.

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.

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 (400 psig/550°F) steam at a typical industrial complex. The steam can be used at various locations throughout the complex. Two coals are studied as feedstock:

a high sulfur Southeastern Ohio coal (Subtasks 3.2 and 3.3 producing about 25 MW of power) and a low-sulfur, low rank coal in Subtask 3.4 (producing about 200 MW of power).

Other factors evaluated as part of the sensitivity analysis include various methods of oxygen production (such as the oxygen ionic membrane technology currently under development); alternative sulfur removal technologies; and alternative technologies for ash removal from the gasifier.

4.3 HEAT INTEGRATION

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

The syngas cooling section of the plant is designed to minimize deposition and erosion problems as a result of dust carried in the syngas. Therefore only one heat exchanger is used before the water scrubber. This single heat exchanger is a fire tube boiler design that cools the hot syngas leaving the third stage cyclone from about 1750°F to about 600°F by producing saturated steam at 415 psig and 450°F. This steam is superheated to 550°F in the HRSG. For maximum thermal efficiency, the 415 psig steam should be superheated with the hot syngas before it enters the steam boiler, but this would add another exchanger to the syngas cooling train and would increase the potential for additional deposition and erosion problems.

The water scrubber has two functions: it scrubs the dust, light oils, HCl , etc. out of the syngas and simultaneously cools the syngas from 600°F to 265°F. The cleaned syngas now is processed to remove contaminants. These include sulfur compounds, mercury and ammonia. This is accomplished first by being reheated to 275°F with steam to ensure that the syngas is dry when it passes through a hydrolysis reactor (to convert the COS to H_2S. The H_2S is removed downstream in the acid gas removal unit). The syngas leaving the hydrolysis reactor then is cooled in a series of three heat exchangers. The first exchanger cools the syngas to about 240°F by preheating boiler feed water only for the upstream steam boiler. The second is an air cooler, and the third is a water cooler that combine to cool the syngas to 110°F.

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

4.4 TECHNOLOGY DRIVERS

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

4.4.1 Cost Drivers

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

Over the past four years, natural gas prices have risen dramatically. With the price of natural gas currently selling for over 6.00 $/MBtu, many companies are worried about their energy supply costs as near term gas prices are expected to continue higher. The rise in gas prices is forcing industry to critically examine their energy supply choices. Recent articles in the press[1] 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 $/MBtu by 2010 and then slowly increase to 5.10 $/MBtu in 2025.[2] 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 per $/MBtu by 2025. Natural gas prices have demonstrated significant volatility over the past few years. This is not expected to change considerably since these variations are based on changes in U.S. supply and demand options and future world events.

Gas prices paid by industry are not fully reported on EIA databases to retain confidentiality of sensitive company data. Typically only about 12% of industrial pricing is reported. Industry payments for gas vary widely; a key determinant in price variation is whether a company is in a position to bypass the local distribution company (LDC) for gas purchases and buy gas directly via a pipeline. When gas is available bypassing the LDC, the price above well-head is about 0.90 $/MBtu on average nationally. However,

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

[2] Annual Energy Outlook 2004 with Projections to 2025, www.eia.doe.gov/oiaf/aco/economic.html

data reported in key industrial states like New York, Pennsylvania, and Ohio, indicates that industries pay as much as 1.00 to 2.20 $/MBtu in transportation cost to the LDC. This could raise the long-term expectation for natural gas price above those reported by the EIA for delivered cost.

Coal can play a greater role for many industrial facilities. In contrast to natural gas, coal prices have remained stable over the past decade. Coal prices to industrial users are typically between 1.25 to 2.00 $/MBtu (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 $/MBtu, depending on the specific fuel transportation factors for 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 $/MBtu. However, conversion of a solid fuel to gas is capital intensive, and the cost is high. Thus, the critical decision for implementation of this technology lies in the long term differential fuel costs between coal and natural gas including the attendant emission controls associated with their use. Although not every industrial facility can benefit from coal gasification on a purely price basis, there are clearly many facilities that can justify a serious evaluation of this technology as long term solution to meeting its energy needs.

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

4.4.2 Emission Drivers

Natural gas has been the industrial fuel of choice for the past 30 years. Gas use 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 boilers from coal 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 this year the U.S. EPA is planning to release new standards for emission controls at industrial plants that will require essentially all sites with combustion facilities rated at over 10 MBtu/hr to apply state-of-the-art emission controls. Emission control will be required for sulfur, NOx, particulates, mercury, and possibly chlorides. Post combustion control for all these emissions will require significant expense for industrial utility systems that are in many cases 40 to 60 years old. Replacement of old coal fired equipment with new systems at an industrial scale is relatively expensive. This application of IGCC technology has been demonstrated to be environmentally superior to post combustion emission controls and can be applied to industrial facilities in a cost effective manner.

4.4.3 Reliability Drivers

Reliability is a tangible factor for industrial applications; however the value of reliability can only be quantified by each facility individually. Costs associated with loss of manufacturing and industrial lost productivity have been studied by EPRI and others. These studies reflect the importance for 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 (when using liquid fuels) and are limited in the annual number of hours for which they can be used.

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

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

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

4.5 PLANT SIZE

The plant consists of 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 two parallel gasifier trains using GTI's U-GAS® fluidized bed gasifiers. For the purposes of the study, the GE 10 engine was selected for the gas turbine. Each turbine requires 140.5 MBtu/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

The gasifier island, syngas cooling and power island consist of two identical parallel trains. The coal handling, syngas cleaning, and auxiliary systems are designed so that on a short term basis, a single train is capable of handling the full throughput needs for the entire facility.

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

A project construction schedule is shown in Addendum H of this report.

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 consideration. 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. Further cost savings have been identified for study, but not yet quantified. These options are investigated in Subtasks 3.3 and 3.4.

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.

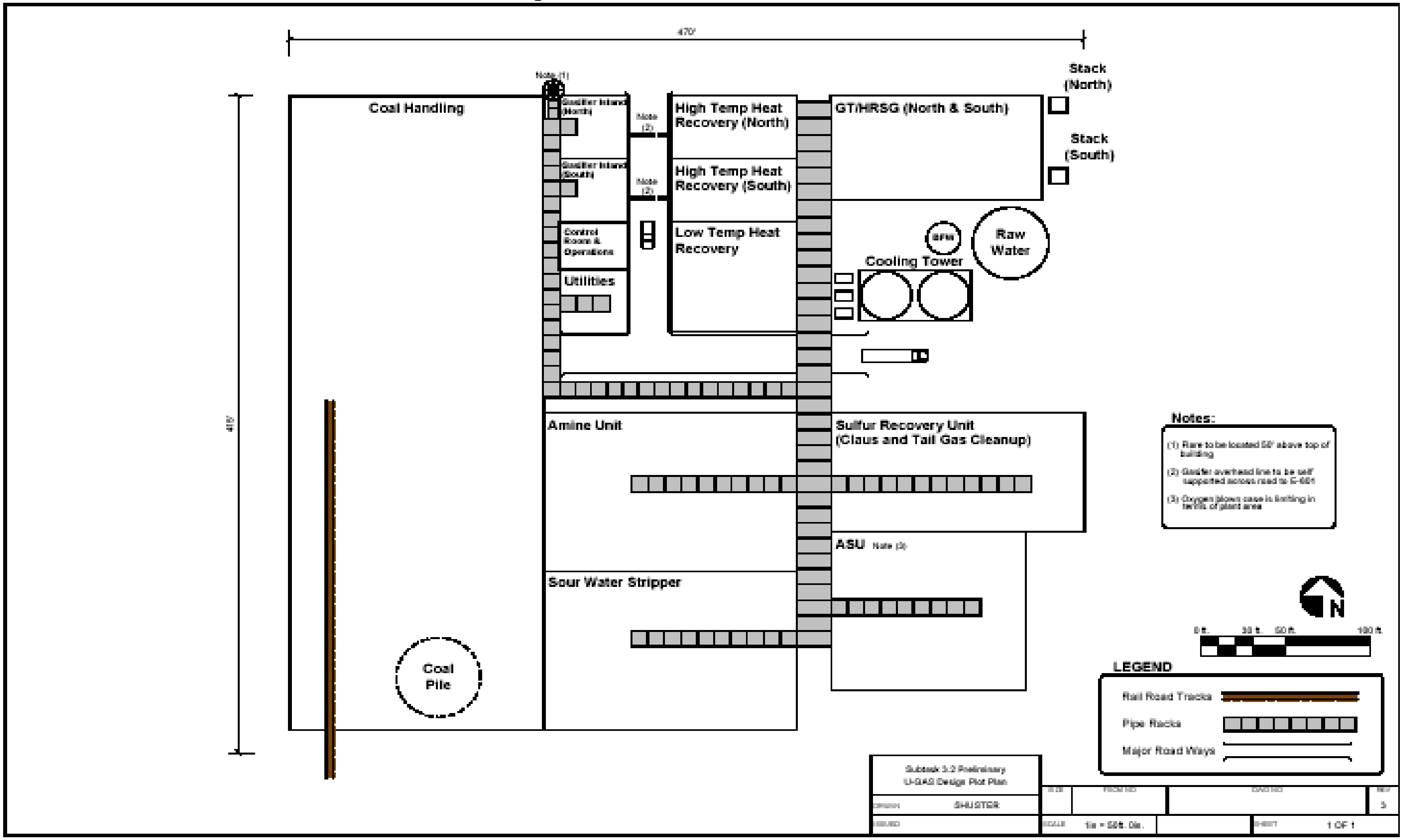
Proliferation – As more IGCC plants are built, their costs will decrease, availability will improve, and companies will be more willing to proceed with the construction of additional IGCC plants.

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

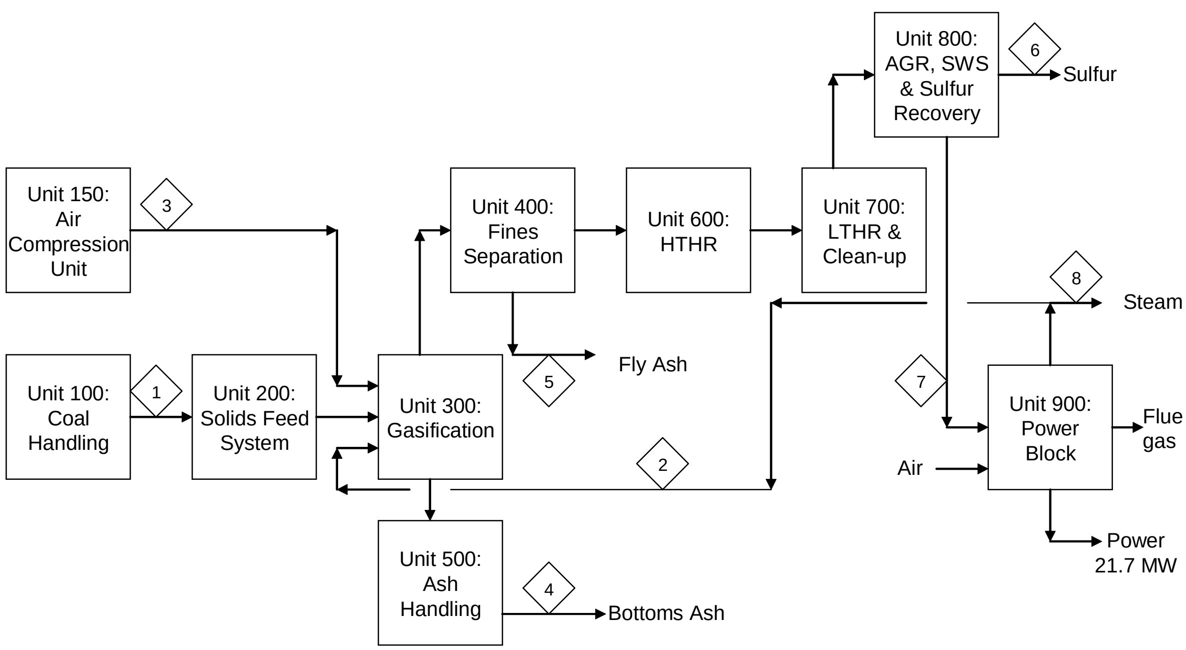


5.1 INTRODUCTION

The study compared the design for two alternative gasification scenarios for an industrial application in Subtask 3.2: an air-blown case and an oxygen-blown case. This section of the report describes the air-blown case. The oxygen-blown case is described in Section 6.

The overall material balance generated using ASPEN is shown in Table 5.1. The complete material balance is shown in the Addendum.

Figure 5.1 Overall Material Balance, Air-Blown 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

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5.2 PLANT CONFIGURATION

5.2.1 Coal Preparation/Handling (Unit 100)

The coal handling system receives and unloads coal from unit-train rail shipments delivered to the plant once/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/hr of fuel, which is transferred to a ready pile. Coal from the ready pile (designed for a one week inventory) is delivered via a reclaim hopper, vibrating feeder and conveyor to a crusher/dryer that prepares the as-received fuel to the gasifier feed specifications. Coal is to be sized to:

- No more than 2% > 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, used to store startup coke, is located next to the main fuel silo.

The Coal Handling System as described herein is a conceptual design. The proposed system will be located in a gasification plant in upstate New York. Southeastern Ohio coal will be delivered by railroad cars to the site, handled and processed for charging the two GTI Gasifiers.

The Coal Handling System starts at the unloading area where unit trains unload coal one car at a time to the under track hopper. The unloading area includes a 80 foot long x 30 foot wide x 20 foot high building which admits a 100 ton capacity rail car. The building is provided with wall mounted infra-red radiant heaters and track heaters for thawing car loads of frozen coal. The building includes a car shaker to loosen frozen coal and to provide effective unloading at the desired rate. The site will receive one train unit, consisting of a minimum of 24 rail cars, each car with 100 ton capacity for a total of 2,400 tons. The industrial facility (customer) is responsible for breaking the train into segments and spotting the coal cars at the unloading facility. No provision is provided for moving the rail cars at the gasification site.

The plant requires 363 tons per day at the silo discharge or a total of 1,815 tons for 5 days consumption. Coal delivery is made every 5th workday. The active pile requires 7 days storage or 2,541 tons based on a design rate of 363 TPD. Coal will be unloaded from the rail car at a rate of 300 TPH and transferred to the active pile storage by belt conveyor. This belt conveyor will include a metal detector and magnetic separator to remove and collect tramp irons. The stacker and reclaim conveyor will transfer the coal to the crusher inside the coal handling building. The 24 hour storage silo is approximately 300 feet from the railcar unloading area. The proposed coal handling building is 100 ft. long x 75 ft. wide x 60 ft. high. The building will contain the coal handling equipment, including the crusher and dryer units. The rest of the coal handling equipment after the vibratory screen discharge feeder will be located outside the building. The coke handling equipment from the delivery truck to the coke silo discharge feeder will also be located outside the coal handling building. All equipment located outdoors will be weather protected and tightly sealed to prevent dust leaks.

The coal specification calls for 98% passing a ¼" screen and no more than 10% less than 100 mesh (fines). The coal also must be dried to 5% surface moisture. To achieve these specifications, the crusher receives coal with the largest lump size of 2" x 0" and breaks the coal to the top size of ¼". The oversized coal is recirculated to the inlet of crusher until the ¼" top size is met. The ¼" sized coal and fines are then conveyed by screw feeder and bucket elevator to the dryer which utilizes plant steam supplied to a dryer heating coil at 400 psig pressure and at 550°F temperature. The steam heat of vaporization is transferred to ambient air-blown to the dryer fluidized bed by a forced draft fan to reduce the coal moisture content from a maximum of 15% to the specified 5%. A 15% moisture content of coal is used for dryer design. Normally the coal moisture content is expected to be less than 10%. The heating requirement for 15% moisture is 17 MBtu/hr, and for 10% about 8.5 MBtu/hr. It is estimated that the steam supply needed for drying 40 TPH to 5% moisture is estimated at 25 klb/hr and for 15.13 TPH it is estimated at 8 klb/hr. The steam supply line will be designed based on 300 feet of 6" nominal diameter pipe and condensate return of the same length with 2.5" nominal diameter pipe. This length of pipe will include expansion loops, vertical risers and vertical drops.

The dried coal is discharged to a vibrating screen where any coal greater than ¼" coal is separated from the fuel and recirculated to the crusher. To ensure that no more than 10% of the coal that is smaller than 100 mesh is fed to the gasifier the coal is applied to a 120-mesh screen. Fines passing through the 120 mesh screen are expected to be of a small amount and will be rejected and collected in a proposed dust collector. Adjustments in grinding can be performed if the quantity of fines becomes significant. If the fines quantity cannot be reduced, a pneumatic transport system can be installed to send the non-specification coal to the facilities' other boilers. This collection and conveying system will not be required if grab sample analysis indicates that total amount of fines are less than 10%, which means that all the coal discharging from the vibrating screen will be transported to the silo.

The dried coal is conveyed by a screw feeder and bucket elevator to the 24 hour primary silo for storage at the rate of 15 to 40 TPH. The coal silo is 32 feet in diameter with a cylindrical height of 42 feet.

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

The distribution screw feeder, which takes the coal or coke from either primary or redundant elevator, will discharge coal to the weigh feeder supplied by GTI at approximately 8 tons each. The distribution screw conveyor includes a grab sampling port before the first weigh feeder opening for coal analysis including moisture content and fines. If the fines exceed the 10% maximum limit, the primary silo will be scheduled for cleaning at a predetermined coal storage level. Cleaning will be performed in accordance with the plant maintenance procedure.

The coke handling system is designed for outdoor installation and is provided for gasifier start-up. Coke is delivered by truck inside the coal handling building and unloaded to a hopper, which feeds the belt conveyor for transferring coke to a bucket elevator. The bucket elevator takes the coke to the top of the 8 hour coke storage silo. The coke silo is approximately 14 feet in diameter by 32 feet cylindrical height. Coke is discharged at the bottom and is conveyed to the redundant elevator, which takes the coke to the common distribution screw conveyor for supplying the gasifier weigh feeders. The coke is fed to the gasifiers at 120 TPD.

The coal handling equipment from the active pile that discharges to the primary silo is designed for a maximum rate of 40 TPH. This rate will allow the plant to fill the primary silo during one shift (8 hours).

5.2.2 Air Compressor (Unit 150)

Two parallel Ingersoll-Rand air compressors, each with 50% capacity, compress the oxygen to the gasifier inlet pressure, 415 psia. Two rotary compressors, each having a flow rate of 48,050 lb/hr of air, 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 400 psig / 550°F steam before entering the gasifier.

5.2.3 Gasification Island

The gasification system is enclosed in a building with the two parallel trains located next to each other.

Each gasifier train consists of the following elements:

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

5.2.3.1 Coal Lockhopper Feed System (Unit 200)

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

5.2.3.2 Gasification (Unit 300)

The gasifier is lined with refractory to minimize heat losses. An outer layer is designed to minimize heat loss, and an 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 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 temperature exiting the gasifier when operated on bituminous coal is approximately 1850°F. The gasifier operates at 340 psig to provide adequate available pressure throughout the plant ahead of the gas turbine.

Air-Blown Operation

In air-blown operation the gasifier consumes 30,250 lb/hr (363 tpd) of dry coal that has a maximum of 5% surface moisture. Air and steam at high pressure are mixed and fed to the gasifier to react with the coal. At design conditions 96,106 lb/hr of air and 9,653 lb/hr of steam are required for gasification.

The product gas composition (mole basis) from the gasifier contains the following major components (re. Table D.2).

CO	20.72%
CO_2	6.88%
H_2	12.09%
H_2O	6.39%
CH_4	4.25%
H_2S	0.52%
COS	0.02%
NH_3	0.15%
HCN	0.02%
N_2	48.97%

Small quantities of light oils (primarily benzene), dust, chlorides, and mercury are also included in the gas stream and must be removed in the downstream cleanup system. Complete details are shown in the material balance in Table D.2 in Addendum D.

5.2.3.3 Startup

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

5.2.3.4 Dust Cyclones

A series of three dust cyclones increases carbon conversion and reduces the contaminant dust concentration in the syngas. Solids separated in the primary and secondary cyclones are recycled back to the gasifier to maximize carbon conversion and process efficiency. Dust collected from the tertiary cyclone is discharged via a lockhopper system to the dust collection and removal system.

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

5.2.3.5 Dust Removal System (Unit 400)

The dust removal system consists of a series of equipment to cool the dust and to transport it via lockhoppers 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. The screw rotates when the valve to the lockhopper is open and is stopped when the lockhopper valve is closed. A refractory lined surge hopper collects dust when the screw is not rotating (lockhopper closed). When the lockhopper is full (confirmed by nuclear level detectors) the upper valve is closed, the vessel 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.

5.2.3.6 Ash Removal System (Unit 500)

The ash removal system consists of a series of equipment to cool the ash and transport it via a lockhopper from the gasifier pressure to storage at atmospheric pressure. A pressurized cooling screw cools the ash from the high temperatures in the gasifier 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. A refractory lined surge hopper collects ash when the screw is not operating (lockhopper closed). When the lockhopper is full (confirmed by nuclear level detectors) the upper valve is closed, the vessel pressure is lowered to atmospheric, and the discharge valve is opened. Ash is then transported via a pneumatic system to a day tank from which it can be disposed or sold.

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

5.2.4 High Temperature Heat Recovery (Unit 600)

5.2.4.1 Introduction

The high temperature heat recovery system recovers the sensible heat from the syngas by producing saturated 425 psig steam, which is routed to HRSG to produce 400 psig superheated steam. The design objectives are to maximize the syngas sensible heat utilization, to maximize reliability, and to minimize operation difficulties. The syngas leaving the gasifiers contains particulates, light oils, chlorides, ammonia, etc. Each of these undesirables alone and in combination significantly impacts the design.

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

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

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

5.2.4.2 Design Basis

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

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

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

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

5.2.4.3 Characteristics of Raw Syngas

Table 5.1 lists the major characteristics of the syngas exiting the gasifier. The residual particulates in the syngas stream leaving the gasifier comprise ash, unburned carbon, and small amounts of trace elements. The oils produced in the U-GAS® gasifier are mainly benzene and naphthalene based compounds. Refer to Section 5.2.12 for a discussion of the light oils present in the wastewater.

Table 5.1 Major Characteristics of Syngas Leaving the Gasifiers

Temperature Exiting Gasifier (°F)	1750
Pressure (psia)	355
Mass Flow Rate (lb/hr)	133,280
Water (lb/hr)	6,200
Oils Condensation Temperatures (°F)	180 ~ 450
Dew Point (°F)	238
Ammonium Chloride Condensation Temperature (°F)	~ 540

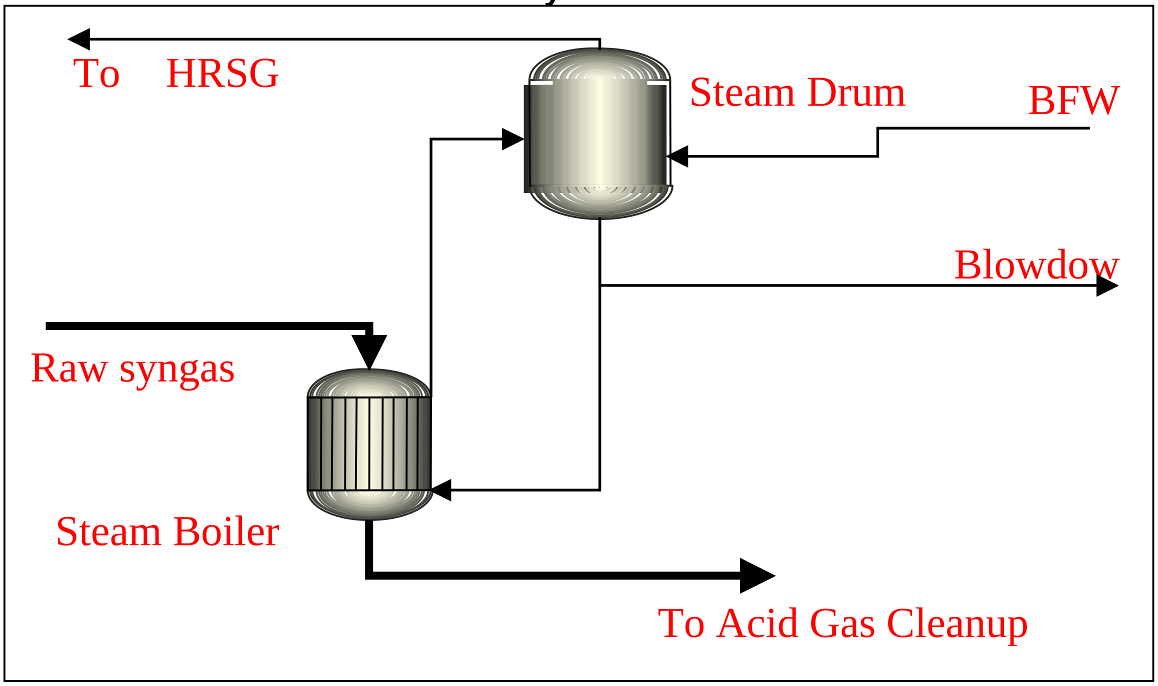
5.2.4.4 System Description

Figure 5.2 shows the schematic flow diagram for the high temperature heat recovery system, which comprises a steam boiler and a steam drum. The raw syngas 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. To prevent foreign matter from accumulating in the steam drum, a small percentage of liquid water is extracted from the steam boiler as blowdown, and it is sent to the wastewater treatment (WWT) unit. For startup, a small pump is required to establish the thermo-siphon flow pattern.

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 downwardly on the tube side while the water flows upwards on the shell side. The average syngas flow velocity in the tubes is about 30 ft/sec, and the overall heat transfer coefficient in the steam boiler is calculated to be about 65 Btu/hr-ft^2-°F. Inconel is recommended for the tubes for better erosion resistance. The heat exchanger has 100 tubes with a length of about 27 ft long and an inside diameter of two inches. To insulate the tube sheet and protect it from erosion, the gas side of the tube sheet is refractory lined, and the tube inlets are equipped with ferrules to minimize the thermal stresses on the tube sheet/tube joints. The steam boiler is large enough to have a 30 minute residence time. The equipment specification is included in Addendum B.

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

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



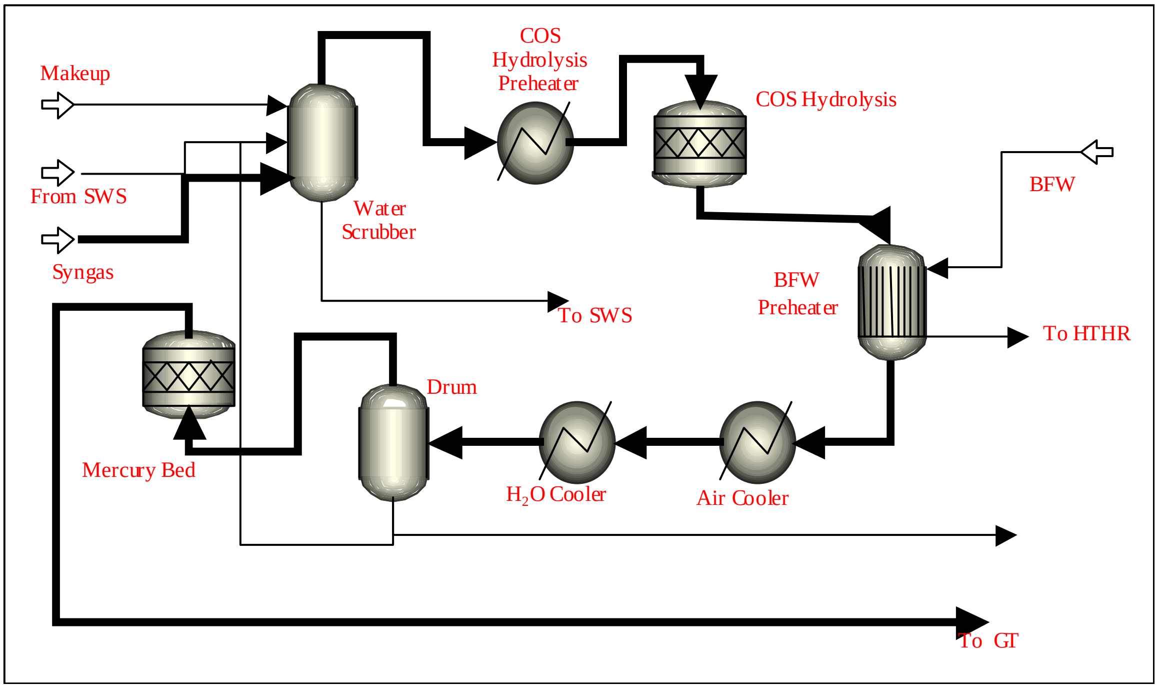
A marked benefit of this design is its simplicity. It minimizes the potential of plugging and damage to the equipment by the particulates. In the steam boiler, the syngas flows downwardly towards the bottom head, reducing the likelihood of tube plugging. In addition, the syngas exits the steam boiler at 600°F, preventing any complications stemming from the condensation of water, oils, and ammonium chloride. The simplicity of the design translates directly into being low cost. On the other hand, the particulates in the syngas stream may result in increased erosion in the steam boiler. Periodic cleaning of the steam boiler is recommended to remove deposits that accumulate on a regular basis (at least once per year). Experience indicates that these issues are at a manageable level and comparable to maintenance issues for traditional combustion systems.

5.2.5 Syngas Cleanup System (Units 700 and 800)

The syngas cleanup system removes particulates, ammonia, chlorides, oils, etc. from the syngas prior to sulfur removal in the acid gas removal system and combustion in the gas turbines. To ensure the proper operation of the acid gas removal system and the gas turbines, it is critical to remove the undesirables from the syngas such as the chlorides, particulates, ammonium chloride, etc. The design objective for the syngas cleanup system is to develop a system that is robust, reliable, and low cost.

The syngas cleanup system consists of two syngas scrubbers, a COS hydrolysis unit, and a low temperature heat recovery unit. Figure 5.3 shows the schematic flow diagram of the syngas cleanup system. The syngas streams from the two trains merge before they enter the COS hydrolysis preheater. The following discussion describes each of the main components in detail. A detailed equipment list can be found in Addendum B.

Figure 5.3 Schematic Flow Diagram of the Syngas Cleanup System



5.2.5.1 Syngas Scrubber

A number of technologies were considered for removal of contaminants from the raw syngas. The objective of the design study was to identify a system that effectively removes the contaminants that is cost effective, proven and reliable. Technologies considered include a wet-scrubber column, ceramic or metallic candle filters, baghouse, venturi scrubber, and electrostatic precipitator. A syngas wet-scrubber column is used at the Tampa Polk Power station to remove the particulates downstream of the high temperature heat recovery system. The use of a wet-scrubber was selected for the base design case. This is because a syngas wet-scrubber column can effectively remove the particulates and also remove chlorides, oils, and other gases such as ammonia. This is due to the intimate contact between the syngas and the water in the scrubber tower. The following discussion briefly describes why other equipment combinations were not selected.

Candle filters are used at the Wabash River plant; they are effective in removing particulates. However, separate processes are still needed to remove the oils and the chlorides present in the gas after filtration. This adds to the overall capital cost, since a scrubber would still be required.

Venturi scrubbers are widely used to remove particulates in similar applications; they are more economical in water usage compared to scrubber columns. However, oils condensed from the syngas may adhere to equipment surfaces because of the small water usage. This can lead to possible agglomeration of the particulates that would create separation problems in down-stream equipment.

Electrostatic precipitators are used on most coal-fired power plants for removal of particulates; however, they have not been proven for treating syngas. In addition, their high capital cost makes this choice less favorable.

The use of a baghouse was considered as an alternative to the wash column for removal of the particulates. The presence of light oils and the acid gases in the syngas may cause binding of the filters and may ultimately damage the fabric. In addition separate processes still are needed to remove the chlorides and oils.

Based on these considerations, syngas scrubber columns were chosen for this application, since the syngas wet-scrubber eliminates the need for a separate process to remove the chlorides and the oils. Note that it is assumed that the oils being purged into the sour water stream can be removed effectively in downstream systems, namely the sour water stripper (SWS) and the wastewater treatment unit before the water is discharged. Destruction of the organic compounds is required to meet environmental criteria since some of these compounds are water soluble. Removing them effectively from the water requires an aerobic treatment system. For this study an existing system is in place to handle this downstream function.

An impingement type of 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. Table 5.2 lists the water sources and the stream data related to the wash columns. The cleaned syngas leaving the wash column is saturated with water and contains twice the amount as was present entering the scrubber. All particulates, oils, most of the chlorides, and a part of the ammonia are removed from the syngas in the wash columns. The detailed stream data can be found in Addendum D.

Table 5.2 Water Balance in the Syngas Scrubber Column

	Inlets				Outlets	
	Syngas	Recycled Sour Water	Fresh Quench Water	Process Condensate	Cleaned Syngas	Sour Water
Temperature °F	600	110	80	110	265	265
Flow, lb/hr	6,200	24,000	15,249	5,874	12,138	39,168

The syngas exit temperature is dictated by requirements for the COS hydrolysis unit immediately downstream of the scrubbers. The manufacture of the COS hydrolysis catalyst recommends that the temperature of the syngas entering the COS hydrolysis reactor at 275°F. Also it is desirable to avoid any condensation of water in the COS hydrolysis unit to prevent degradation of the catalyst. Thus, the syngas passes through a COS hydrolysis preheater to raise its temperature to 275°F or 10°F above the scrubber discharge temperature of 265°F. The pressure inside the scrubber columns is about 345 psia.

For reliability and turndown considerations, the design calls for one syngas scrubber per train. Following the syngas scrubber, the washed syngas streams are combined into a large single train before entering the COS hydrolysis preheater. The use of two wash columns instead of one avoids complications due to the particulate erosion when one wash column is used, then equipment such as valves needs to be installed to regulate the flows of the two particulate laden syngas streams. The particulates in the syngas could render severe damages to the valves making a tight seal impossible. On the other hand, after the syngas is cleaned in the scrubber and the streams are free of particulates, they are much less likely to damage the flow regulation equipment if they merge. This allows complete and safe isolation of one gasification train for maintenance.

5.2.5.2 COS Hydrolysis Unit

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). For the air-blown case, 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 COS to H_2S to achieve 99% total sulfur recovery.

In a COS hydrolysis unit, COS reacts with water on the catalyst to produce CO_2 and H_2S. This reaction is slightly exothermic. To prevent catalyst degradation, it is desirable to keep water from condensing in the reactor. The syngas leaving the water scrubber at 265°F is saturated with water. A small heater is used to raise the syngas above its dew point. It is recommended that the syngas enter the unit at a temperature of 275°F, which favors the shifting of the hydrolysis reaction towards the formation of H_2S. The heater duty is about 0.47 MBtu/hr. A typical shell-and-tube heat exchanger is used to heat the syngas with 400 psig steam condensing on tube outer surface.

Süd-Chemie group is a specialist in the field of chemistry for surfaces comprising the most finely distributed inorganic matter and is one of the world's leading companies producing catalysts, adsorbents and additives. It has extended experience providing catalysts for COS hydrolysis. Based on the syngas compositions and flow rate, they provided a design recommendation using their catalyst. Detailed stream compositions around the COS hydrolysis reactor can be found in Addendum D.

5.2.5.3 Low Temperature Heat Recovery

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 of 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. The heated BFW then goes to the steam drum in the high temperature heat recovery system to produce saturated 425 psig steam. The syngas leaving the BFW preheater then 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. The advantages of using a combination of an air cooler and a water cooler are two fold: (1) it conserves water usage and, (2) it prevents scale build up on the heat exchanger tube surfaces. The syngas exits the BFW preheater at about 239°F. At this temperature if the syngas is cooled by cooling water, localized boiling may take place causing scale build up and fouling.

Table 5.3 lists the cooling capacity of each of these heat exchangers. Downstream of the third stage water cooler, a flash drum is used to separate the syngas from the condensate. During the cooling process, a substantial amount of NH_3 is dissolved in the process condensate. Half of the process condensate is routed to the syngas scrubber, and the other half is routed to the sour water stripper for NH_3 removal. Detailed compositions for all the streams can be found in Addendum D.

Table 5.3 Duties of the Heat Exchangers in the Low Temperature Cooling Section

Heat Exchanger	**Syngas Temperature (°F) Inlet - Outlet**	**Duty (MBtu/hr)**
BFW Preheater	275 - 239	5.56
Air Cooler	239 – 140	12.09
Water Cooler	140 - 110	2.04

5.2.5.4 Mercury Removal

Introduction

Mercury present in the coal will partition primarily to the syngas stream. The mercury content of the design coal used in this study is approximately 0.12 ppmw, based on USGS coal analysis data. The mercury concentration in the syngas for the air-blown case is about 40 µg/Nm^3, and the mass flow rate is approximately 0.0036 lb/hr. The mercury removal system is designed to achieve greater than 90% removal.

Basis

The technology used for mercury control is based on the equipment design at Eastman Chemical Company's Chemicals-from-Coal facility, which began operations in 1983. That facility employs carbon beds to remove mercury from cooled syngas. For Eastman Chemical Company, the purpose of the mercury removal is to protect the acetyl chemical product from any mercury contamination so consistent, high levels of mercury removal are required. Sulfur-impregnated activated carbon is used as the adsorbent in packed beds that operate at 86°F and 900 psi. Mercury removal of 90 to 95 percent has been reported with a bed life of 18 to 24 months. Eastman has 20 years of demonstrated vapor-phase mercury removal and has yet to experience any mercury contamination in its product.[3, 4]

Process Description

Calgon Carbon Corporation (www.calgoncarbon.com) provided the mercury control equipment design. The equipment consists of a single cylindrical adsorber vessel, 9 feet in diameter and 10 feet tangent to tangent. The vessel is packed with 20,000 lbs of Calgon Carbon HGR® sulfur impregnated activated carbon. The expected bed life is approximately 3-5 years.

Special Considerations

Mercury control using sulfur impregnated activated carbon is highly temperature dependent and requires a process temperature near 100°F.

Results/Conclusions

[1] The Cost of Mercury Removal in an IGCC Plant, Prepared for the US DOD/NETL, Parsons Infrastructure and Technology Group Inc., September, 2002.

[2] David L. Denton, Coal Gasification – Today's Technology of Choice and Tomorrow's Bright Promise, Presented at the AIChE – East TN Section Meeting, October 29, 2003.

[3] *The Cost of Mercury Removal in an IGCC Plant*, Prepared for the US DOE/NETL, Parsons Infrastructure and Technology Group Inc., September, 2002

[4] David L. Denton, *Coal Gasification – Today's Technology of Choice and Tomorrow's Bright Promise*, Presented at the AIChE – East TN Section Meeting, October 29, 2003.

The mercury control equipment is based on a commercially proven, reliable design. The equipment is expected to meet or exceed the design target of 90% mercury removal from in the syngas.

5.2.5.5 Acid Gas Removal and Clean-up

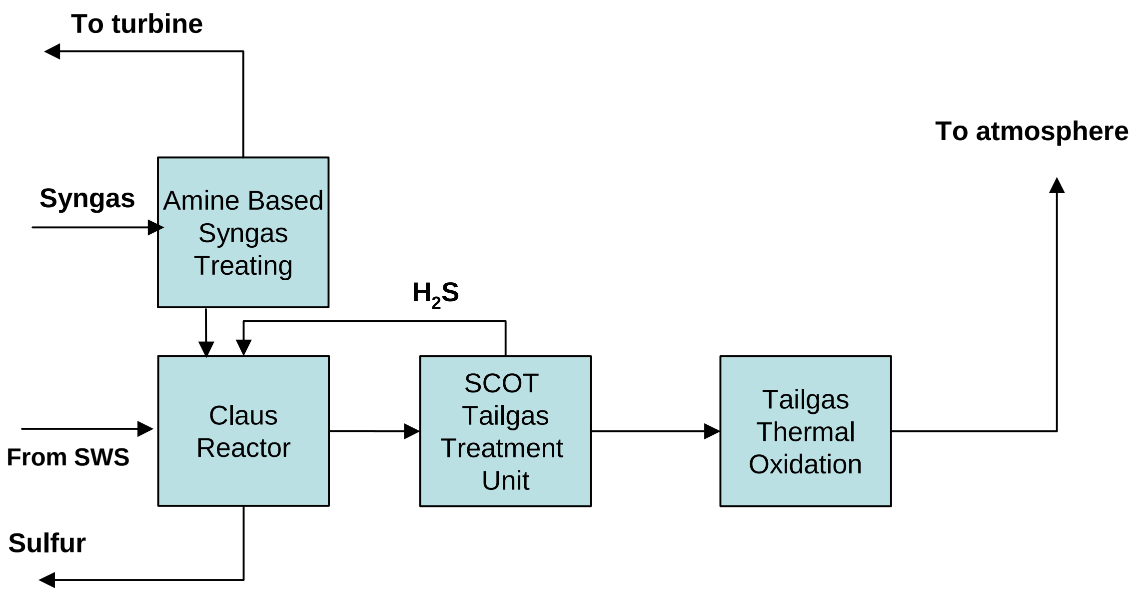
The syngas leaving the mercury drum is routed to an acid gas removal system to remove H_2S. Ortloff Engineers, Limited, who is a recognized leader in the area of sulfur recovery and sour gas processing plant designs, provided the design for this unit.

A gas treatment system features UOP's Selective AGFS process, which selectively removes most of H_2S, but allows most of the CO_2 and other species to remain in the syngas stream. By allowing most of the CO_2 to slip through the system, the sizes of the downstream gas cleaning equipment can be made smaller than otherwise possible with other process designs. This reduces the capital and operating costs associated with this system. Figure 5.4 shows the block diagram of the recommended design. The amine based acid gas removal unit consists mainly of an absorber and a regenerator. The treated syngas then flows to the gas turbines.

The acid gas stream leaving the regenerator can be converted either into sulfuric acid or elemental sulfur. Based on demonstrated performance on syngas and on the required scale of production (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 an amine system. It is then routed back to the Claus reactor. Note that the sour gas (HCN, CO, CO_2, H_2S, NH_3, etc.) collected from the SWS also is treated in this system to recover any sulfur in the sour water. This results in very high overall sulfur recovery, on the order of 99.8% or higher. The elemental sulfur produced in the Claus reactor can be sold as a source of revenue or disposed in a landfill.

The treated gas leaving the SCOT unit then is incinerated in a tailgas thermal oxidation (TTO) unit before being released to the atmosphere. Natural gas is used in the TTO to incinerate the effluent, and a waste heat recovery system is included in the TTO to generate high pressure and low pressure steam. These steams along with the steam generated in the Claus reactor are used in the reboiler of the amine stripper. The vent gas is dispersed to the atmosphere at about 550°F to avoid any condensation of SO_2.

Figure 5.4 Block Flow Diagram for the Acid Gas Removal System



5.2.6 Power Block - Gas Turbines and HSRG (Unit 900)

5.2.6.1 Introduction

The primary purpose for the gasification of coal in this application is to generate electric power and steam. Two General Electric (GE 10) combustion turbines (CT) were selected as the basis for this study. The design power output with natural gas for these engines is 11.25 MW at ISO conditions. Steam is generated with the exhaust from each turbine in a two-pressure heat recovery steam generator (HRSG). The power block consists of two parallel CT/HRSG trains.

5.2.6.2 Basis

The power block is designed around the two CT's. The exhaust gas exiting the CT is routed through the associated two-pressure HRSGs. The individual HRSGs are designed such that three specific process conditions are met:

- Stack temperature remains above the acid-dew point so that condensation and corrosion does not occur within the system
- 50 psig superheated steam (~353°F) is generated such that the process steam demands of all gasifier and gas clean-up processes are self sufficient (including gas clean-up operations and sour water treatment)
- Balance of steam generation is 400 psig superheated steam (~550°F)

5.2.6.3 Process Description

Clean syngas is sent to the CT at 120°F and 295 psia. Figure D.5 in Addendum D illustrates the CT/HRSG.

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 phenomena is sometimes referred to as the “syngas boost”. Modeling for this study 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. See “Special Considerations” below for additional discussion of CT modeling and performance.

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

Flue Gas Flow – Exhaust gas exiting the CT (about 890°F, 15.6 psia) 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 (~40,000 Lb/hr) entering the 50 psig evaporator is extracted as liquid water and sent to the 400 psig evaporator. The remaining 1/3 of the water (~19,000 lb/hr) 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 38,000 Lb/hr of 400 psig saturated steam is generated in the evaporator. Approximately 4% of the inlet water mass flow (~1,500 lb/hr) 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, producing approximately 63,000 lb/hr of 400 psig superheated steam (~550°F).

5.2.6.4 Special Considerations

The firing of low- to medium- Btu coal derived syngas in a combustion turbine will result in performance different than the combustion of natural gas or syngas from non-coal feeds. The only reliable way to estimate turbine performance is to have performance testing conducted by the manufacturer. The turbines selected for this application have had some prior testing done for coal derived syngas of similar composition and quality. However, data is lacking, especially in the performance curves for modeling CTs in heat balances such as the GateCycle program used here. A more detailed discussion of the modeling of the power block is included in Addendum A.

For CTs to be able to burn different fuels (e.g., natural gas, coal derived syngas), appropriate burners must be designed. While GE has conducted prior work on low-Btu syngas burners for the GE 10 turbine, the current availability of such burners is not clear. Communication with GE has indicated that should the application of this study move forward, additional performance testing and perhaps burner development would be necessary. Costs associated with those activities have not been included in the economics of this study.

Many IGCC designs employ the use of air extraction from the GT compressor as the initial stages 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 the base case design for two reasons. First, it adds increased plant complexity and poses integration issues that were considered too complex for this level of study. Second the GE 10 turbine has not been thoroughly evaluated for extraction air.

Furthermore, the presence of trace amounts of H_2S in the syngas, as well as some trace level impurities may necessitate the use of special materials, either in the CT, HRSG, or both. Those needs have not been assessed or estimated in this study.

5.2.6.5 Results/Conclusions

For this application, two parallel CT/HRSG trains were designed based on turbine availability and system reliability. The CTs are GE 10s and the HRSGs are designed to provide self sufficient quantities of 50 psig superheated steam. The balance of steam production by the HRSG is 400 psig superheated steam exported to the industrial facility for process consumption. Table 5.5 summarizes the net output of the power block. Table 5.4 does not include the power and steam demands of the gasifier operations.

Table 5.4 Power and Steam Net Output (Power Block only)

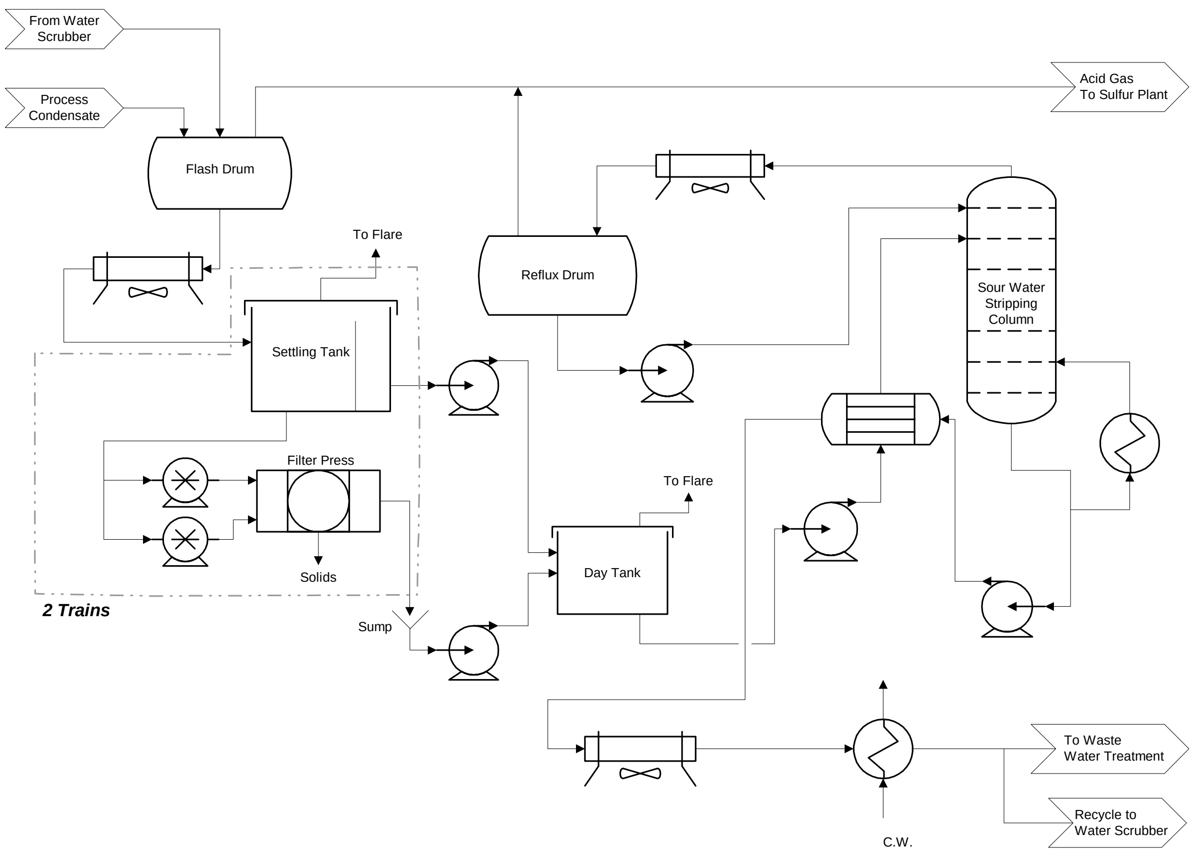
	Single Train (1 of 2)	**Combined Trains (2 of 2)**
Power Output	14.9 MW	29.8 MW
50 psig Superheated Steam Generation	19,123 Lb/hr	38,246 Lb/hr
400 psig Superheated Steam Generation	62,908 Lb/hr	125,815 Lb/hr

5.2.7 Sour Water Treatment Unit (Unit 800)

5.2.7.1 Introduction

Sour water containing ammonia, hydrogen sulfide, residual particulates and other low level impurities is sent to the sour water treatment unit. The sour water treatment unit processes the effluent from both syngas water scrubbers and the process condensate from the flash drum upstream of the amine system. The sour water treatment unit consists of a flash drum, settling tank, a day tank, a sour water stripping column, and associated heat exchangers and pumps. The settling tank removes particulates and insoluble oils. The filter presses dewater the agglomerated sludge from the bottom of the settling tanks. The day tank provides for water storage during stripper outages. Vapors from the flash drum and stripping column are sent to the sulfur plant. Stripped water from the bottom of the column is recycled to the syngas scrubber with a blowdown stream sent to the wastewater treatment plant. Any vapors emanating from the settling tank and day tank are sent to the flare. Figure 5.5 is a schematic flow diagram of the sour water treatment unit for the air-blown case.

Fig 5.5 Sour Water Treatment Unit Schematic



5.2.7.2 Basis

The sour water stripper for the air-blown case is a two-stage unit consisting of a flash drum followed by a distillation column. Most of the volatiles exit the flash drum in the vapor phase and are routed to the sulfur plant. The liquid stream from the flash drum that contains significant amounts of ammonia and hydrogen sulfide is sent to the stripping column. The stripping column produces a liquid effluent stream containing no more than 50 ppmw ammonia and less than 10 ppmw hydrogen sulfide.[5] The resulting stream is then cooled to 110°F before either being recycled to the syngas water scrubber or being blowdown to the wastewater treatment plant.

5.2.7.3 Process Description

Sour water from the water 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 meet the process requirements of the settling tank. This additional cooling reduces 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, 5–10 µm, a flocculent is recommended to agglomerate the small particles and increase their settling velocities. The design basis was adopted from the successful operation at the Polk Power Stations. 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 may also be used. Other chemical treatment 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 will be used to transport the slurry from the bottom of the settling tank to the filter press. Filter presses 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 is typically 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

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

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 plant to prevent the buildup of any non-volatile impurities within the system, and the remainder is recycled back to the syngas water scrubber. The specific design information and simulation results are included in Addendum D.

5.2.7.4 Special Considerations

The settling tanks were designed for atmospheric pressure to avoid pressurized vessels and to reduce the cost of the tanks. An air fin cooler is used to cool the incoming sour water to condense all of the H_2S. This method is used to avoid the need for a compressor to compress the corrosive H_2S vapor to the flare. The cooler reduces the vapor in the tank, leaving only a small amount of non-condensable, which are sent to the flare.

Because of the corrosive nature of hydrogen sulfide, stainless steel or stainless steel cladding was specified for the material of construction.

Additional design considerations included a study to minimize the water content of the vapor stream sent to the sulfur plant. It is recommended that the water vapor not exceed 5 percent of the gas stream entering the acid gas stripper column.

Past experience with lower temperature fluidized bed gasification systems have demonstrated the presence of various light oils, including water-soluble phenols, in the raw gas.[6,7] The U-GAS® technology used here has been characterized as having a very low light oil content.[8] However, recent data on U-GAS® light oil generation for the specific coal is lacking, and pilot scale testing should be conducted to gather design data and to determine solubility data of low-level oils and trace organics for the specific design coal. Furthermore, research is recommended to determine the fate of such organics, primarily for the more soluble compounds that may end up in the water scrubber discharge and process condensate when processed in aerobic treatment processes.

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

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

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

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

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 green field 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 green field system has not been included in this study.

5.2.7.5 Results/Conclusions

The sour water feed rate was slightly more than 45,000 lbs per hour (~90 gallons/minute). The resulting stripper column was designed with 21 stages (including the condenser) and a tray spacing of 2 feet. The column diameter measures 3 feet. It is constructed of stainless steel clad carbon steel with stainless steel internals. Additionally, the sour water flash drum, column distillate drum, and overhead condenser are constructed of carbon steel with stainless steel cladding. Stainless steel also was selected for the tube side of the stripper feed preheater, but the shell is carbon steel. The product recycle water has a design ammonia concentration of less than 50 ppmw, and the H_2S and COS concentrations each are less than 1 ppmw.

5.2.8 Offsites/Utilities (Unit 1000)

The outside battery limits (OSBL) facilities consist of sections provided to support the gasification units in terms of utilities and other auxiliary facilities. A list of these units, a general description, and the basis used for determining their capacities are listed below.

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

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

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

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

5.2.8.1 Steam System

Two levels of steam are provided within the onsite facilities. Both high pressure (400 psig) and low pressure (50 psig) are generated onsite. No additional steam generation facilities are required in the OSBL. For a greenfield plant site a start-up boiler will normally be required (10 thousand pounds per hour at 400 psig, superheated to 550^oF). Since most industrial facilities will have existing boiler capacity as in this brownfield site, it was assumed that steam for the start-up can be obtained from the existing plant. The steam production and consumption are shown in Table 5.5.

Table 5.5 Steam Production/Consumption

(thousand lb/hr)

High Pressure (400 psig/550°F) Steam		
	Production	130.7
	Consumption	29.0[1]
	Export	101.7
Low Pressure (50 psig/350°F) Steam		
	Production	38.3
	Consumption	38.3
	Export	0

[1] Design maximum is 40.0 thousand lb/hr and corresponding export is 90.8 thousand lb/hr

5.2.8.2 Condensate Collection System

Condensate polishing and deaeration of boiler feed water will be performed at the existing boiler house of the industrial site. Therefore, a condensate collection system is required to collect condensate produced in the new plant and transfer it to the existing industrial facility. Once processed, the treated and deaerated water will be returned to the new plant as fresh boiler feed water. A small storage tank will maintain about one hour of storage, and this amount is sufficient to fill the boiler during startup.

The system handles condensate at two pressures: 50 and 400 psig. The condensate flowrates are listed in Table 5.6. The condensates are sent in two separate lines to the existing industrial facility where they are polished and deaerated.

Table 5.6 Condensate Basis

50 psig condensate flow rate	38,200 lb/hr
400 psig condensate flow rate	30,300 lb/hr
Total condensate flowrate	68,500 lb/hr = 137 gpm
BFW requirement	242,200 lb/hr = 484 gpm
Deaerated water storage tank	32,000 gallons

Deaerated water from the existing industrial facility is received in a deaerated water storage tank located at the new plant. This storage tank has the capacity to hold 1 hour requirement of the deaerated water under the assumption that the industrial site has a larger amount of holdup upstream. In order to prevent the deaerated water in the storage tank from absorbing oxygen from the air, a small amount of steam is continuously injected at the bottom of the storage tank. Water from the deaerated water storage tank is pumped by boiler feed water booster pumps (one operating, one spare) to the suction of the boiler feed water pumps.

5.2.8.3 Cooling Water System

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

Table 5.7 Cooling Water Basis

Cooling water normal flow rate	2,810 gpm
Cooling water maximum flow rate	3,650 gpm
Inlet temperature	80°F
Outlet temperature	100°F
Pump discharge pressure	50 psig

5.2.8.4 Safety Shower / Eye Wash System

The safety shower and eye wash system consists of a safety shower water tank, pump and a heater/cooler to keep the water in circulation at near ambient temperature. The potable water system for the industrial site gasification facility also provides a continuous and sufficient quantity to the emergency shower and eye wash (ESEW) stations.

5.2.8.5 Raw Water / Fire Water System

The raw water system receives raw water from the existing industrial facility and stores it in the raw water/fire water storage tank located in the new plant. The water received from the existing industrial facility is lake water that has been filtered and chlorinated. Water from the raw water/fire water storage tank is pumped by the cooling tower make-up pumps to the cooling tower, and with the firewater pumps to the fire fighting users. Adequate water quantity is assumed to be available at this site. The raw water / fire water basis is shown in Table 5.8.

Table 5.8 Raw Water / Fire Water Basis

Raw water consumption by facility	115,000 gal/day
Firewater consumption in 4 hrs	120,000 gal/day
Storage tank capacity	250,000 gallons

5.2.8.6 Drinking (Potable) Water System

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

The function of the potable water system is to distribute potable water (supplied by the city) to various areas inside the industrial gasification site. The potable water system is adequate to provide a continuous and sufficient quantity to the plant bathroom facilities, drinking fountains, emergency shower and eye wash (ESEW) stations and various sinks (lab, maintenance, control room). The potable water provided by the city is estimated at 900 gallons per day with a peak demand requirement of 20 gpm.

Potable water shall be supplied to the following areas of the industrial gasification facility:

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

5.2.8.7 Compressed Air System

The compressed air system provides oil free compressed air while maintaining a minimum pressure of 100 psig in the distribution headers. Each individual compressor (total of 2) is capable of providing 600 scfm at discharge pressure of 125 psig to the dryers. It is assumed that the system will be interconnected and backed up by the industrial site air supply. The compressed air is dried to a dew point of –40°F using heatless desiccant dryers. The compressed air basis is shown in Table 5.9.

Table 5.9 Compressed Air Basis

Instrument air flow rate	1,000 SCFM
Service or plant air flow rate	120 SCFM
Total design air flow rate	1,200 SCFM
Compressed air pressure	100 psig

5.2.8.8 Natural Gas System

Natural gas will be supplied to the facility from the main natural gas header from outside the complex. Natural gas will be used as fuel for the Claus process and for the pilot in the flare. Two knock out drums will be provided in the system.

The natural gas system is designed to provide natural gas as a start up fuel for gasifier, as fuel to the Claus Plant, and as a primary fuel for the auxiliary boiler if required. Typically a natural gas pressure of 31 psig is required.

5.2.8.9 Flare System

Specific design philosophies and instrumented control systems are usually employed in Gasification plant designs to mitigate certain relief scenarios and reduce the load on the flare. Such detailed design load calculations will be part of basic/detailed engineering for the specific facility. For purposes of this study, the scope of facilities includes a flare with a design capacity of 140 MBtu/hr (the syngas produced by one gasifier when one gas turbine is lost). This would be a steam-assisted flare with natural gas being used as pilot fuel.

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

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

5.2.8.10 Nitrogen System

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

- A dedicated nitrogen system for onsite process applications
- A general purpose nitrogen system for all other applications

The two systems will be independently piped from the source.

Subsystem 1 consists of two liquid storage tanks followed by one vaporizer that vaporizes the liquid nitrogen drawn from the storage tanks. The nitrogen is vaporized by atmospheric (ambient) heat. The vaporized nitrogen is then sent to the nitrogen users in the gasification island. Because the liquid nitrogen tanks are under pressure, the subsystem does not need any pump for transferring liquid nitrogen from the tanks to the vaporizer.

Subsystem 2 consists of one liquid storage tank followed by one vaporizer that vaporizes the liquid nitrogen drawn from the storage tank. The nitrogen is vaporized by atmospheric (ambient) heat. The vaporized nitrogen is then sent to the nitrogen users in the general plant area. Because the liquid nitrogen tank is under pressure, the subsystem does not need any pump for transferring liquid nitrogen from the tank to the vaporizer.

Nitrogen requirements will be provided from a liquid nitrogen tank and evaporators. The liquid nitrogen storage and evaporation system consisting of the nitrogen unloading facility, liquid nitrogen tank, evaporators and associated controls will be leased from a standard manufacturer who supplies liquid nitrogen.

A larger quantity of nitrogen will be required during initial start-up, which will be made available from a road tanker mounted storage vessel and evaporating system. The start-up nitrogen will be routed through the two nitrogen systems explained earlier. The nitrogen basis is shown in Table 5.10.

Table 5.10 Nitrogen Basis

Sub-System 1, for Gasification Island:		
	Design flowrate of N_2	14,200 scfh
	N_2 supply pressure	450 psig
Sub-System 2, for General Plant Use:		
	Design flowrate of N_2	900 scfh
	N_2 supply pressure	50 psig

5.2.8.11 Wastewater Collection, Treatment and Disposal System

All the rainwater falling on non-contaminated areas of the industrial gasification facility is allowed to rundown into storm water drains. Any wastewater and potentially contaminated storm water is collected and sent to the existing industrial facility for wastewater treatment.

Wastewater sumps are located at the ends of the gasification plant area for collection of the wastewater. Wastewater from the water seal drum in the flare system is collected in a dedicated sump and pumped to the two gasification unit sumps. The wastewater in the gasification unit sumps is pumped by the sump pumps to the final wastewater transfer sump. The final wastewater transfer sump also receives boiler blowdown and cooling tower blowdown.

The water in the final wastewater transfer sump is pumped with the final wastewater transfer pump to the existing industrial facility for wastewater treatment.

There are waste streams generated in the facility, which will be collected, conveyed and treated, as necessary, before disposal. The following has been considered for this study.

Non-contaminated surface water

All non-contaminated rainwater falling on non-contaminated areas is allowed to run down into storm water drains, which are connected to the area drainage system by gravity.

Potentially contaminated wastewater system

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

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

Wastewater from flare system

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

Final wastewater disposal sump

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

5.2.8.12 Electrical Distribution

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

Table 5.11 Electrical Basis

Gross power production	29.83 MW
Internal power consumption	8.13 MW
Power export	21.70 MW

5.2.8.13 Miscellaneous Facilities

Interconnecting Piping

The system consists of the six lines that transport liquid or gas streams from the existing industrial facility to the OSBL or vice versa.

The following six lines are in the interconnecting piping system:

• 50 psig condensate	OSBL to existing industrial facility
• 400 psig condensate	OSBL to existing industrial facility
• Boiler feed water	Existing industrial facility to OSBL
• Raw water	Existing industrial facility to OSBL
• Natural gas	Existing industrial facility to OSBL
• Wastewater	OSBL to existing industrial facility

Pipe Racks

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

Roads

Adequate roads will be incorporated to suit the layout of the facility in the scope.

Site Development

The site is a reasonably flat piece of land. In the absence of a survey map for the proposed plot, a provision has been allotted for site development in the estimate. The flood level, has not been established at the proposed site. It is assumed that the plot is above flood level, and no provision has been made in the estimate in this regard. No piling is assumed to be required.

Miscellaneous Works

Other miscellaneous works in this category included in the scope of facility are equipment foundations and the two wastewater collection sumps.

Buildings

Synergies with the existing facilities are assumed to be maximized, and the buildings are, therefore, limited to process buildings for the gasification island, turbines, coal handling areas, and an onsite control room/plant office building. The former also contains any electrical and utility facilities rooms, as necessary.

5.3 EMISSIONS

Gasification systems are inherently less polluting than combustion systems because the pollutants (sulfur, mercury, chlorine, and others) are removed from the syngas before it is sent to the combustion turbine. Pollutant 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.

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

5.3.1 Sulfur

Sulfur is removed from the syngas by a two-step process. First the syngas is passed over a COS hydrolysis catalyst at 275°F to convert the COS to H_2S. The H_2S is removed from the syngas by UOP's Selective AGFS (Acid Gas removal – Formulated Solvents) process that was designed by Ortloff Engineers, Ltd. This combination has a sulfur removal rate of greater than 99.8%. The sulfur is recovered as elemental sulfur by a Claus process with a Shell Claus Off-gas Treating (SCOT) process, where the residual sulfur compounds are converted back to H_2S and subsequently captured. The combined SO_2 release rate from the gas turbine and the incinerator is 5.1 lb/hr or 0.013 lb per MBtu (HHV) of energy input. For the oxygen-blown case, the combined SO_2 release rate is 5.0 lb/hr or 0.013 pounds per MBtu of energy input. The net result of this processing scheme is an overall sulfur removal rate of 99.7%.

5.3.2 NOx and CO

The firing of combustion turbines on coal-derived syngas requires the proper design of turbine components. The specific design can influence the emission rates of NOx and CO. For this application, 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/MBtu – 0.35 lb/MBtu). A prior CO emissions estimate for the GE 10 was 20 ppmvd @ 15% O_2 (~0.04 lb/MBtu). Including balance of plant emissions (e.g., incinerator off gas and flare), total facility CO emissions are estimated at less than 0.05 lb/MBtu.

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/MBtu 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.[13] 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

[13] 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.

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 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.[14] 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.[15]

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, sulfur recovery unit tail gas incinerator, the flare system, and possible fugitive emissions from equipment leaks.

Most of the upstate New York area is classified as attainment[16] 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.

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

[15] 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

[16] 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

While specific emission levels for the application under study would be site specific, for comparison carbon monoxide emission limits included in the operating permits for TECO Polk Power Station and the Wabash River Repowering Project were 0.392 lb/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.

5.3.3 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 for both the air and oxygen-blown cases are estimated at 0.00036 lb/hr (0.95 lb/TBtu) and 0.00034 lb/hr (0.87 lb/TBtu), respectively. 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.

5.3.4 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 green field 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 green field system has not been included in this study.

5.3.5 Emissions Summary

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

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

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.

5.4 TRADE-OFF STUDIES

A variety of trade-off studies were examined as part of this subtask. The objective for this subtask was to provide a simple and reliable facility, which had a major impact on the evaluation of the trade-off studies. The trade-off studies are summarized below.

- Use a low pressure gasifier and compress the syngas going to the gas turbine. This idea was rejected as impractical.

- Consider alternative scrubber inlet temperatures of 450°F and 500°F for higher thermal energy recovery. The issue of corrosion problems caused by possible HCl condensation precluded these options. As the design progressed it was noted that 600°F would be better since it would eliminate possible precipitation of ammonium chloride. This idea was accepted based on these technical reasons.

- Compare a bag house versus candle filters for dust removal from the syngas. Rejected this idea for technical reasons – a bag house with light oils would be messy and a safety hazard. The candle filters were eliminated from consideration for a more conservative scrubber design.

- Generate electricity in a steam turbine with excess 50 psig steam. This idea was rejected because the goal is to make 400 psig steam for the industrial site and there is no excess 50 psig steam available.

- Superheat all the steam in the HRSG. Accepted for technical reasons. By placing the steam boiler first maximizes steam production and simplifies the cooling system.

- Use 50 psig steam instead of 400 psig steam for preheating the COS reactor feed. This was rejected based on the pressure difference in the case of a tube breach. That is, with 50 psig steam there is the possibility of contaminating the steam system with syngas (operating at over 300 psig) if there were to be a leak.

- Place a 400 psig steam superheater as the first cooler in the syngas cooling train. Rejected for technical reasons. There exists a need to keep the tubes cool for durability. In addition it was decided that all the superheating of steam would take place in the HRSG for simplicity.

- Move the 50 psig steam superheater after the 400 psig steam coil in HRSG. Accepted this idea since it reduces the total surface area required in the HRSG.

- Consider an alternate media for coal drying. Will use steam for heating the air since it is assumed that over the long-term natural gas is a premium fuel at the site.

- Use the flue gas heat from the HRSG for coal drying. Since the flue gas will be high in moisture and near its dew point, this was not considered practical. It was decided to use indirect steam heat as the source of heat for this application.

- Remove the 50 psig steam generation from the syngas cooling train. Accepted this idea since this simplifies the syngas cooling water design.

- Utilize the heat from the intercoolers of the air compressor. Rejected this idea because it adds unnecessary integration complications. Furthermore, operating problems could result in damage to the expensive air compressors.

- Use a thermosyphon reboiler for circulation around the 400 psig steam drum in the syngas cooling train. Accepted this idea because this provides a safer, simpler, more reliable system. A small circulation pump is provide for start-up.

- Use two sour water strippers (one for each gasifier train) to increase availability. Rejected this idea and replaced it by a simpler, less expensive alternative (holding tank with one day capacity).

- Combine steam generation in the syngas cooling area by using a high pressure water pump around loop and then flashing to low pressure. Rejected this idea due to cost and final design considerations of system.

- Evaluate alternative configurations to supply high-pressure air to the gasifiers. Main air compressor replication for availability (i.e., 3 @ 50% versus 2 @ 100% versus 2 @ 50% versus 1 @ 100%). Rejected this idea since compressors are expensive and will not payout (use 2 @ 50% for operating flexibility).

- Use two sets of cyclones from each gasifier for increased performance. Rejected this idea as impractable.

- Use bleed air from the first stage of the main air compressor to transport solids into the truck loading hopper. Rejected this idea as impractable.

- Use syngas to preheat the gas turbine feed. Rejected this idea for technical reasons due to the possibility of allowing dirty syngas into the clean turbine feed.

- Use sour water stripper bottoms water (i.e., stripped water) as the feed for the syngas scrubber. This is an acceptable idea that targets a goal of minimizing water make-up, but requires a significant purge stream to remove unknown materials form the system. Operating experience could allow reduction of the purge stream.

- Evaluate a gas turbine feed preheater. Rejected this idea because the fuel savings is offset by the lower steam production.

- Balance 50 psig steam production with demand, and if extra 50 psig steam is needed, let down 400 psig steam on a short term basis. This idea was accepted since the industrial site does not have 50 psig steam available, and the gasification unit should be self sufficient.

- Use 50 psig steam drives on the air blowers. Rejected this idea because it is more economical to make and export 400 psig steam.

- Use only one startup heater for two gasifiers. Rejected this idea because the savings from eliminating one heater will be less than the cost of the additional high-temperature valving and manifold.

- Enclose the gasification island within a building (versus an open structure around the gasification system). Accepted this idea since the climate around the industrial site in upstate New York is severe in winter, and an open structure could reduce the plant availability.

- Co-firing syngas in the HRSG when one turbine is out of service. Rejected this idea due to the stack temperature limitations and also because the expected occurrence of this operating state is low, gasifiers are capable of 50% turndown (no need to flare or shutdown a single gasifier train during GT/HRSG outage), and the desire to keep process simple by avoiding interconnections and instrumentation.

Numerous trade-off studies that were rejected for Subtask 3.2 will need to be revisited during the optimization portion of Subtasks 3.3 and 3.4.

5.5 PLANT COSTS

5.5.1 Basis

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

For this evaluation all the investment costs are for the second quarter 2004 at 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.

5.5.2 Methodology

5.5.2.1 Equipment Design

The equipment for both the air- and oxygen-blown cases were designed using the material and energy (M&E) balances developed specifically for Subtask 3.2. Various groups developed the M&E balances. Raymond Professional Group (RPG) developed the coal handling and preparation area. GTI developed the gasification island. Ortloff Engineers, Ltd. (Ortloff) prepared the sulfur removal and recovery systems. Nexant and NETL developed the remainder of the ISBL facilities and the balance of the plant (BOP).

RPG and GTI provided process flow diagrams (PFD’s) for their portion of the study. The BOP process flow diagrams were developed based on the ASPEN and GATECYCLE computer simulations and previous experience with similar systems. The M&E balances and PFD’s are shown in Addendum D.

Using the M&E balances and PFD's established the operating and design conditions for the individual pieces of equipment pieces. The equipment was then sized and material were selected to provide a 20-year life. RPG provided the equipment list and sizing for Unit 100. The equipment sizing for Units 200-500 was prepared by GTI. The design for the equipment in Units 150 and 600 through 1000 (excluding the sulfur removal and recovery system) were prepared by Nexant and NETL using the ASPEN and GATECYCLE heat and material balances as a basis. The sulfur removal and recovery system design was provided by Ortloff. The equipments lists are provided in Addendum B.

5.5.2.2 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 costs at 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. In the same way 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, gas turbine, and air compressor package unit.

A third approach was to request quotes for the installed cost of complete units. This method was employed for the coal preparation and handling unit (from Raymond Professional Group, RPG), HRSG, and the sulfur removal and sulfur recovery units (from Ortloff Engineers, Ltd.).

The capital cost estimate for the various buildings was handled separately. The gasifier structure building, control building and turbine building were estimated using PDQ$. The coal handling building was estimated by RPG and is included in the coal prep and handling.

5.5.3 Results

Table 5.12 shows the EPC (engineering, procurement and construction) cost for the Subtask 3.2 Preliminary Design for the Eastern coal, air-blown case. These costs are on a second quarter 2004 basis. The investment is adjusted for labor rates and productivity in upstate New York.

Table 5.12 Capital Cost Summary, Preliminary Design for the Eastern Coal, Air-Blown Case
(US$)

Description	Total Project Cost*	Percent of Total
Coal Preparation and Handling	7,551,000	8.4
Air Compressor	4,321,000	4.8
Coal Feeding	1,739,000	1.9
Gasification	5,168,000	5.7
Dust Removal	2,629,000	2.9
Ash Removal	2,217,000	2.5
Gas Cooling	2,373,000	2.6
Gas Cleaning	2,812,000	3.1
Sour Water Stripper	2,979,000	3.3
Acid Gas Removal and Sulfur Recovery	10,800,000	12.0
Gas Turbine and HRSG	29,890,000	33.2
Offsites and Auxiliaries	14,583,000	16.2
Buildings	2,913,000	3.2
TOTAL	**89,976,000**	**100.0**

* All plant EPC costs mentioned in this report are second quarter 2004 cost estimates with an accuracy of +30/-15% and 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 in most cases and most of the gasifier island is assumed to be stick built. Examples where modular design and construction was applied include the sulfur clean-up and air separation units.

5.6 AVAILABILITY ANALYSIS

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 both the Subtask 3.2 IGCC co-production (export power and steam) plant designs to account for forced and scheduled outages to determine expected annual revenue and expense cash flows. Based on these cash flows, financial analyses were performed to evaluate the comparative economics of the various plant configurations and alternate design options.

The effect of sparing (back-up equipment or parallel trains of reduced capacity) can have a significant influence on the capacity factor (availability) of a plant depending upon the amount of spare equipment or parallel trains that are present. Sparing is most effective in increasing the overall plant availability when those portions of the plant with the lowest on-stream factors are replicated. Because reliability is key to the Subtask 3.2 design, sparing played an important role in the design development to provide optimum on-stream capacity while also attempting to maintain economic viability. Availability analysis for the Subtask 3.2 air-blown case resulted in annual average on-stream factors of 85.67%.

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

5.7 FINANCIAL ANALYSIS

The general methodology followed for performing the financial analysis was outlined in Section 3.6. Inputs were placed into the Nexant developed IGCC Financial Model Version 3.01 to obtain the results elaborated upon in this section and Section 6.6. Please refer to Addendum C to view the base case model inputs.

The plant EPC cost entered into the financial model was taken from the analysis done in Section 5.5, with only a few modifications. The main difference is the estimated cost for the gasification section of the plant. The cost for this plant area has been increased so when the project contingency of 15% is applied across the entire facility, the net impact is a 25% process contingency increase on the gasification section. Process feeds and products come directly from the plant configuration outlined in Section 5.2.

“Guaranteed Availability” entered into the financial model refers to plant operations excluding scheduled maintenance outages. This number only gives insight into plant availability for times where the plant is scheduled to operate. The detailed availability analysis in Section 5.6 calculates overall yearly availability, which provides the total availability regardless of if outages are scheduled or not. Therefore, the reported availability in Section 5.6 of 85.67% is the “Guaranteed Availability” of 90.9% times the percentage of time the plant is scheduled to operate (8256 hours/year, or 92.4% of the time).

5.7.1 Results

For an air-blown facility with EPC costs of 90.0 M$ and a project life of 20 years, the return on investment (ROI) is expected to be 5.9%, with a net present value (NPV) of -14.6 M$ given a 10% discount factor. Table 5.13 outlines the rate of return, NPV, payback year, and required electricity and steam selling prices to obtain a 12% ROI with other entries fixed. It is important to keep in mind that there are two major products from this facility, electricity and steam, and both values must be considered when determining the suitability of this project. Besides the base case, a "high" and "low" estimate is listed reflecting the current cost accuracy assumption of +30/-15%.

Table 5.13 Air-Blown Financial Cost Summary

	Cases		
	Base	**Low -15% EPC**	**High +30% EPC**
ROI (%)*	5.9	10.7	< 0
NPV @ 10% Discount Rate (M$)	-14.6	2.26	-52.8
Number of Years to Payback	17	14	>20
Electricity Selling Price for 12% ROI (cents/kWh)**	9.02	8.4	13.2
Steam Selling Price for 12% ROI ($/ton)***	17.56	13.8	34.3

* With an export power price of 8.0 cents/kWh and a steam price of 12 $/ton

** With a steam price of 12 $/ton

*** With an export power price of 8.0 cents/kWh

For the base case, Table 5.14 breaks down the total plant cost including EPC costs, all fees, start-up costs, and costs occurred from project financing. The "High" and "Low" case costs would be proportionately changed by the percentage difference in EPC costs.

As mentioned in Section 5.5, Subtask 3.2 represents a case focused on reliability and simplicity, positioned in an industrial site without taking any credits for already existing facilities. Future analysis will consider plant cost savings more closely. Project developers should take into account facilities and services available at the site where this gasification plant is to be located in order to reduce costs and unit requirements.

Also note that the analysis does not compare other options to IGCC, that is installation of environmental controls or natural gas combined cycle (NGCC). A prospective user of the technology must factor their estimates of alternative compliance costs to meet new emission rules, or the need to expand or replace their existing utility systems as a comparison to the options offered in this study.

With co-production of electricity and steam at this facility, comparisons must be made to similar facilities or all outputs converted to one energy form in order to determine the competitiveness of industrial gasification. If a steam turbine existed, the electrical output from this unit would be between 5 to 8 MW with no steam export. Please note that this scenario does not reflect an optimized design to maximize electricity output, but rather converts the energy potential in the export steam to electricity. Based on the plant EPC costs, this equates to an electricity cost of ~3000-3300 $/kW.

Another way to evaluate the cost is to look at the capital cost of the plant required to produce syngas for the power island. This is about 3.75 $/MBtu produced. (This value is an approximation calculated by simply removing the investment cost of the gas turbine and HRSG from the overall capital.) This method allows for a simple comparison with anticipated fuel costs to determine if gasification is a cost effective means to meet electricity and steam demands from a combined cycle system powered by premium fuels. This compares to a typical cost for utility scale IGCC facilities of 3.00-3.50 $/MBtu.

Table 5.14 Air Blown Base Case Total Plant Costs

Construction/Project Cost (in Thousand Dollars)		
Capital Costs	**Category**	**Percentage**
EPC Costs	$90,430	72%
Initial Working Capital	$1,111	1%
Owner's Contingency (% of EPC Costs)	$13,565	11%
Development Fee (% of EPC Costs)	$3,617	3%
Start-up (% of EPC Costs)	$1,809	1%
Initial Debt Reserve Fund	$0	0%
Owner's Cost (in thousand dollars)	$3,617	3%
Additional Capital Cost	$0	0%
Total Capital Costs	**$114,149**	**91%**
Financing Costs		
Interest During Construction	$8,572	7%
Financing Fee	$2,430	2%
Additional Financing Cost #1	$0	0%
Additional Financing Cost #2	$0	0%
Total Financing Costs	**$11,002**	**9%**
Total Project Cost/Uses of Funds	**$125,150**	**100%**
Sources of Funds		
Equity	$42,551	34%
Debt	$82,599	66%
Total Sources of Funds	**$125,150**	**100%**

5.7.2 Sensitivities

All financial model inputs were varied to determine the project financial sensitivities. Parameter 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 process variable 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 model inputs of highest sensitivity.

The two parameters that had the greatest impact on overall project finances were guaranteed availability and the electricity tariff level. "Tariff" refers to the sales or purchase value of a commodity. In this case, "Electricity Tariff" is used to refer to the sales value for the electricity that the plant generates. "Tariff" is more comprehensive than just referring to the sales value, since it also refers to the marginal price that the industrial client currently pays for electricity. Total operating hours and electricity escalation, because of their direct relationship to availability and electricity value, also were found to have a strong financial impact. Figure 5.6 shows the relationship that varying the guaranteed availability has on the NPV assuming a 10% discount rate. Figure 5.7 shows the relationship between availability and ROI.

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. While Gasification economics continue to show promise, project developers should consider the operating hours required for the facility to be economically justified.

Figure 5.6 Effect of Availability on Air-Blown Project NPV (10% Discount Rate)

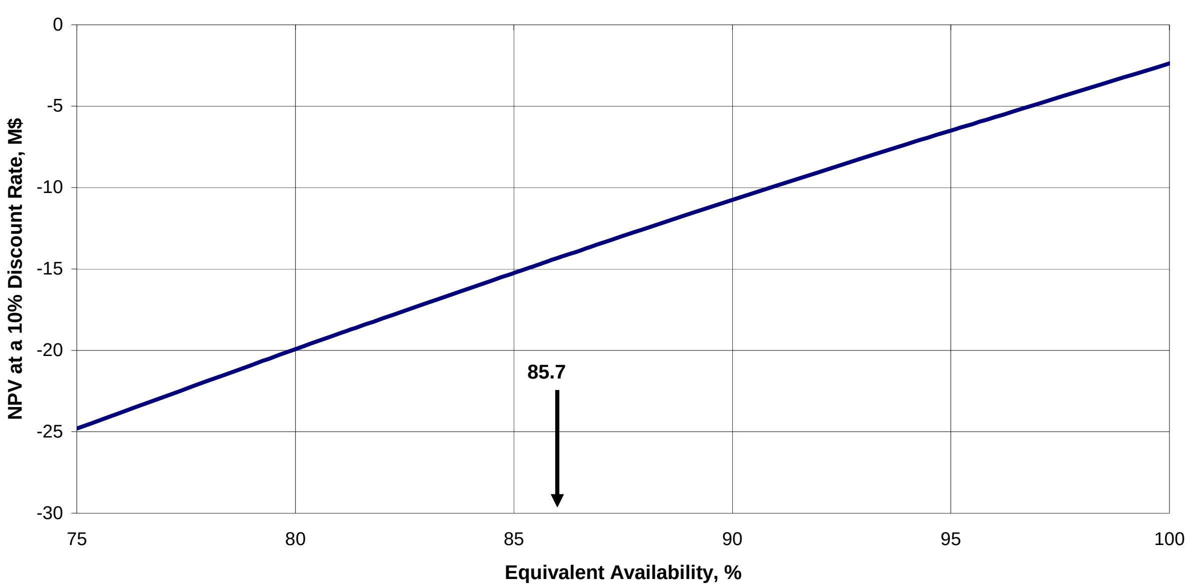


Figure 5.7 Effect of Availability Tariff on Air-Blown Project ROI

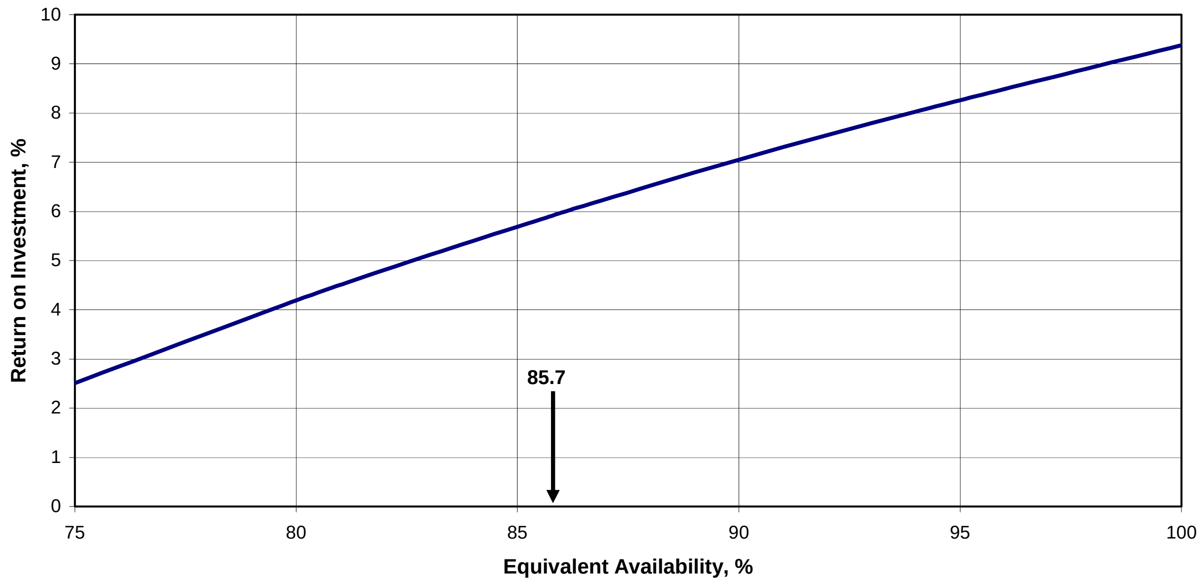


Figure 5.8 shows the relationship between the electricity tariff and NPV assuming a 10% discount rate and steam value of 12 $/ton. Figure 5.9 shows the relationship between the electricity tariff and ROI.

Figure 5.8 Effect of Electricity Tariff on Air-Blown Project NPV
(Discount Rate = 10% and Steam Value = 12 $/ton)

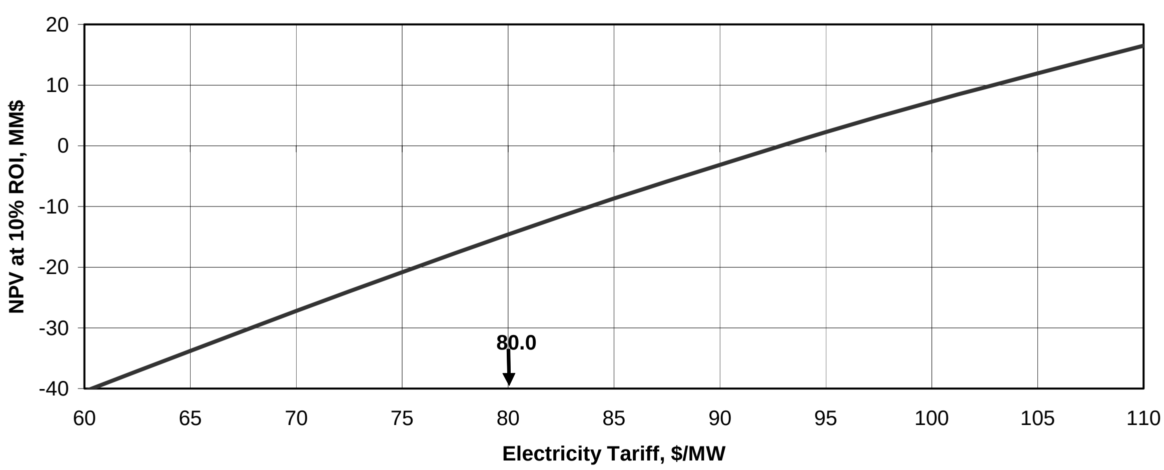
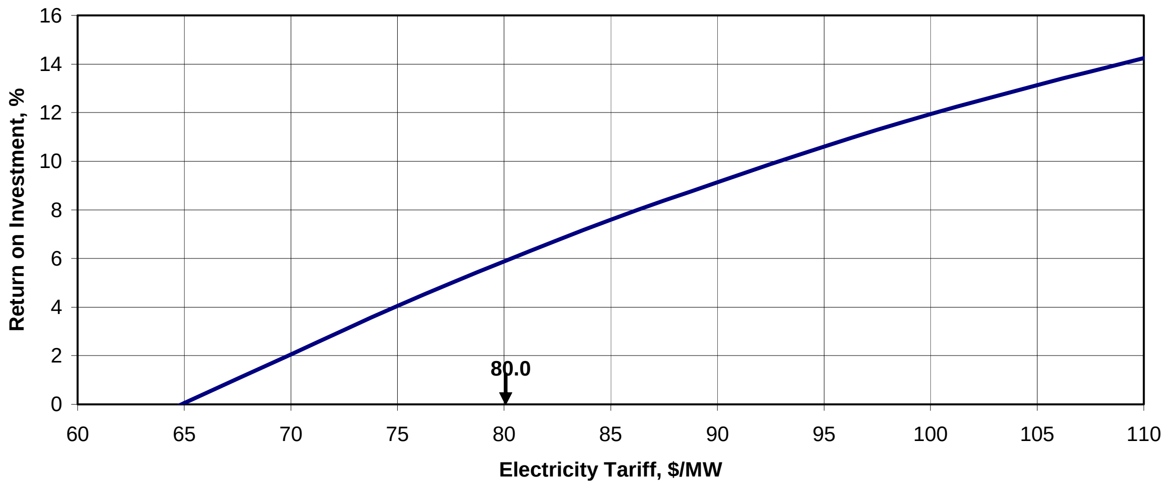


Figure 5.9 Effect of Electricity Tariff on Air-Blown Project ROI
(Steam Value = 12 $/ton)



Marginal rates for electricity have varied widely in recent years and locations due to fluctuating natural gas prices, changing market conditions, and new state and federal electricity regulations. With relatively stable prices for coal, development of an industrial Gasification unit for power and steam can act as a hedge against fluctuating prices and supply. Comparing the electricity value required to obtain a given level of NPV versus other marginal costs for electricity generation is not a complete comparison due to the export steam generated. Any analysis of other generation options versus this project must look at both electricity and steam generation requirements.

Economic life of the facility was found to be important, large changes in the total years of operation were required to vary the economics considerably. A 25% change in project life, from 20 to 25 years improved the present value by 6.6 M$. While this is significant, it is unclear what the sustained life of an IGCC facility of this nature will be due to a lack of industrial comparisons available. Decreasing the project life from 20 to 15 years had an even greater impact, decreasing the NPV to -14.8 M$.

All other process variables tested were found to have much less significance in impacting the overall plant economics. Steam value/escalation, interest rate, coal price/escalation, O&M costs, percentage of plant financing via debt, and contingency fee rate were found to have the next greatest level of impact on facility economics after the variables mentioned above. However, for the ranges tested, none of these inputs varied the NPV by more than ~8 $M or the ROI by more than 4 percent. Table 5.15 shows some of the impacts that these variables have. The base case entries are in parenthesis after the sensitivity value:

Table 5.15 Other Financial Model Input Sensitivities, Air-Blown Case

	New Value	NPV (M$, 10% discount rate)	ROI (%)
Economic Life (Years)	25 *(20)*	-8.7 *(Base = -14.6)*	8.1 *(Base =5.9)*
Steam Value ($/ton)	14 *(12)*	-9.1	7.5
Interest Rate (%)	6 *(8)*	-6.7	8.0
Coal Value ($/ton)	42*34.6)*	-20.5	4.2
Fixed O&M (% EPC)	3.5*(3.0)*	-15.7	5.5
Debt (%)	50 *(66)*	-17.3	6.2
Contingency Fee (% EPC)	10 *(15)*	-11.3	6.7

Other model inputs tested, including sulfur value, ash value, O&M escalation rates, and depreciation, among others, had even less significance than the variables listed in Table 5.15. When evaluating the economics for a facility of this size, availability and electricity tariff value should be focused on the most when considering the range of financial outcomes. Other inputs, while important to a complete picture of a facility's financial potential, will have the impact of these two factors.

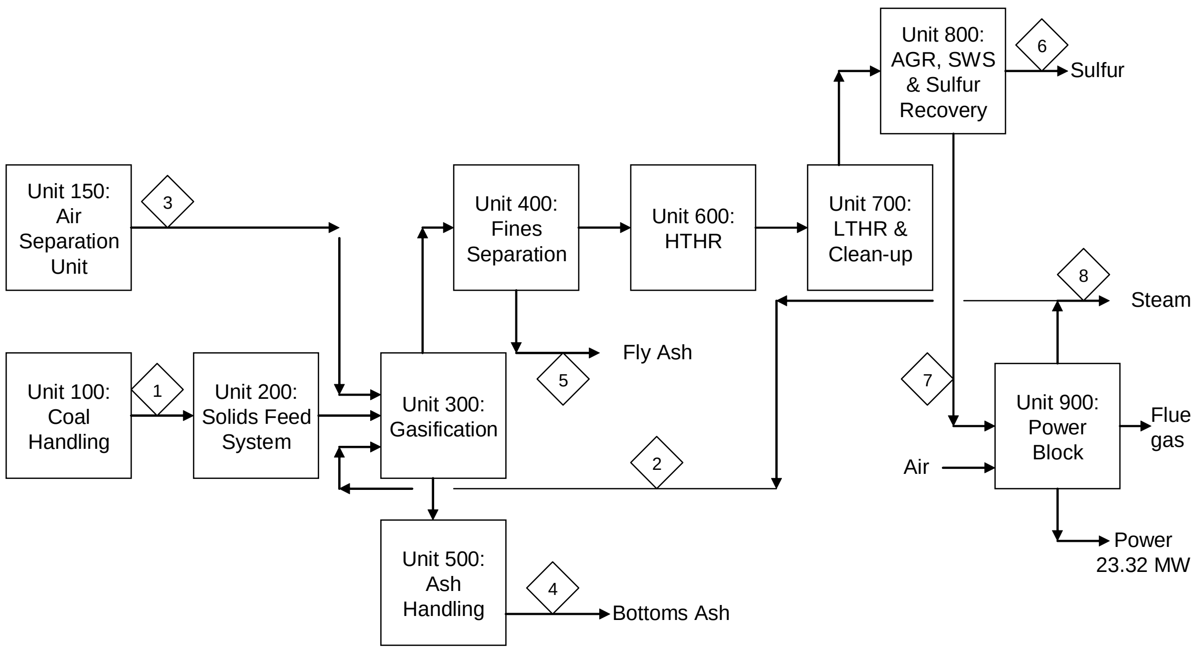
From the analysis above, it can be stated that the project finance inputs are robust and would require large changes in the model assumptions to obtain results that are substantially different. The two entries most critical to the financial analysis, availability and electricity price, can vary significantly based on plant design, facility location, and energy commodity values. Because of the range of values possible for individual facilities in these critical areas, the results of this analysis should not be applied to every facility considering industrial-sized 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 proper model entries.

6.1 INTRODUCTION

This section describes the oxygen-blown case, highlighting the differences between the cases.

The overall material balance generated using ASPEN is shown in Table 6.1. The complete material balance is shown in the Addendum D.

Figure 6.1 Overall Material Balance, Oxygen-Blown 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	28,400	30,208	19,685	872	594	863	50,881	26,800

6.2 PLANT CONFIGURATION

6.2.1 Coal Preparation/Handling (Unit 100)

The coal handling (i.e., crushing and drying) is the same configuration, flow and equipment as used for the air-blown case (see Section 5.2.1) because the coal flow rates are essentially the same.

6.2.2 Air Separation Unit (Unit 150)

A PRISM® APack™ Generator air separation unit from Air Products provides 230 tons per day of 95% pure oxygen at about 70°F and 15 psig. Two parallel oxygen compressors, each with 50% capacity, compress the oxygen to the gasifier inlet pressure, 415 psia. Two Dresser-Rand reciprocating compressors, each having a flow rate of 9,840 lb/hr of 95% oxygen, are used to provide operating flexibility during operations at reduced capacity. Each compressor is a five-stage machine with four intercoolers, and requires a 600 BHP electric motor drive.

The oxygen is discharged from the compressors at 231°F and 419 psia and is heated to 500°F with superheated 400 psig / 550 °F steam before entering the gasifier.

6.2.3 Gasification Island (Units 200-500)

The gasification island design for the oxygen-blow case is essentially the same as for the air-blown case except for:

In oxygen-blown operation the gasifier consumes 28,400 lb/hr (341 tpd) of dry coal that has a maximum of 5% surface moisture. Oxygen and steam at high pressure are mixed and fed to the gasifier to react with the coal. Design conditions of 19,685 lb/hr of oxygen and 30,208 lb/hr of steam are required for gasification. The product gas composition (mole basis) from the gasifier contains the following major components (re. Table D.6).

CO	22.65%
CO_2	15.07%
H_2	26.58%
H_2O	28.21%
CH_4	5.47%
H_2S	0.70%
COS	0.02%
NH_3	0.20%
HCN	0.02%
N_2	1.08%

Small quantities of light oils (primarily benzene), dust, chlorides, and mercury are also included in the gas stream and must be removed in the downstream cleanup system. Complete details are shown in the material balance in Table D.6 in Addendum D.

6.2.4 High Temperature Heat Recovery (Unit 600)

6.2.4.1 Characteristics of Raw Syngas

The particulates in the syngas stream leaving the gasifier comprise ash, unburned carbon, and a small amount of trace elements. Table 6.1 lists the major characteristics of the syngas exiting the gasifier.

Note that for the oxygen-blown case the mass flow rate is much smaller than that for the air-blown case. The water vapor content in the syngas is much greater than that in the air-blown case, about twice as much because steam is used to moderate the gasifier temperature.

Table 6.1 Major Characteristics of Syngas Leaving the Gasifiers

Temperature Leaving the Gasifier (ºF)	1750
Pressure (psia)	355
Mass Flow Rate (Lb/hr)	76200
Water (Lb/hr)	19313
Oils condensation Temperatures	180 ~ 450
Dew Point (ºF)	220
Ammonium Chloride Condensation Temperature (ºF)	~ 540

6.2.4.2 High Temperature Heat Recovery System

The high temperature heat recovery system design for the oxygen-blown case is similar to that for the air-blown case, consisting of a steam boiler and a steam drum. A thermosyphon loop is formed between the steam drum and the boiler. A pump is used at startup. The steam boiler is a vertical firetube exchanger with the syngas flowing downward on the tube side. BFW at 250°F flows into steam drum, where it condenses some of the saturated steam generated in the steam drum; the liquid water circulates back to the steam boiler where it flows upward on the shell side. The drum has a 30 minute residence time. The steam boiler tubes are made of inconel for better corrosion resistance. A total of 80 2-inch ID tubes are used. The average gas velocity is designed for about 30 ft/sec, and the overall heat transfer coefficient across the tubes is calculated to be about 65 Btu/hr-ft^2-ºF.

The high temperature heat recovery system produces about 40,000 lb/hr of saturated 400 psig steam, which is then routed to the HRSG for superheating. A list of equipment can be found in the Addendum B.

6.2.5 Syngas Cleanup System (Units 700 and 800)

Similar to the air-blown case, the syngas cleanup system comprises two syngas scrubber columns, one COS hydrolysis reactor, a number of heat exchangers for syngas cooling and heating, and a sulfur impregnated activated carbon bed for mercury removal. The following subsections highlight the design of these units.

6.2.5.1 Syngas Scrubber

Table 6.2 shows the water balance for the syngas scrubber columns. In this case the raw syngas contains about three times the amount water as the treated syngas, whereas in the air-blown case, the treated syngas contains about twice the amount of water as the raw syngas. In addition, the amount of water used in the scrubber is about

129,000 lb/hr, compared to 45,000 lb/hr for the air-blown case. The extra quench water is needed to cool the larger amount of water vapor contained in the raw syngas. Similar to the air-blown case, only a small portion of the water used (about 25% of the total) is make-up water.

Table 6.2 Water Balance in the Sungas Scrubber Column

	Inlets				Outlets	
	Syngas	**Recycled Sour Water**	**Fresh Quench Water**	**Process Condensate**	**Cleaned Syngas**	**Sour Water**
Temperature ºF	600	110	80	110	265	265
Flowr, lb/hr	19,313	90,000	35,762	3,153	6,521	141,707

6.2.5.2 Low Temperature Heat Recovery System

A series of three heat exchangers are used to cool the syngas, with the first one being a BFW preheater, the second an air cooler, and the last a water cooler. The heat exchanger duties for these heat exchangers are listed in Table 6.3.

Table 6.3 Duties of the Heat Exchangers in the Low Temperature Cooling Section

Heat Exchanger	Syngas Temperature (ºF) Inlet - Outlet	Duty (MBtu/hr)
BFW Preheater	275 - 227	5.56
Air Cooler	239 – 140	12.09
Water Cooler	140 - 110	2.04

6.2.5.3 Mercury Removal

Introduction

As in the air-blown case, the mercury present in the coal will partition primarily to the syngas stream. The same design coal used in the air-blown case is used in the oxygen-blown case, with an approximate mercury concentration of 0.12 ppmw. Because the oxygen-blown case uses slightly less coal feed and has a smaller volume flow of syngas, the concentration and mass flow rate of mercury in the syngas is different than that of the air-blown case. For the oxygen-blown case, the mercury concentration in the syngas is on the order of 60 µg/Nm3 and the mass flow rate is approximately 0.0034 Lb/hr. As in the air-blown case, mercury control is designed to achieve greater than 90% removal.

Basis

The basis for the oxygen-blown case is the same as the air-blown case, that of the Eastman Chemical Company's Chemicals-from-Coal facility.

Process Description

The mercury control equipment for the oxygen case is the same as that of the air-blown case, a vendor equipment design supplied by Calgon Carbon Corporation (www.calgoncarbon.com). In respect to mercury content, the process streams of the air-blown case and the oxygen-blown case differ slightly in terms of syngas concentration, but are almost equivalent in terms of mass flow. Despite the approximately 50% increase in syngas mercury concentration (40 µg/Nm3 for the air-blown case compared to 60 µg/Nm3 for the oxygen-blown case), the syngas concentration is low enough and the mass flow rate is essentially equal such that the design would be the same for both the air-blown and oxygen-blown cases. The equipment consists of a single adsorber vessel, 9 feet in diameter and 10 feet on the straight side. The vessel is packed with 20,000 lbs of Calgon Carbon HGR® sulfur impregnated activated carbon. The expected bed life is approximately 3-5 years.

Special Considerations

As in the air-blown case, mercury control using sulfur impregnated activated carbon is highly temperature dependent and requires a process temperature of near 100°F.

Results/Conclusions

The mercury control equipment is based on a commercially proven, reliable design. The equipment is expected to meet or exceed the design removal of 90% of the mercury in the syngas.

6.2.5.4 Acid Gas Removal and Clean-up

The amount of H_2S contained in the syngas for the oxygen-blown case is 859 lb/hr compared to 918 lb/hr for the air-blown case or about 6 percent less. The design of the acid gas removal and clean up is essentially the same as for the air-blown case described in Section 5.2.4.

6.2.6 Power Block - Gas Turbines and HSRG (Unit 900)

6.2.6.1 Introduction

As in the air-blown case, the primary products of this application are power and steam. Power is generated by two General Electric (GE) combustion turbine/generator sets (CT) (11.25 MW ISO conditions, natural gas DLE). Steam is generated using a two-pressure heat recovery steam generator (HRSG). The power block consists of two parallel CT/HRSG trains.

6.2.6.2 Basis

The basis for the oxygen-blown case is the same as the air-blown case. The power block is designed around the two CTs. The exhaust gas exiting the CT is routed through the associated two-pressure HRSGs. The individual HRSGs are designed such that three specific process conditions are met:

- Stack temperature remains above the acid-dew point to avoid condensation and corrosion within the system.
- 50 psig superheated steam (~350°F) is generated such that the process steam demands of all gasifier and gas clean-up processes (including gas clean up operations and sour water treatment) are satisfied.
- Balance of steam generation is 400 psig superheated steam (~550°F).

6.2.6.3 Process Description

Clean syngas is sent to the CT at 120°F and 295 psia. For the oxygen-blown case, the syngas is diluted with steam for NOx control. Figure D.5 in Addendum D illustrates the CT/HRSG.

Modeling for this study estimated that each CT would generate approximately 14.86 MW of net power, slightly less than the air-blown case but greater than the 11.25 MW ISO rating for natural gas. As in the air-blown case, the syngas boost effect 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 (about 890°F, 15.6 psia) 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 58% of the water flow (~34,200 lb/hr) entering the 50 psig evaporator is extracted as liquid and sent to the 400 psig evaporator. The remaining 42% of the water (~24,800 lb/hr) exits the evaporator as 50 psig saturated steam. This flow then goes to the 50 psig superheater, where it is heated to approximately 350°F.

The liquid water exiting the 50 psig evaporator flows to the 400 psig evaporator. Approximately 38,000 lb/hr of 400 psig saturated steam are generated in the evaporator. Approximately 4% of the inlet water flow (~1,300 lb/hr) is blowdown from the system. The saturated steam exiting the evaporator is mixed with the 400 psig saturated steam coming from the gasifier waste heat boiler. The mixed saturated steam then goes to the 400 psig superheater, producing approximately 52,500 lb/hr of 400 psig superheated steam (~550°F).

6.2.6.4 Special Considerations

The same issues outlined in the air-blown case are applicable to the oxygen-blown case.

6.2.6.5 Results/Conclusions

For this application, two parallel CT/HRSG trains were designed based on turbine availability and system reliability. The CT's are General Electric GE 10s and the HRSGs are designed to provide self sufficient quantities of 50 psig superheated steam. The balance of steam production by the HRSG is 400 psig superheated steam exported to the industrial facility for process consumption. Table 6.5 provides summary of the net output of the power block. Table 6.4 does not include power and steam demands of the gasifier operations.

Table 6.4 Power and Steam Net Output (Power Block Only)

	Single Train (1 of 2)	Combined Trains (2 of 2)
Power Output	14.85 MW	29.7 MW
50 psig Superheated Steam Generation	24,842 lb/hr	49,684 lb/hr
400 psig Superheated Steam Generation	52,483 lb/hr	104,966 lb/hr

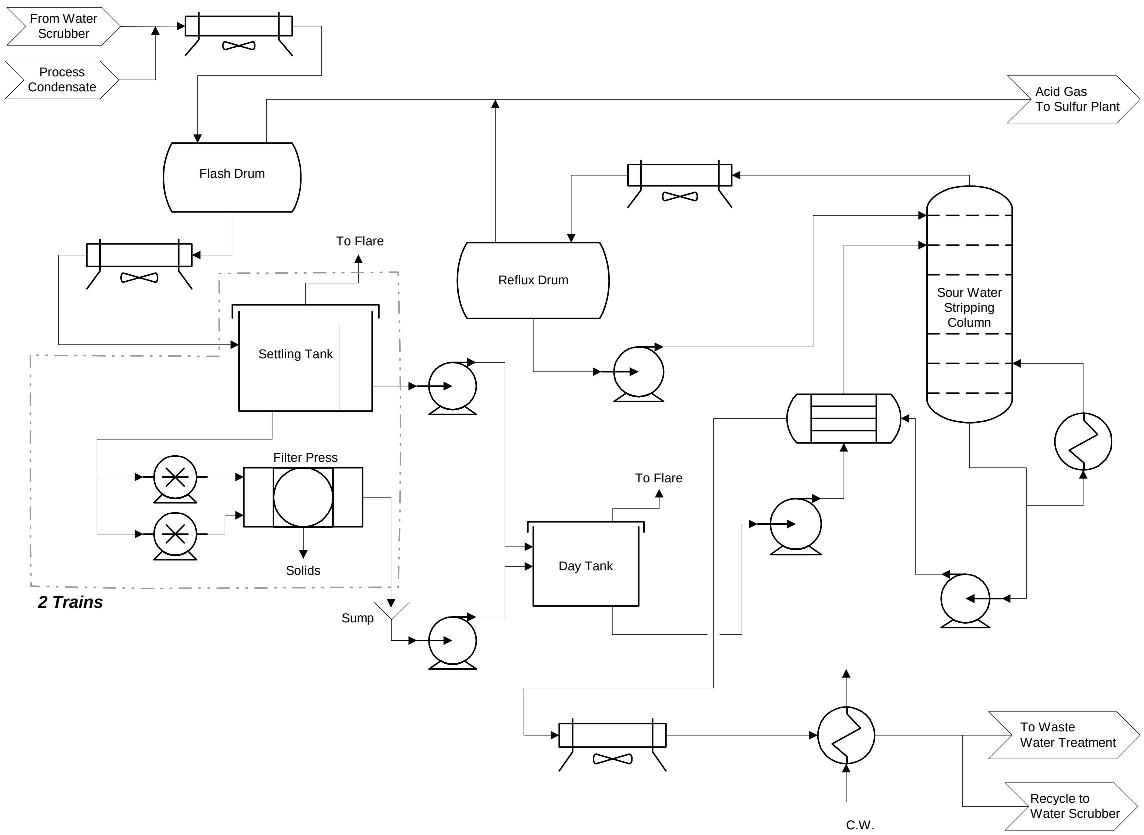
Compared to the air-blown case, the oxygen-blown case produces slightly less power; more 50 psig superheated steam; and less 400 psig superheated steam. The decreased power output is a result of the slightly lower mass flow rate of syngas through the combustion turbine. The 50 psig superheated steam production is increased to meet a greater internal demand, a result of the significantly higher amount of sour water treatment required for the oxygen-blown case compared to the air-blown case. The decrease in 400 psig steam production is a result of a HRSG design to produce additional 50 psig steam and b) the gasifier block of the oxygen-blown case provides approximately 20% less saturated steam to the HRSG (~39,000 lb/hr) compared to the air-blown case (~50,000 lb/hr).

6.2.7 Sour Water Stripper (Unit 800)

6.2.7.1 Introduction

The sour water treatment unit for the oxygen-blown case is fundamentally similar to the air-blown case. A few process differences exist because the oxygen-blown case treats a significantly greater volume of sour water, 145,000 lbs per hour (~784 gallons/minute) for the oxygen case compared to 45,000 lbs per hour (~90 gallons/minute) for the air case. Figure 6.2 is a schematic flow diagram of the sour water treatment unit for the oxygen-blown case.

Fig 6.2 Sour Water Treatment Unit Schematic (Oxygen-Blown Case)



6.2.7.2 Basis

The design basis of the sour water stripper for the oxygen-blown case is the same as that of the air-blown case. The stripping column produces a liquid effluent stream containing no more than 50 ppmw ammonia and less than 10 ppmw hydrogen sulfide.[1] The resulting stream is cooled to 110°F before either being recycled to the syngas water scrubber or being blowdown to the wastewater treatment plant.

6.2.7.3 Process Description

Sour water from the water scrubber is mixed with a portion of the process condensate, cooled to 171°F, and sent to the flash drum where it is flashed to near atmospheric pressure. This cooling minimizes the water in the vapor product that leaves the flash drum. The vapor stream leaving the flash drum consists primarily of CO_2, water and H_2S. The liquid stream is then cooled to 161°F (approximately 10°F below the liquid saturation point at atmospheric pressure) to meet the process requirements of the settling tanks. As in the air-blown case, the additional cooling reduces the chance for any off-gassing in the settling tanks. The remaining process units are identical to the air-blown case, the settling tanks, filter presses, day tank storage, preheater, stripper

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

column, bottoms product cooling. Other than size, there are no significant differences in the settling tanks, filter presses, or day tank for the oxygen-blown case compared to the air-blown case. The specific design information and simulation results are included in Addendum B.

6.2.7.4 Special Considerations

The same special considerations (corrosiveness, water vapor, etc.) of the air-blown case also apply to the oxygen-blown case. For the same reason described in the air-blown case, attempts to minimize the water content of the vapor stream sent to the sulfur plant were considered during the design. The same constraint that the water content of the vapor stream going to the sulfur plant contains less than 5 percent (approximately 200 pounds per hour) of the total feed stream to the sulfur treatment plant also was applied to the oxygen-blown case.

As with the air-blown case, the presence and fate of any dissolved organics (benzene and toluene derivatives, including phenols) in the sour water streams also must be verified to provide a clear understanding of the wastewater treatment needs.

6.2.7.5 Results/Conclusions

For the oxygen-blown case the sour water feed rate was slightly less than 145,000 lbs per hour (~290 gallons/minute). The resulting stripper column consists of 21 stages (including the condenser), with tray spacing of 2 feet. The column diameter measures 4.75 feet. It is constructed of stainless steel clad carbon steel with stainless steel internals. The sour water flash drum, column distillate drum, and overhead condenser are constructed of carbon steel with stainless steel cladding. Stainless steel also was selected for the tube side of the stripper feed preheater, but the shell is carbon steel. The sour water cooler is an air finned design, with the surfaces that contact the sour water being constructed of stainless steel. The product recycle water has a design ammonia concentration of less than 50 ppmw, and the H_2S and COS concentrations each are 1 ppmw or less.

6.2.8 Offsites/Utilities

Offsites and utilities are essentially the same as for the air-blown case. The oxygen-blown case has higher cooling water consumption, slightly lower nitrogen demand, lower electrical power consumption, lower steam production and higher 400 psig steam consumption as shown in Table 6.5.

Table 6.5 Air- and Oxygen- Blown Utility Requirements

	Air-Blown Case	Oxygen-Blown Case
Electrical Power, kW		
Gross Production	29,830	29,694
Consumption	8,130	6,371
NET	**21,700**	**23,323**
Cooling Water requirements, gpm	2,810	4,054
Nitrogen consumption, SCFH	14,164	13,246
Steam – 400 psig, thousand lb/hr		
Production	130.7	110.0
Consumption	29.0	83.2*
NET	**101.7**	**26.8***
Steam – 50 psig, thousand lb/hr		
Production	38.3	49.7
Consumption	38.3	49.7
NET	**0.0**	**0.0**

* Design maximum is 94.1 thousand lb/hr and corresponding export is 15.8 thousand lb./hr

6.3 EMISSIONS

Refer to Section 5.3 for a discussion of emissions.

6.4 TRADE-OFF STUDIES

Many of the trade-off studies that were examined as part of the air-blown case are applicable to the oxygen-blown case and are discussed in Section 5.4. The trade-off studies applicable to the oxygen-blown case are summarized below..

- Use water from the bottom of the scrubber either before or after stripping to preheat the oxygen, nitrogen and/or fuel gas going to the turbine. It is not recommended to adopt this alternative design for the industrial site since a) for the oxygen or nitrogen the temperature driving force is small and b) for the fuel gas it results in an additional plant availability penalty.

- Use steam instead of nitrogen as the diluents for NOx control. This idea was accepted based on a payout for the nitrogen compressor of almost 7 years (i.e., steam at a cost of $6 per thousand pounds is more economic).

Numerous trade-off studies that were rejected for Subtask 3.2 will need to be revisited during the optimization portion of Subtask 3.3 and 3.4.

6.5 PLANT COSTS

For the oxygen case the approach was the same as for the air-blown case except that several cost estimates were prorated from the air case. Specifically, the gas turbine, HRSG, and acid gas removal and sulfur recovery units were estimated in this manner.

Table 6.6 shows the EPC cost for the Subtask 3.2 Preliminary Design for the Eastern coal, oxygen-blown case. These costs are on a second quarter 2004 basis for upstate New York.

Table 6.6 Capital Cost Summary, Preliminary Design for the Eastern Coal, Oxygen-Blown Case
(US$)

Description	Total Project Cost*	Percent of Total
Coal Preparation and Handling	7,551,000	7.6
Air Separation	14,909,000	14.9
Coal Feeding	1,686,000	1.7
Gasification	4,412,000	4.4
Dust Removal	2,478,000	2.5
Ash Removal	2,039,000	2.0
Gas Cooling	1,830,000	1.8
Gas Cleaning	2,579,000	2.6
Sour Water Stripper	4,110,000	4.1
Acid Gas Removal and Sulfur Recovery	10,800,000	10.8
Gas Turbine and HRSG	29,890,000	30.0
Offsites and Auxiliaries	14,614,000	14.6
Buildings	2,913,000	2.9
TOTAL	**99,811,000**	**100.0**

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

6.6 AVAILABILITY ANALYSIS

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 both the Subtask 3.2 IGCC co-production (export power and steam) plant designs to account for forced and scheduled outages to determine expected annual revenue and expense cash flows. Based on these cash flows, financial analyses were performed to evaluate the comparative economics of the various plant configurations and alternate design options.

The effect of sparing (back-up equipment or parallel trains of reduced capacity) can have a significant affect on the capacity factor (availability) of a plant depending upon the amount of spare equipment or parallel trains that are present. Sparing is most effective in increasing the overall plant availability when those portions of the plant with the lowest on-stream factors are replicated. Because reliability is key to the Subtask 3.2 design, sparing played an important role in the design development to provide optimum on-stream capacity while also attempting to maintain economic viability. Availability analysis for the Subtask 3.2 oxygen-blown case result in annual average on-stream factors of 82.65%. The availability of the air separation unit is the reason for the lower availability of the oxygen-blown case.

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

6.7 FINANCIAL ANALYSIS

The approach followed in Section 5.6 for the air-blown case was replicated for the oxygen-blown case. Changes were made in the Plant Inputs section of the financial model to reflect the differences between the air-blown and oxygen-blown case. While many of the unit costs decreased slightly due to the lower quantity of syngas that had to be processed in the oxygen-blown case, the EPC cost of the oxygen production facility is 9.9 M$ greater than the air-blown case due predominantly to the cost of the air separation unit. Since contingency fees, development costs, owner's costs, and O&M costs are all based off a percentage of the total EPC value, these costs increased proportionately. Feed requirements and product output rates also changed. The greatest impact is seen in the decrease in steam export due to the parasitic requirements of the oxygen production facility and the steam injection for NOx control.

The scenario inputs for the variable financial entries into the model are unchanged from the air-blown case. The increase in the plant EPC costs is not expected to change the conditions of project financing or other financial assumptions. In addition, the oxygen plant should not increase the construction time or change the plant's economic life. Commodity tariffs and escalation rates were kept the same as for the air-blown case. See Addendum C for the financial model entries for the oxygen-blown case.

As with the air-blown case, the difference between the "Guaranteed Availability" entered into the financial model and the availability analysis performed in section 6.6 is that the latter takes into account time for scheduled outages. Thus, the reported availability in Section 6.6 of 82.65% is the "Guaranteed Availability" of 87.7% times the percentage of time the plant is scheduled to operate (8256 hours/year, or 92.4% of the time).

6.7.1 Results

For an oxygen-blown facility with EPC costs of 99.8 M$ and a project life of 20 years, the return on investment (ROI) is expected to be less than zero, with a net present value (NPV) of -48.6 M$ given a 10% discount factor. Table 6.7 outlines the rate of return, NPV, payback year, and required electricity and steam selling prices to obtain a 12%

ROI with other entries fixed. "High" and "low" estimates are listed as well to reflect the current cost accuracy assumption of +30/-15%.

Table 6.7 Oxygen--Blown Financial Cost Summary

	Cases			
	Base	**Low -15% EPC**	**High +30% EPC**	**Air-Blown Base Case**
ROI (%)*	<0	1.5	<0	5.9
NPV @10% Discount Rate, (M$)	-48.6	-26.9	-70.3	-14.6
Payback Year	>20	>20	>20	17
Electricity Selling Price for 12% ROI (¢/kWh)**	11.8	10.8	14.2	9.02
Steam Selling Price for 12% ROI ($/ton)***	>40	61	>100	17.56

* With an export power price of 8.0 cents/kWh and a steam price of 12 $/ton

** With a steam price of 12 $/ton

*** With an export power price of 8.0 cents/kWh

Comparing the results for the oxygen-blown case to the air-blown case, it is clear that the air-blown case is superior on a financial basis. The oxygen-blown case has higher net EPC costs, while producing significantly less steam for export. The advantage gained by having an oxygen plant, smaller process equipment sizes and slightly more net electricity export, does not outweigh the oxygen plant cost and large steam requirements. A side-by-side comparison of the air-blown and oxygen-blown cases, as can be seen by comparing the oxygen-blown base case to the last column in Table 6.8, shows how the air-blown case is superior in all financial categories.

For the base case, Table 6.8 breaks down the total plant cost including EPC costs, all fees, start-up costs, and costs occurred from project financing. The "High" and "Low" case costs would be proportionately changed by the percentage difference in EPC costs.

Table 6.8 Oxygen-Blown Base Case Total Plant Costs

Construction/Project Cost (in Thousand Dollars)		
Capital Costs	**Category**	**Percentage**
EPC Costs	$100,194	72%
Initial Working Capital	$962	1%
Owner's Contingency (% of EPC Costs)	$15,029	11%
Development Fee (% of EPC Costs)	$4,008	3%
Start-up (% of EPC Costs)	$2,004	1%
Initial Debt Reserve Fund	$0	0%
Owner's Cost (in thousand dollars)	$4,008	3%
Additional Capital Cost	$0	0%
Total Capital Costs	**$126,205**	**91%**
Financing Costs		
Interest During Construction	$9,477	7%
Financing Fee	$2,686	2%
Total Financing Costs	**$12,163**	**9%**
Total Project Cost/Uses of Funds	**$138,367**	**100%**
Sources of Funds		
Equity	$47,045	34%
Debt	$91,323	66%
Total Sources of Funds	**$138,367**	**100%**

As mentioned in the air-blown case, this analysis represents the economics for a design focused on simplicity and maximum availability. Future analysis will look more closely at economic optimization that may be able to decrease the overall cost of the facility. Existing facilities at the site where this unit is located also may decrease the overall cost.

If a steam turbine existed, the electrical output from this unit would be between 2.5 and 5 MW. Based on the plant EPC costs, this equates to an investment cost of 3450-3800 $/kW. The incremental gas cost for the oxygen-blown plant is about 4.44 $/MBtu. (This value is an approximation calculated by simply removing the investment cost of the gas turbine and HRSG from the overall capital.)

6.7.2 Sensitivities

Because there has been no change in the financial assumptions made between the air-blown and oxygen-blown cases, the parameters found to be most sensitive in the oxygen-blown case are the same as in the air-blown. Guaranteed availability and electricity tariff rate were again found to be the most sensitive model inputs. Figures 6.3

and 6.4 show the impact that changes in the availability has on the project NPV and ROI at a 10% discount rate. Figures 6.5 and 6.6 show the impact that changes in the electricity tariff has on the project NPV and ROI at a 10% discount rate. All other model inputs were held constant during this analysis (steam value = 12 $/ton).

Figure 6.3 Effect of Availability on Oxygen-Blown Case NPV
(Discount Rate = 10%)

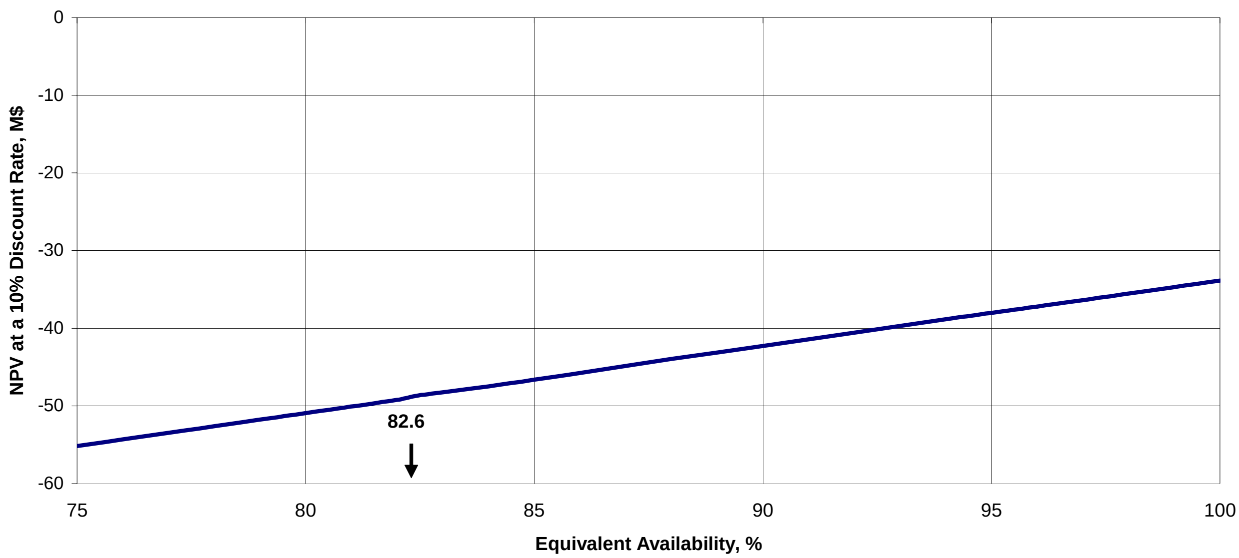


Figure 6.4 Effect of Availability on Oxygen-Blown Case ROI

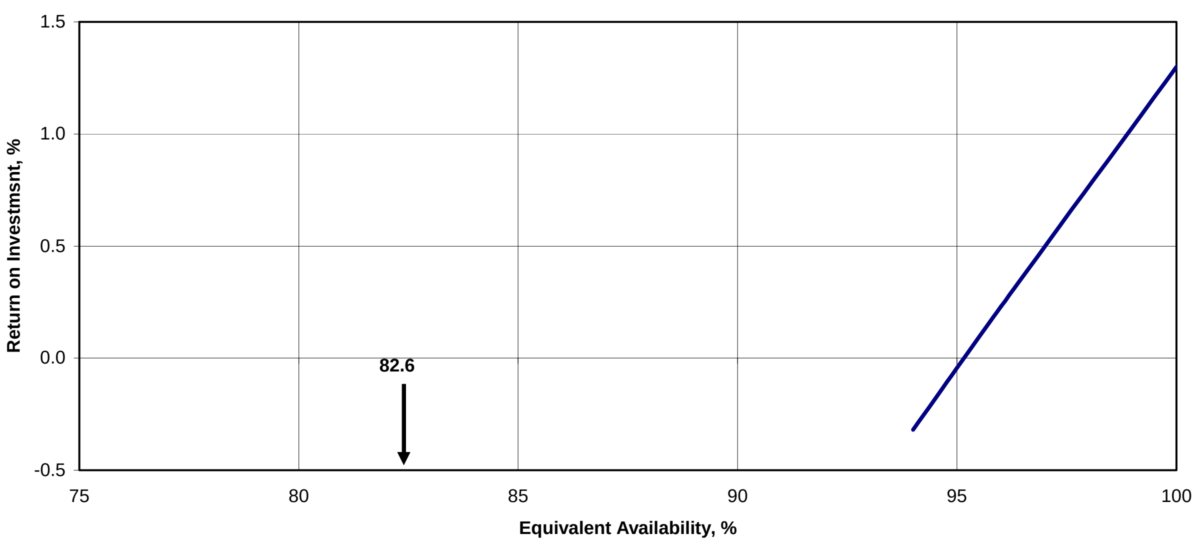


Figure 6.5 Effect of Electricity Tariff on Oxygen-Blown Case NPV
(Discount Rate = 10% and Steam Value = 12 $/ton)

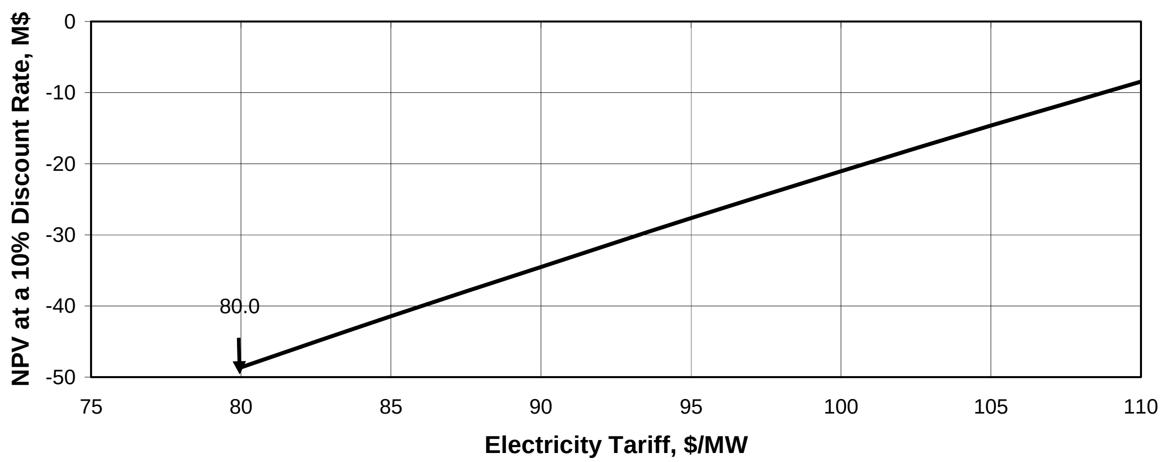
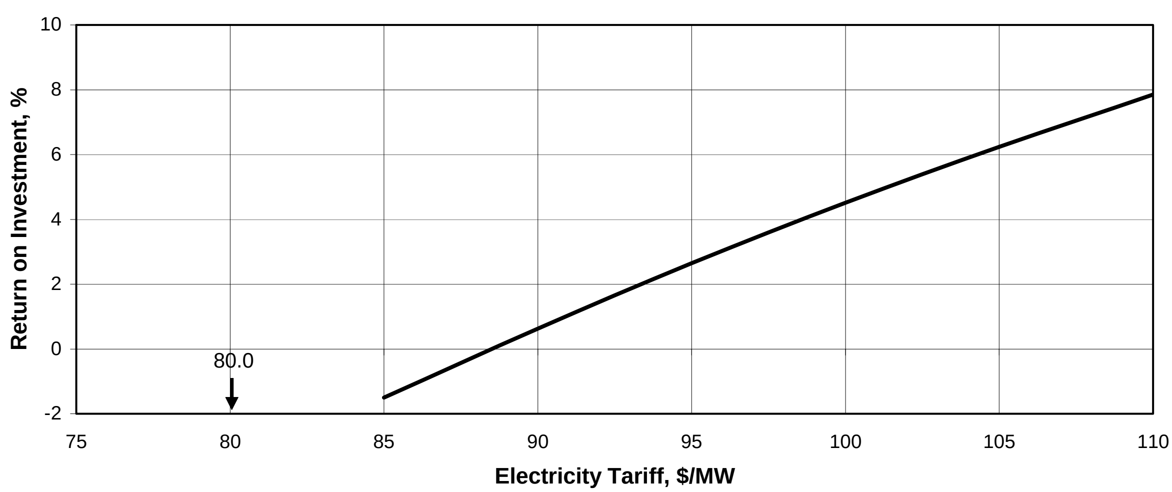


Figure 6.6 Effect of Electricity Tariff on Oxygen-Blown Case ROI
(Steam Value = 12 $/ton)



The trends seen in these two figures are similar to those witnessed for the air-blown case. High levels of availability greatly assist in assuring the plant will be economically justified. In facilities where co-production of electricity and steam are required, gasification equipment can be competitive, yet plant size should be taken into consideration for optimal gasification design. Plant size will have a large impact on whether the facility should be air or oxygen-blown. The NPV numbers seen here are lower when compared to the air-blown case for a similar range.

As with the air-blown case, all other process variable tested were found to have much less significance in impacting the overall plant economics. Economic life, steam value/escalation, interest rate, coal price/escalation, O&M costs, percentage of plant financing via debt, and contingency fee rate, were found to have the next greatest level of impact on facility economics after availability and the electricity tariff. Steam value/escalation has a much smaller impact than in the air-blown case due to the smaller quantity of export steam. With the exception of the interest rate, none of these inputs varied the NPV by more than 10 M$ for the ranges tested. The higher plant EPC cost increases the sensitivity of the model to interest rate and fees that are a percentage of the EPC cost. Table 6.9 shows some of the impacts that these variables have. The base case entries are in parenthesis after the sensitivity value:

Table 6.9 Other Financial Model Input Sensitivities, Oxygen-Blown Case

	New Value	**NPV (M$, at 10% discount rate)**
Economic Life (Years)	25 *(20)*	-44.5 *(Base Case = -48.6)*
Steam Value ($/ton)	14 *(12)*	-47.0
Interest Rate (%)	6 *(8)*	-38.5
Coal Value ($/ton)	42 *(34.6)*	-54.5
Fixed O&M (% EPC)	3.5 *(3.0)*	-50.0
% Debt	50 *(66)*	-50.4
Contingency Fee (% EPC)	10 *(15)*	-44.5

Other plant inputs tested for sensitivity had a less significant impact than the items listed in Table 6.9. Availability and the electricity tariff rate remain the items that should have the most focus when determining the range of financial outputs. The higher EPC cost in the oxygen-blown case creates a greater range of financial outputs for similar model changes when compared to the air-blown case. This slightly decreases the robustness of the financial model for considering this specific case. While the model is still robust for the situation presented, industrial facilities considering gasification for electricity and steam production must have a financial analysis performed that reflects their unique situation.

The term "robustness" for the financial model refers to the overall sensitivity to financial model to modifications in the inputs. Considering that only two inputs, electricity tariff and availability, have significant impact on model outputs, and that even those inputs have to be changed significantly to strongly impact the outputs, the model is considered robust. This sentence refers to the fact that the oxygen case ("this case") is more sensitive to overall changes in the model than the air case due to the higher EPC costs.

7.1 SUMMARY

The air-blown and oxygen-blown conceptual designs were developed for a GTI U-GAS® fluidized bed gasification facility. Table 7.1 summarizes these two designs.

Table 7.1 Overall Plant Summary

	Air-Blown Case	**Oxygen-Blown Case**
Design Inputs		
Coal Feed, moisture-free tpd	345.7	323.8
Coal Feed, moisture-free lb/hr	28,810	26,980
Fuel (Natural Gas), MBtu/hr	5.1	7.3
Makeup Water Input from the Industrial Facility		
Boiler Feed Water, gpm	495	473
Quench Water, gpm	30	70
Cooling Tower Makeup Water, gpm	53	72
Design Outputs		
Export Power, MW	21.7	23.3
Export Steam (400 psig, 550°F), Mlb/hr	101.72	26.75
Sulfur, lb/hr	899	863
Ash, lb/hr	2,097	1,465
Condensate (to industrial facility), Mlb/hr	60.9	65.5
EPC Cost, M$*	90.0	100.2
Plant EPC Cost, $/kW**	3,090	4,057
Plant Energy Input, k$/MBtu/hr	229.9	263.6
Plant Energy Output, k$/MBtu/hr	469.2	907.3
Equivalent Availability, %	85.7	82.6
Return on Investment, %***	5.9	<0
Cold Gas Efficiency, % (HHV basis)	79.3	83.1
Net CHP Efficacy, % (HHV basis)	49.0	29.1

* 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.

The air-blown design exports about 1.6 MW less power. However, it is less costly because it eliminates the costly air separation unit (ASU or oxygen plant) even though it requires larger processing equipment to handle the inert nitrogen that is contained in the syngas. Although the oxygen-blown case has smaller processing equipment, it exports less steam than the air-blown case because it consumes more steam to control the gasifier temperature, and consequently, requires more water (both boiler feed water and quench water). In the oxygen-blown case, steam is also used to control the NOx generation (while in the air-blown case the nitrogen contained in the air is the diluent). Since the costs of the processing factors tend to compensate, the investment cost difference between the two designs is about that of the cost of the air separation unit.

The return on investment (ROI) for the air-blown case is higher than for the oxygen-blown case due mostly to the lower investment and higher steam export. The resulting ROI's are 5.9% for the air-blown case and less than zero for the oxygen-blown case. The NPV at a 10% discount rate is –14.6 million dollars for the air-blown case and –48.6 million dollars for the oxygen-blown case. Plant net CHP efficacy is 49% for the air-blown case and 29% for the oxygen-blown case.

The two parameters that have the greatest impact on overall project finances were capital investment, guaranteed availability and the electricity tariff level. All other process variables tested (steam value/escalation, interest rate, coal price/escalation, O&M costs, percentage of plant financing via debt, and contingency fee rate) were found to have much less significance in impacting the overall plant economics.

7.2 CONCLUSIONS

This study has shown that:

- A ROI of 5.9% 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). The cost elements developed by this study should be useful as building blocks for developing reasonable cost estimates for plants of varying size within the 5-100 MW size range defined by this study.

- 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.

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

- As a result of this study, a list of potential enhancements has been identified that should provide additional cost savings as some of the improvements are researched, developed and implemented, such as:
    - Economy of scale (i.e., single train gasifier island)
    - The Stamet "solids" feeding 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 to reduce water consumption and capital investment
- Improved heat integration
- Simplified sour water stripper
- Improved sulfur removal methods including warm sulfur removal (e.g., LO-CAT® system)
- Warm mercury removal systems
- Improved particulate removal systems

Subtask 3.3 employed a number of these improvements in the alternate design for the Air-Blown Eastern Coal Case.

- 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 1000°F)
  - Determining the combustion turbine performance capabilities for the desired engine(s) (both output and emissions)
  - Further study of the ash characteristics associated with the char
- Based on the simulations prepared for this study the design should meet emission targets established by the DOE in their roadmap for 2010 (re. Section 5.3).

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

7.3 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, increasing the overall efficiency.

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

- Although it is known that reducing the moisture content of the coal feed going to the gasifier is more efficient than evaporating the moisture in the gasifier, it has not been established that 5% is the optimum moisture content of the gasifier feed. 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).

A.1 ASPEN

A.1.1 Gasification Island

A.1.1.1 Basis

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

A.1.1.2 Introduction

GTI's gasifier, U-GAS®, is a fluidized bed gasifier that can operate over a wide range of temperatures from 1650°F to 2000°F for bituminous coal. The gasifier also can operate over a wide range of pressures ranging from 1 to 70 atmospheres. The pressure selected is sufficient to provide the syngas to the gas turbine without compression. Higher pressures tend to favor methane formation. The gasifier has two cyclone separators that recycle the unburned carbon particles and flyash back to the gasifier. The oxidant, steam and fuel enter the gasifier from the bottom. Aspen Plus® version 11.1.1 was used to model both the GTI gasifier and the coal drying unit.

The gasifier temperature was set at 1850°F and the outlet pressure was set at 354.7 psia. Eight individual reactions were specified and the chemical equilibrium of each reaction was restricted by varying the temperature approach of each individual reaction. This technique is reasonable since the gasifier is not completely homogenous.

The models for the air-blown and oxygen-blown cases required separate sets of parameters to match the gasifier and the unburned coal (char) compositions. Different temperature approaches (for the restricted chemical equilibrium) were needed to model the two different cases. In addition to the different temperature approaches, the split fractions used for the unburned coal (char) were different for each case.

A.1.1.3 Setup

Coal Prep – Drying

The component attributes for the coal were taken from a Southeastern Ohio coal sample analyzed by AEP. The "as determined" proximate analysis for the coal sample was used for the PROXANAL attribute in ASPEN. The "dry basis" ultimate analysis for the coal sample was used for the ULTANAL attribute. These coal properties are shown is Table A.1. Trace components in the coal such as Cl, Hg, etc., were neglected because this information was not available. Therefore, no comparison could be made.

Table A.1 Typical SE Ohio Coal Properties

Ultimate Analysis	**Dry Basis**
Carbon	74.65
Hydrogen	5.79
Nitrogen	1.54
Sulfur	3.32
Oxygen	8.79
Ash	5.91
Total	100

Proximate Analysis	**As Determined**
Total Moisture	3.04
Volatile Matter	41.92
Fixed Carbon	49.31
Ash	5.73
Total	100

An R-STOIC reactor block was used to dry the coal. In the R-STOIC block, a portion of the coal reacts to form water. A FORTRAN calculator block was used to calculate the amount of water that had to be removed so that the dried coal would have 5% moisture content. The model also is setup to incorporate nitrogen if used in the drying process. A FLASH2 block is used to separate the dried coal from the moisture and the nitrogen. A schematic diagram of the ASPEN simulation is given in Figure A.1.

The Gasifier

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

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

- SEP – This block, named UNREACT, selectively separates specific amounts of each component into various streams. The block is used to bypass the unburned coal around the RGIBBS reactor block.

- RGIBBS – This block, named GASFIRXR, models the gasification of the coal. The RGIBBS block models chemical equilibrium by minimizing the Gibbs free energy.

- SSPLIT – This block, named ASHSEP, splits the bottom ash and flyash from the syngas.

- MIX – This block, named BOTMIX, mixes the unburned carbon and sulfur that bypassed the RGIBBS block with the bottom ash.

Split Fractions - Modeling the unburned coal and ash

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

Tuning the Gasifier

After the input data was entered and the split fractions for the unburned coal were calculated, the model was executed to produce a baseline. Tables A.2a and A.2b list the reactions and the temperature approaches from the gasifier temperature for the two models. Reaction one, the formation of methane, was the first reaction that was adjusted. The temperature approach for that reaction was varied until the difference in methane composition was less than 1% from the GTI stream. Next, the temperature approaches for reactions 5 and 7 were adjusted until the H_2S and COS components were less than 1%. Reaction 6 was adjusted so that the NH_3 was less than 1% and then reaction 8 was adjusted so that the HCN was within 2%. Reactions 2 and 5 were adjusted to fine tune the CO, CO_2, H_2 and H_2O components. Once these 4 components along with CH_4 were within less than 0.1%; the reactions for the NH_3, HCN, COS, and H_2S components were adjusted as required.

Table A.2a Temperature Approach for the Air-Blown Case

Reactions	ΔT
1. $C + 2H_2 = CH_4$	-135.6
2. $CO + H_2O = CO_2 + H_2$	-88
3. $H_2 + S = H_2S$	165
4. $C + O_2 = CO_2$	0.0
5. $C + 0.5O_2 = CO$	3.5
6. $0.5N_2 + 1.5H_2 = NH_3$	-805
7. $CO + S = COS$	240
8. $NH_3 + CO = HCN + H_2O$	64

Table A.2b Temperature Approach for the Oxygen-Blown Case

Reactions	ΔT
1. $C + 2H_2 = CH_4$	-116.2
2. $CO + H_2O = CO_2 + H_2$	-88
3. $H_2 + S = H_2S$	170
4. $C + O_2 = CO_2$	0.0
5. $C + 0.5O_2 = CO$	3.5
6. $0.5N_2 + 1.5H_2 = NH_3$	-980
7. $CO + S = COS$	178
8. $NH_3 + CO = HCN + H_2O$	1,030

A.1.1.4 Results

The model was able to match GTI's results with a high level of accuracy. The results from the ASPEN models compared to GTI's results are shown in Tables A.3a and A.3b. For the air-blown case, CO had the largest delta of +11.99 pounds per hour out of 31,146.8 pounds per hour and for the oxygen-blown case CO yielded that largest delta of -1.6 pounds per hour out of 24,107.4 pounds per hour. The air-blown model predicts the overall HHV of the main syngas components going to the gas turbine (CO, H_2, and CH_4) within -0.011% of the GTI results. For the air-blown case: GTI = 302.269 MBtu/hr and Model = 302.234 MBtu/hr. The oxygen-blown model predicts the overall HHV of the syngas going to the gas turbine within -0.0046% of the GTI results: GTI = 308.39 MBtu/hr and Model = 308.37 MBtu/hr. This model is valid for various size gasifiers and should be applicable for perturbations within reasonable coal properties.

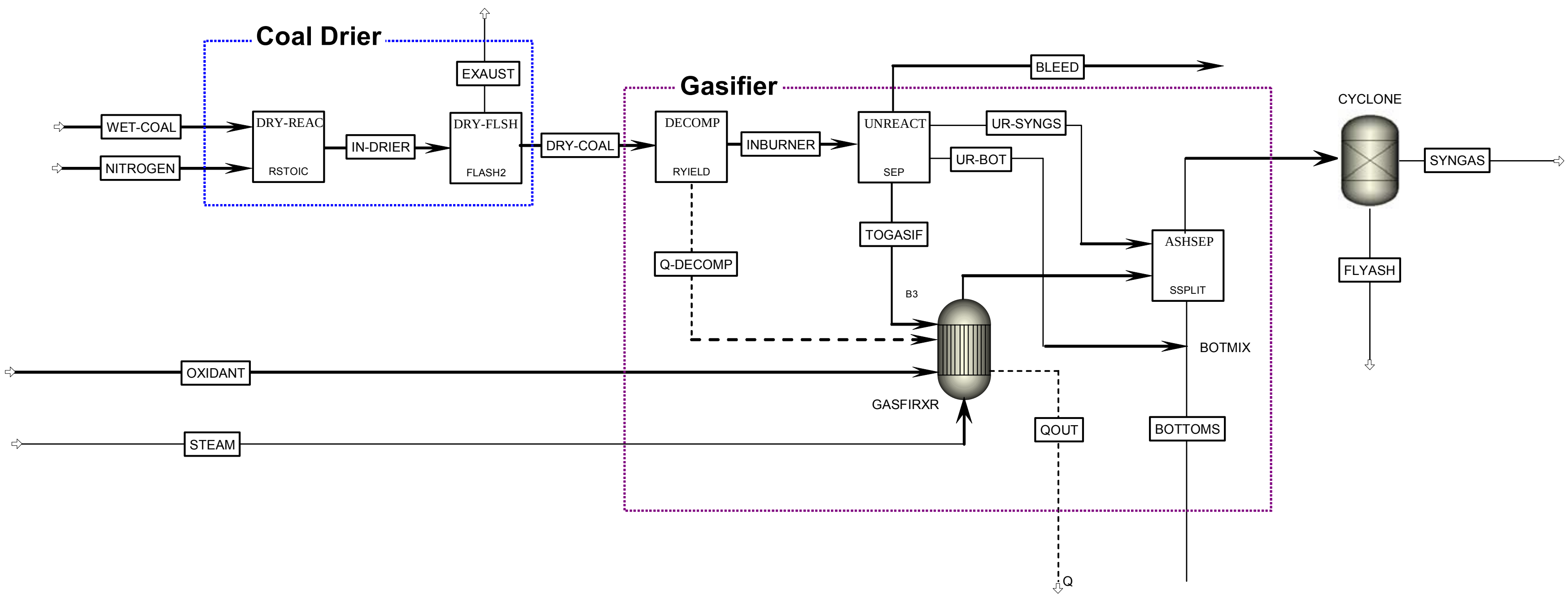
Table A.3a Results for the Air-Blown Case

Stream Composition	GTI (lb/hr)	Model (lb/hr)	Δ (lb/hr)	% Error
CO	31,158	31,146.8	-11.17	-0.04%
CO_2	16,221	16,233.0	11.99	0.07%
H_2	1,309	1,308.4	-0.62	-0.05%
H_2O	6,183	6,180.0	-3.01	-0.05%
CH_4	3,650	3,652.2	2.18	0.06%
H_2S	943	942.7	-0.31	-0.03%
COS	42	41.8	-0.16	-0.38%
NH_3	135	134.7	-0.25	-0.2%
HCN	17	17.2	0.21	1.2%
N_2	73,623	73,622.7	-0.28	<0.001%

Table 3b Results for the Oxygen-Blown Case

Stream Composition	GTI (lb/hr)	Model (lb/hr)	Δ (lb/hr)	% Error
CO	24,109	24,107.4	-1.61	-0.01%
CO_2	25,201	25,202.6	1.57	0.01%
H_2	2,036	2,035.6	-0.40	-0.02%
H_2O	19,313	19,311.4	-1.56	-0.01%
CH_4	3,337	3,337.7	0.71	0.02%
H_2S	905	904.6	-0.40	-0.04%
COS	40	40.2	0.18	0.45%
NH_3	126	126.9	0.89	0.71%
HCN	16	15.7	-0.27	-1.7%
N_2	1,152	1,152.0	-0.02	-0.002%

Figure A.1 ASPEN Block Diagram



A.1.2 Syngas Cleanup System

Aspen Plus provides a number of physical property methods for calculation of stream thermodynamic 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, cautions need to be exercised in evaluating the syngas water scrubber and the flash drum downstream of the low temperature heat recovery system, since some of the gases are dissolved in the sour water and process condensate, which will be treated in the sour water stripper. It is important to realistically estimate the sour 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 an acid gas removal system, the ElectrolyteNRTL property method set is suggested. A portion of the NH_3, H_2S, and CO_2 in the syngas are dissolved in the sour water and process condensate. To make sure that the acid gases in the sour water and process condensate are correctly accounted for, the Aspen Plus simulation developed for the current design incorporates the results obtained using both the ElectrolyteNRTL method and the 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 wastewater treatment system.

A.1.3.2 Sour Water Streams

The largest sour water feed is from the water scrubber down stream of the high temperature heat recovery boiler. A portion of the process condensate is also 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 (<10 microns) that are not removed by the cyclone particulate removal system. Some condensed oils (benzene and toluene derivatives) also may be in the sour water.

A.1.3.3 Sour Water Stripper

The sour water treatment unit processes the effluent from the syngas water scrubber and the process condensate from the flash drum upstream of the amine system. This unit consists of a flash drum, settling tank, day tank, sour water stripping column, and associated heat exchangers and pumps. The settling tanks remove particulates and insoluble oils. The filter presses dewater the agglomerated sludge from the bottom of the settling tanks. The day tank provides for water storage during stripper outages.

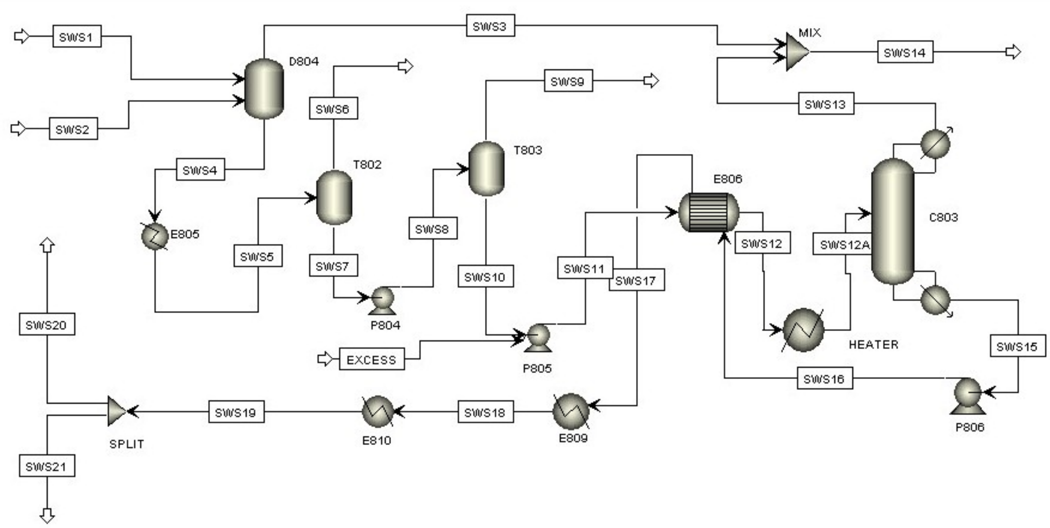
Vapors from the flash drum and stripping column are sent to the sulfur plant. Stripped water from the bottom of the column is recycled to the water scrubber with a blowdown stream sent to the wastewater treatment plant. Any necessary purges from the settling tanks and day tank are sent to the flare.

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

A.1.3.4 Sour Water Stripper Modeling – Air-Blown Case

The sour water stripper was modeled using ASPEN Plus Version 11.1. Figure A.2 shows the ASPEN process flowsheet for the air-blown case.

Figure A.2 ASPEN Process Flowsheet – Air-Blown Case



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

The flash drum (Figure A.2, D-804) was specified (atmospheric pressure, zero duty) to provide a bottom stream of saturated liquid near atmospheric pressure. This condition is necessary so that no significant off-gassing would occur in the settling tank (T-802) or the day tank (T-803). As a design margin, a cooler (E-805) was used to cool the liquid to 10°F below saturation temperature. The settling tank and day tank were modeled as

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

simple flash drums with zero pressure drop and zero heat duty. Because the day tank was designed for the storage of one day accumulation of sour water, an excess stream (EXCESS) was included that was specified as the same composition of the material entering the day tank at one-tenth the mass flow rate to allow for sizing the equipment at 110% of the design rate. For operating stream compositions (heat and material) the excess stream was specified as zero flow.

The preheater (E-806) was modeled as a countercurrent heat exchanger to recover heat from the column bottoms stream. Because ASPEN blocks are calculated sequentially, an additional heat exchanger (HEATER) was used for the initial modeling run (zero pressure drop and temperature specified near column operation) to speed up model convergence. After the column (see below) converged, the duty of this additional heat exchanger (HEATER) was set to zero and the model was re-run.

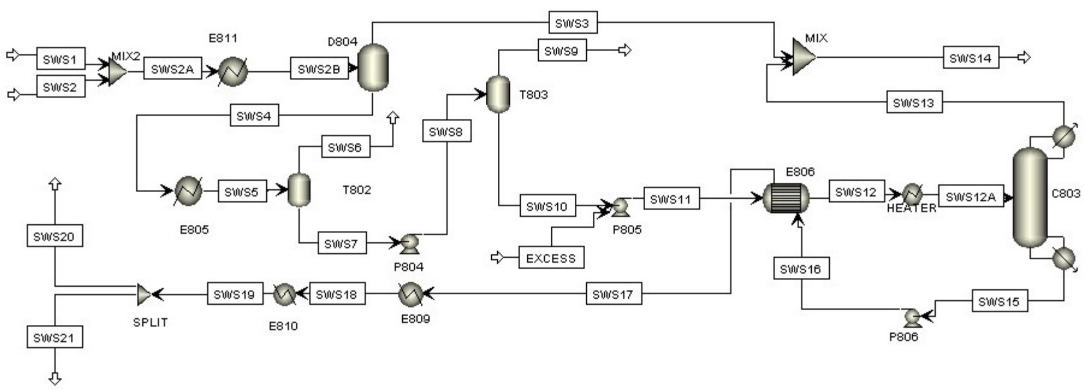
The column (C-803) was modeled using a RADFRAC block. The column was specified similar to the design described in the literature with 20 actual stages (10 theoretical stages) plus a partial condenser and kettle reboiler. Sieve trays were specified for the design. The kettle reboiler duty was specified to meet the desired discharge water composition (<50 ppmw NH_3). Meeting the ammonia specification also satisfied the H_2S specification (<10 ppmw). The reflux ratio was manually adjusted until the mass flow of the water was between 25 wt% and 30 wt% of the total vapor stream to the sulfur plant.

Cooling of the hot stream downstream of the pre-heater (E-806) was modeled using two coolers in series based on the assumption that 140°F is the economic break-point between an air-finned cooler (E-809) and a water cooler (E-810). The cooled water stream at 110°F is split between recycle water to the water scrubber and discharge to the wastewater treatment plant.

A.1.3.5 Sour Water Stripper Modeling – Oxygen-Blown Case

The ASPEN modeling for the oxygen-blown case is fundamentally similar to the air-blown case except for the mixing and cooling (E-811) of the sour water streams prior to the first flash drum. A different composition of the sour water streams required a lower temperature to drive off more CO_2 while maintaining a low mass flow of water in the vapor stream leaving the first flash. Figure A.3 represents the ASPEN process flowsheet for the oxygen-blown case.

Figure A.3 ASPEN Process Flowsheet – Oxygen-Blown Case



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 Addendum G of this report. Modeling of the power block was accomplished using the GateCycle computer modeling program for Windows Version 5.34.0.r, and for each case, air- and oxygen-blown, a single model represented 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 turbine selected for the facility is the General Electric GE10 (11.25 MW ISO conditions, natural gas DLE). For the purpose of modeling, a Nuovo Pignone PGT10B turbine (a forerunner to the GE10. General Electric (GE) had supplied some performance data previously for the Nuovo Pignone turbine) was selected from the GateCycle turbine library. Syngas composition was generated using ASPEN 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.

It is important to note that 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 is 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 GE10 turbine is not commercially demonstrated for use on coal-derived syngas, there is not currently sufficient data available for use. Use of the GE software library PGT10B data provided results reasonably consistent with GE performance data mentioned above.

Another area of uncertainty lies in the performance of the CT for the oxygen-blown case. Prior information from GE indicated that some method of combustion NOx control (e.g., water or steam injection) would be required for the oxygen-blown case. However, the information from GE is insufficient for use in GateCycle modeling, and for this analysis the syngas from the oxygen-blown case was diluted with steam to an LHV equivalent to that of the air-blown case.

A.2.3 HRSG Modeling

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

- Stack temperature remained above the acid-dew point so that condensation and corrosion did not occur within the system (~240ºF).
- 50 psig superheated steam (~353ºF) was generated such that the process steam demands of all gasifier and gas clean-up processes were self sufficient (including gas clean up operations and sour water treatment).
- Balance of steam generation was 400 psig superheated steam (~548ºF).

The modeling was accomplished by inserting the appropriate HRSG components downstream of the turbine exhaust. Figure A.4 provides a screen capture of the GateCycle flow diagram used for both the air-blown and oxygen-blown cases.

Figure A.4 Gate Cycle Process Flow Diagram – Air-Blown and Oxygen-Blown Cases

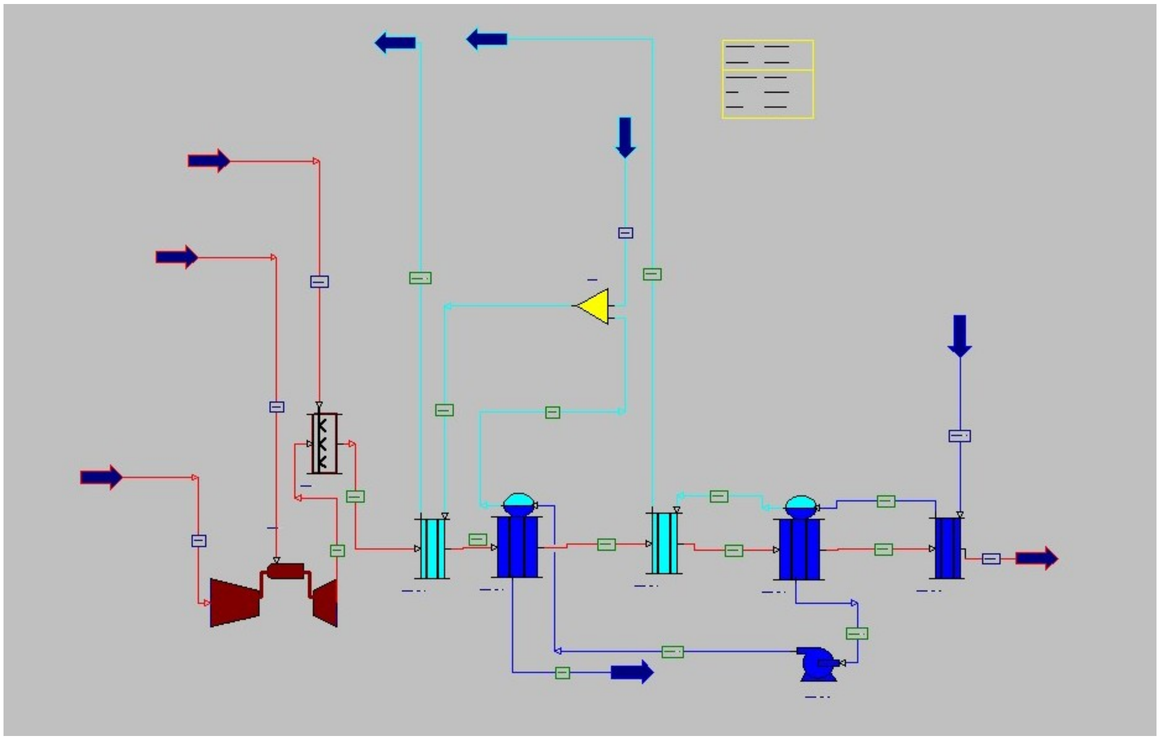


Figure A.4 shows the HRSG was modeled to produce both 400 psig and 50 psig superheated steams to meet the three requirements described above. The key input parameters were stack outlet temperature, degree of superheat for 50 psig and 400 psig steam, 400 psig steam input (from gasification operations), and pinch ΔT of the 400 psig evaporator. 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.

Also shown in Figure A.4 is a duct-burner between the turbine exhaust and the 400 psig superheater. The duct-burner was included to provide the flexibility to evaluate other scenarios and conduct trade-off studies. When not in use the duct-burner was specified in a manner that eliminated its presence in the system (zero fuel flow, zero pressure drop). Once all blocks were assigned inputs, including pressure drop and blow-down from the 400 psig evaporator, GateCycle was run and heat and material balances were calculated. Based on the heat and material balances, GateCycle also calculated the steam production and surface area of each block.

A.2.4 Comparison of Air-Blown Process with Oxygen-Blown Process

Because the syngas LHV of both cases was very similar (see explanation of steam dilution above), the primary differences in HRSG design for the air-blown case and the oxygen-blown case was in the surface areas of the evaporators and the 400 psig superheater. Table A.4 presents the calculated surface area for each of the HRSG components.

Table A.4 Surface Area Comparison

(Square feet)

	Air-Blown	Oxygen-Blown
Economizer	17,031	16,804
50 psig Evaporator	46,788	50,101
50 psig Superheater	400	403
400 psig Evaporator	28,328	19,256
400 psig Superheater	1,900	1,581

The differences in HRSG surface area for the two cases was a direct result of the increased 50 psig steam demand and decreased 400 psig steam import from the HTHR system for the oxygen-blown case. Because the oxygen-blown case has a higher demand for 50 psig steam there is less energy available for 400 psig steam generation, and therefore, smaller surface areas for the 400 psig components. Although the oxygen-blown case produced more 50 psig steam than the air-blown case, the size differences for the 50 psig superheater and evaporator were not as great as those for the 400 psig system. This is because the oxygen-blown case has higher gas inlet temperatures to both the 50 psig superheater and evaporator (due to the lower production rate of 400 psig steam).

B.1 COAL PREPARATION

The equipment included in the coal preparation area (Area 100) 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 AIR-BLOWN CASE

The equipment in Areas 150 though 1000 for the air-blown case includes:

Area 150 Air or Oxygen Supply				
Identification	No.	Description	Comments	Unit Size
K-151	2	Air Compressor package	850 GPM cooling water W/ 25F delta	4000bhp
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 ft^2
Area 200 Coal Feeding				
Identification	No.	Description	Comments	Unit Size
T-201	2	Weigh Hopper		
D-202	2	Lock Hopper		
D-203	2	Surge Hopper		
S-201	2	Rotary Feeder		
S-203	2	Screw Feeder		
Area 300 Gasification				
Identification	No.	Description	Comments	Unit Size
R-301	2	Gasifier		
		Refractory		
		Internals		
H-301	2	Startup Heater		

Area 400 Dust Removal				
Identification	No.	Description	Comments	Unit Size
CY-401	2	Primary Cyclone		
CY-402	2	Secondary Cyclone		
CY-403	2	Tertiary Cyclone		
		Refractory		
		Connecting Refractory Pipe		
D-401	2	Cyclone Surge Hopper		
		Refractory		
S-401	2	Cyclone Transport Screw		
D-402	2	Cyclone Lock Hopper		
T-403	2	Cyclone Pneumatic Transport Hopper		
S-403	2	Dust Feeder		
T-404	1	Dust Storage Silo		
Area 500 Ash Removal				
Identification	No.	Description	Comments	Unit Size
D-501	2	Ash Surge Hopper		
		Refractory		
S-501	2	Ash Transport Screw		
D-502	2	Ash Lock Hopper		
T-503	2	Ash Pneumatic Transport Hopper		
S-503	2	Ash Feeder		
T-504	1	Ash Storage Silo		
Area 600 Gas Cooling				
Identification	No.	Description	Comments	Unit Size
D-601 NA/SA	2	High Pressure Steam Drum	CS Horizontal, D=7.5ft L=41ft P_{des} =470psig T_{des} = 500F	D = 5ft, and L = 17 ft
E-601 NA/SA	2	High Pressure Steam Boiler	SHELL: DP= 450 psig; DT= 550 F; TUBE: DP= 385 psig;DT= 1800 F; Inconel Tubes	560 ft^2
P-601	1	HP Steam Boiler Start-up Pump		flowrate: 250 gpm; head = 100 ft, 9.9hp
P-602	3	Fresh Quench Water pump		flowrate: 10 gpm; head = 420 psia, 2.5hp

Area 700 Gas Cleaning				
Identification	No.	Description	Comments	Unit Size
C-701 NA/SA	2	Syngas Scrubber Column	Inerts = 410SS, Vessel = CS 8 Trays, P_{des} = 352psig, T_{des}=295F	D=3.27 ft h =21 ft
E-701	1	COS Hydrolysis Reactor Preheater	SHELL: DP= 450 psig; DT= 600 F; TUBE: DP= 385 psig;DT= 325 F; CS Tubes , AEU	88 ft^2
E-702	1	COS Reactor Effluent Cooler/BFW H	SHELL: DP= 500 psig; DT= 300 F; TUBE: DP= 385 psig;DT= 325 F; CS Tubes , AEU	2224 ft^2
E-703	1	Effluent air cooler	cool effluent to 140 F with air	28 hp, surface area = 38.1x10^3 ft^2, bare tube area = 2350 ft^2; 410SS tubes
E-704	1	COS Reactor Effluent Water Cooler	SHELL: DP= 150 psig; DT= 170 F; TUBE: DP= 385 psig;DT= 190 F; 410SS Tubes , AEU	1291 ft^2
P-701 A/B/C	3	Condensate Recirculation Pump		6 gpm, dp = 150 psi, efficiency = 0.95, 110 F recycled condensate water, 0.6hp
P-702 A/B/C	3	BFW Pump		53 gpm, dp = 400 psi, efficiency = 0.95, 150 F BFW, 42 hp
P-703 A/B	3	Recycle SWS Water Pump		12 gpm, dp = 400 psi; e = 0.95, 6.0 hp
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 condenser drum	vertical drum, pres = 300 psia, DP=315 psig, DT=160F, CS	Dia = 4, H = 5.1 ft,
S-701	Lot	Sud Chemie C53-2-01 1/8" catalyst	Loading supplied by Sud Chemie	
R-711	1	Mercury Adsorption Vessel	Information supplied by Calgon Carbon	9.0 ft ID by 10.0 ft TT
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-803	1	Sour Water Stripper Column	Internals = 410SS, Vessel = CS 21 Trays, Pdes = 50psig, Tdes=317F	D=3 ft Tray Spacing=2 ft, T-T=60 FT
D-804	1	Sour Water Feed Flash Drum	410SS clad CS Horizontal, Pdes =50psig Tdes = 290F	D=7.7ft L=10.3ft
D-805	1	Sour Water Distillate Drum	410SS clad CS Horizontal, Pdes =45psig Tdes = 236F	D=4.9ft L=5.1ft
E-805	1	Settler Pre-Cooler - Air Fin	SS Construction, Fan HP = 5	Bare tube area = 260 ft2
E-806	1	SWS Feed Pre-Heater	TEMA Type AEU, 410SS tubes	area = 830.1 ft2
E-807	1	SWS Condenser - Air Fin	SS Construction, Fan HP = 24	Bare tube area = 1190 ft2
E-808	1	SWS Kettle Reboiler	TEMA Type BKU, CS	area = 675.7 ft2
E-809	1	Recycle Water Cooler - Air Fin	CS Construction, Fan HP = 11	Bare tube area = 540 ft2
E-810	1	Recycle Water Cooler - Water Coole	TEMA Type AEU, CS Construction	area = 170.5 ft2
F-804A/B	2	Filter Press	Information supplied by US Filter	25.6ft x 7.75ft
P-804A/B	2	Post Day Tank Pump		103 gpm, 29psi dp, 1.8bhp
P-805A/B	2	SWS Reflux Pump		23.7 gpm, 25psi dp, 0.4bhp
P-806A/B	2	Stripper Bottom Pump		106 gpm, 30psi dp, 1.9bhp
P-807A/B	2	Post Settling Tank Water Pump		93 gpm, 20psi dp, 1.1bhp
P-808 A/B/C/D/E	5	Slurry Pump	2 - 100psi air powered diaphragm pumps per filter press	1.5 gpm, 100psi dp
P-809 A/B	2	Sour Water Sump Pump	to pump sour water from sump to day tank	1.5 gpm, 11 psi dp
T-802	2	Settling Tank	plus weir, and cone bottom Pdes =16.2psig Tdes = 236F	D=20ft H=15ft
T-803	1	Sour Water Storage (Day Tank)	Pdes =25psig Tdes = 236.4F	D=40ft H=16ft
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 ERI	

Area 1000 Offsites and Auxiliaries				
Identification	No.	Description	Comments	Unit Size
		Steam generation system	all steam generated on-site	
		Condensate collection system	**equipment:** cooler, storage tank, pumps, polisher, deaerator	137 gpm
		Demineralized water system	equipment: feed tank, feed pumps, mixed bed exchangers, storage tank, blowers, caustic storage tank heater, caustic dilution heater, acid pumps, caustic pumps, regeneration pumps, caustic storage, caustic dilution tank; (reverse osmosis option to be a trade-off study)	
		Cooling water system	supply temp = 80°F, return temp = 100°F, **equipment:** cooling tower, circulation pumps	3650 gpm
		Safety shower/eye wash system	installed cost	
		Raw water/fire water system	**equipment:** feed tank, feed pumps, blowers, filter, sludge pumps, backwash pumps, soda ash pumps, lime slurry pumps, sulfuric acid pumps, coagulant metering pumps, spray water booster pumps, polymer feed systems, filter press, softener reactor, sludge thickener, sulfuric acid storage tank, lime silo, soda ash silo, lime dilution tank, soda ash dilution tank, coagulant feed tank	115 kgpd
		Drinking (potable) water system	installed cost	
		Compressed air system	**equipment:** 3 compressors (2 working, 1 stand-by), desiccant air dryers, IA receiver tank, PA receiver tank	1200 scfm
		Natural gas supply system		
		Flare system	**equipment:** elevated flare, pilot and knock out drum	140 MM Btu/hr
		Nitrogen system		system 1 = 14,200 scfh; system 2 = 900 scfh
		Waste water collection, treatment and disposal system		
		Electrical distribution system		consumption = 7.25 MW; export = 22.25 MW
		Interconnecting piping		
		Telecommunications systems		

B.3 OXYGEN -BLOWN CASE

The equipment in Areas 150 though 1000 for the oxygen-blown case includes:

Area 150 Air or Oxygen Supply				
Identification	No.	Description	Comments	Unit Size
S-161	1	Air Separation Plant	Per Air Products quote	230 tpd
	2	Oxygen Compressor	Dresser-Rand	555.2 BHP
E-	2	Heat Exchanger	SHELL: DP= 450 psig; DT= 600 F; TUBE: DP= 450 psig;DT= 600 F; CS Tubes , AEU	101 ft^2
Area 200 Coal Feeding				
Identification	No.	Description	Comments	Unit Size
T-201	2	Weigh Hopper		
D-202	2	Lock Hopper		
D-203	2	Surge Hopper		
S-201	2	Rotary Feeder		
S-203	2	Screw Feeder		
Area 300 Gasification				
Identification	No.	Description	Comments	Unit Size
R-301	2	Gasifier		
		Refractory		
		Internals		
H-301	2	Startup Heater		

Area 400 Dust Removal				
Identification	No.	Description	Comments	Unit Size
CY-401	2	Primary Cyclone		
CY-402	2	Secondary Cyclone		
CY-403	2	Tertiary Cyclone		
		Refractory		
		Connecting Refractory Pipe		
D-401	2	Cyclone Surge Hopper		
		Refractory		
S-401	2	Cyclone Transport Screw		
D-402	2	Cyclone Lock Hopper		
T-403	2	Cyclone Pneumatic Transport Hopper		
S-403	2	Dust Feeder		
T-404	1	Dust Storage Silo		
Area 500 Ash Removal				
Identification	No.	Description	Comments	Unit Size
D-501	2	Ash Surge Hopper		
		Refractory		
S-501	2	Ash Transport Screw		
D-502	2	Ash Lock Hopper		
T-503	2	Ash Pneumatic Transport Hopper		
S-503	2	Ash Feeder		
T-504	1	Ash Storage Silo		
Area 600 Gas Cooling				
Identification	No.	Description	Comments	Unit Size
D-601 NA/SA	2	High Pressure Steam Drum	CS Horizontal, P_{des} =470psig T_{des} = 500F	Dia=4.6ft L=15.6f
E-601 NA/SA	2	High Pressure Steam Boiler	SHELL: DP= 450 psig; DT= 550 F; TUBE: DP= 385 psig;DT= 1800 F; Inconel Tubes	435.3 ft^2
P-601	1	HP Steam Boiler Start-up Pump		flowrate: 250 gpm; head = 100 ft, 9.9hp
P-602	3	Fresh quench water pump		flowrate: 4 gpm; head = 420, 1hp

Area 700 Gas Cleaning				
Identification	No.	Description	Comments	Unit Size
C-701 NA/SA	2	Syngas Scrubber Column	Inerts = 410SS, Vessel = CS 8 Trays, P_{des} = 352psig, T_{des}=295F	D=2.65 ft h =28.5 ft
E-701	1	COS Hydrolysis Reactor Preheater	SHELL: DP= 450 psig; DT= 600 F; TUBE: DP= 385 psig;DT= 325 F; CS Tubes , AEU	48.3 ft^2
E-702	1	COS Reactor Effluent Cooler/BFW Heater	SHELL: DP= 500 psig; DT= 300 F; TUBE: DP= 385 psig;DT= 325 F; CS Tubes , AEU	1738 ft^2
E-703	1	Effluent air cooler	cool effluent to 140 F with air	12.2 hp, surface area = 16.4x10^3 ft^2, bare tube area = 1070 ft^2 410 SS tubes
E-704	1	COS Reactor Effluent Water Cooler	SHELL: DP= 150 psig; DT= 170 F; TUBE: DP= 385 psig;DT= 190 F; 410SS Tubes , AEU	709 ft^2
P-701 A/B	3	Condensate Recirculation Pump		3 gpm, dp = 150 psi, efficiency = 0.95, 110 F recycled condensate water., 0.3 hp
P-702 A/B	3	BFW Pump		41 gpm, dp = 400 psi, efficiency = 0.95, 150 F BFW, 10.1 hp
P-703 A/B	3	Recycle SWS Water Pump		92 gpm, dp = 400 psi; e = 0.95, 23 hp
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
D-701	1	Effluent condenser drum	vertical drum, Des P=330 psig, DesT=160F, CS	a=5.5ft L=11.5ft
S-701	Lot	Sud Chemie C53-2-01 1/8" catalyst	Loading supplied by Sud Chemie	
R-711	1	Mercury Adsorption Vessel	Information supplied by Calgon Carbon	9.0 ft ID by 10.0 ft TT

Area 800 Acid Gas Removal and Sulfur Recovery				
Identification	No.	Description	Comments	Unit Size
C-803	1	Sour Water Stripper Column	Internals = 410SS, Vessel = CS 21 Trays, Pdes = 50psig, Tdes=317F	D=4.75 ft Tray Spacing=2 ft, T-T=60 FT
D-804	1	Sour Water Feed Flash Drum	410SS clad CS Horizontal, Pdes =40psig Tdes = 221F	)=10.6ft L=16.8ft
D-805	1	Sour Water Distillate Drum	410SS clad CS Horizontal, Pdes =45psig Tdes = 258.6F	D=5.8ft L=6.9ft
E-805	1	Sour Water Sub-Cooler - Air Fin	SS Construction, Fan HP = 4	Bare tube area = 220 ft2
E-806	1	SWS Feed Pre-Heater	TEMA Type AEU, 410SS tubes	area = 3689 ft2
E-807	1	SWS Condenser - Air Fin	SS Construction, Fan HP = 36	Bare tube area = 1820 ft2
E-808	1	SWS Kettle Reboiler	TEMA Type BKU, CS	Bare tube area = 1336.2 ft2
E-809	1	Recycle Water Cooler - Air Fin	CS Construction, Fan HP = 21	Bare tube area = 1070 ft2
E-810	1	Recycle Water Cooler - Water Cool	TEMA Type AEU, CS Construction	area = 546.6 ft2
E-811	1	Sour Water Pre-Cooler	SS Construction, Fan HP = 31	Bare tube area = 1540 ft2
F-804	2	Filter Press	Information supplied by US Filter	25.6ft x 7.75ft
P-804A/B	2	Post Day Tank Pump		325 gpm, 60psi dp, 4.6bhp
P-805A/B	2	SWS Reflux Pump		45.3 gpm, 25psi dp, 0.7bhp
P-806A/B	2	Stripper Bottom Pump		339 gpm, 30psi dp, 6.2bhp
P-807A/B	2	Post Settling Tank Pump		296 gpm, 20psi dp, 3.6bhp
P-808 A/B/C/D/E	5	Slurry Pump	2 - 100psi air powered diaphragm pumps per filter press	1.5 gpm, 100psi dp
P-809 A/B	2	Sour Water Sump Pump	to pump sour water from sump to day tank	1.5 gpm, 11 psi dp
T-802	2	Settling Tank	plus weir, and cone bottom Pdes =16.2psig Tdes = 211F	D=28ft H=21ft
T-803	1	Sour Water Storage (Day Tank)	Pdes =25psig Tdes = 236.4F	460,200, gal (45' DIA X 40')

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 ERI	

Area 1000 Offsites and Auxiliaries				
Identification	No.	Description	Comments	Unit Size
		Steam generation system	all steam generated on-site	
		Condensate collection system	**equipment:** cooler, storage tank, pumps, polisher, deaerator	100 gpm
		Demineralized water system	**equipment:** feed tank, feed pumps, mixed bed exchangers, storage tank, blowers, caustic storage tank heater, caustic dilution heater, acid pumps, caustic pumps, regeneration pumps, caustic storage, caustic dilution tank; (reverse osmosis option to be a trade-off study)	
		Cooling water system	supply temp = 80°F, return temp = 100°F, **equipment:** cooling tower, circulation pumps	5000 gpm
		Safety shower/eye wash system		
		Raw water/fire water system	**equipment:** feed tank, feed pumps, blowers, filter, sludge pumps, backwash pumps, soda ash pumps, lime slurry pumps, sulfuric acid pumps, coagulant metering pumps, spray water booster pumps, polymer feed systems, filter press, softener reactor, sludge thickener, sulfuric acid storage tank, lime silo, soda ash silo, lime dilution tank, soda ash dilution tank, coagulant feed tank	158 kgpd
		Drinking (potable) water system		
		Compressed air system	**equipment:** 3 compressors (2 working, 1 stand-by), desiccant air dryers, IA receiver tank, PA receiver tank	1200 scfm
		Natural gas supply system		
		Flare system	**equipment:** elevated flare, pilot and knock out drum	140 MM Btu/hr
		Nitrogen system		system 1 = 14,200 scfh; system 2 = 900 scfh
		Waste water collection, treatment and disposal system		
		Electrical distribution system		consumption = 5.64 MW; export = 22.86 MW
		Interconnecting piping		
		Telecommunications systems		

The inputs into the financial model are listed below. A greater elaboration on some of the assumptions made:

- Fees: Per NETL guidelines and past team experience, a 10% fee rate was used to include project development, start-up costs, licensing, permitting, spares, training, construction management, commissioning, transportation, and owner's costs. This fee was entered in the model by placing 4% as the development fee, 2% as start-up costs, and the remainder of the fee in "owner's costs". Owner's costs are a dollar value calculated to equal 10% of the plant EPC cost when combined with the development fee and start-up costs.

Interest during construction, financing costs, and working capital requirements have all been entered separately.

Unit engineering and installation are already included in the plant EPC costs. These factors were used when scaling-up the as-built unit costs.

Royalties and land costs have not been included in the model. Since this project is for existing industrial applications, it is assumed that the land cost is negligible. Royalties will vary considerably based on the technology vendor.

- Fixed and Variable Operations and Maintenance Costs: The entries for fixed and variable O&M costs reflect work done on previous gasification studies and NETL guidelines, with a few modifications for the plant location. Typical default values for O&M combine to equal 4.2% of the plant EPC cost. However, this number reflects a United States Gulf Coast (USGC) plant site. Costs for labor and equipment in upstate New York were found to be 15-25% higher than USGC costs, increasing the O&M costs to 5%. This number is consistent with NETL guidelines for plant analysis.

- Operating Hours: Operating hours are defined as the total hours available for plant operation after scheduled outage time has been deducted. The amount of planned plant downtime for scheduled maintenance, 21 days, is based on the estimated requirements for the gasifier as provided by GTI. It is assumed that other scheduled maintenance can be done during this time. While this number will vary throughout the life of the plant due plant turnarounds and major maintenance, this was determined to best reflect total planned plant outages throughout the life of the facility

- Start-Up Scenarios: The financial model allows the user to input a different availability for the first two years of plant operation than that for the rest of the plant life. From previous design and operations experience, the total first year availability was estimated to be 69%. The second year of operation was set to be the same as that for the remaining years.

Table C.1 Financial Model Entries—Plant Inputs

	Air Base Case	Oxygen Base Case
Project Name	Industrial Gasification	Industrial Gasification
Project Location	Upstate NY	Upstate NY
Primary Output/Plant Application **(Options: Power, Multiple Outputs)**	Multiple Outputs	Multiple Outputs
Primary Fuel Type (**Options: Gas, Coal, Petroleum Coke, Other/Waste)**	Coal	Coal
Secondary Fuel Type (**Options: None, Gas, Coal, Petroleum Coke, Other/Waste)**	None	None
Plant Output and Operating Data : Note - All ton units are US Short Tons (2000 lbs)		
Syngas Capacity (Mcf/Day)	0	0
Gross Electric Power Capacity (MW)	29.8	29.7
Net Electric Power Capacity (MW)	21.7	23.3
Steam Capacity (Tons/Hr)	50.9	13.4
Hydrogen Capacity (Mcf/Day)	0	0
Carbon Monoxide Capacity (Mcf/Day)	0	0
Elemental Sulfur Capacity (Tons/Day)	10.9	10.4
Slag Ash Capacity (Tons/Day)	25.2	17.6
Fuel (Tons/Day)	0	0
Chemicals (Tons/Day)	0	0
Environmental Credit (Tons/Day)	0	0
Operating Hours per Year	8256	8256
Guaranteed Availability (percentage)	0.909	0.877
Enter One of the Following Items(For Each Primary/Secondary Fuel) Depending on Project Type:		
Primary Fuel Heat Rate (Btu/kWh) based on HHV **FOR POWER PROJECTS**	0	0
Secondary Fuel Heat Rate (Btu/kWh) based on HHV **FOR POWER PROJECTS**	0	0
Primary Fuel Annual Fuel Consumption (in Mcf *OR* Thousand Tons) **FOR NON POWER PROJECTS**	113.5	102.8
Secondary Fuel Annual Fuel Consumption (in Mcf *OR* Thousand Tons) **FOR NON POWER PROJECTS**		
Initial Capital and Financing Costs (enter 'Additional Costs' in thousand dollars)		
EPC (in thousand dollars)	90,430	100,194
Owner's Contingency (% of EPC Costs)	15%	15%
Development Fee (% of EPC Costs)	4%	4%
Start-up (% of EPC Costs)	2%	2%
Owner's Cost (in thousand dollars) **COMBINED WITH DEVELOPMENT AND S/U TO = 10%**	3,617	4,008
Operating Costs and Expenses		
Variable O&M (% of EPC Cost)	1.5%	1.5%
Fixed O&M Cost (% of EPC Cost)	3.0%	3.0%
Additional Comments	Industrial Gasification Facility --AIR BLOWN	Industrial Gasification Facility--OXYGEN BLOWN

Table C.2 Financial Model Entries—Scenario Inputs

Capital Structure		
Percentage Debt	66%	
Percentage Equity	34%	
Project Debt Terms		
Loan 1: Senior Debt		
% of Total Project Debt **(total for Loans 1,2, and 3 must = 100%)**	100%	
Interest Rate	8%	
Financing Fee	3%	
Repayment Term (in Years)	15	
Grace Period on Principal Repayment	1	
First Year of Principal Repayment	2009	
Loan Covenant Assumptions		
Interest Rate for Debt Reserve Fund (DRF)	4%	
Debt Reserve Fund Used on Senior Debt **(Options: Yes or No)**	No	
Percentage of Total Debt Service used as DRF	20%	
Depreciation : "SL" for Straight-Line OR "DB" for 150% Declining Balance		**Method**
Construction (Years) : *Note - DB Method Must be 15 or 20 years*	15	SL
Financing (Years) : *Note - DB Method Must be 15 or 20 years*	15	SL
Working Capital		
Days Receivable	30	
Days Payable	30	
Annual Operating Cash (Thousand $)	$50	
Initial Working Capital (% of first year revenues)	7%	

ECONOMIC ASSUMPTIONS

Cash Flow Analysis Period	
Plant Economic Life/Concession Length (in Years)	20
Discount Rate	10%
Escalation Factors	
Project Output/Tariff	
Electricity: Capacity Payment	3.0%
Electricity: Energy Payment	3.0%
Steam	3.0%
Elemental Sulfur	3.0%
Slag Ash	3.0%
Fuel/Feedstock	
Gas	4.0%
Coal	2.0%
Petroleum Coke	2.0%
Other/Waste	2.0%
Operating Expenses and Construction Items	
Variable O&M	3.0%
Fixed O&M	3.0%
Other Non-fuel Expenses	3.0%
EPC Costs	3.0%
Tax Assumptions	
Tax Holiday (in Years)	0

Income Tax Rate	40%		
Subsidized Tax Rate (used as investment incentive)	0%		
Length of Subsidized Tax Period (in Years)	0		

FUEL/FEEDSTOCK ASSUMPTIONS

Fuel Prices : For the Base Year, then escalated by fuel factors in B71-B74 above	
Gas ($/Mcf)	4.68
Coal ($/US Short Ton) @ 5% moisture	34.60
Petroleum Coke ($/US Short Ton)	0.00
Other/Waste ($/US Short Ton)	0.00
Alternatively, use Forecasted Prices (From Fuel Forecasts Sheet)? (Yes/No)	No

TARIFF ASSUMPTIONS

INITIAL TARIFF LEVEL (In Dollars in the first year of construction)	
Electricity Payment ($/MWh)	80.00
Steam ($/US Short Ton)	12.00
Elemental Sulfur ($/US Short Ton)	26.52
Slag Ash ($/US Short Ton)	10.00

CONSTRUCTION ASSUMPTIONS	**Base Year :**	2005	
Construction Schedule	**A**		
Construction Start Date	5/1/2005		
Construction Period (in months)	32		
Plant Start-up Date *(must start on January 1)*	1/1/2008		
EPC Cost Escalation in Effect? (Yes/No)	No		
Percentage of Cost for Construction Periods	**Three Year Period**		
Enter for Five, Four or Three Year Periods (To the Right --->)	**Year 1**	**Year 2**	**Year 3**
Capital Costs : Unescalated Allocations	30.0%	50.0%	20.0%
EPC Costs: Escalated Allocations (Use EPC Escalation Sheet to Calculate)	28.6%	50.6%	20.8%
EPC Costs	0.0%	0.0%	100.0%
Initial Working Capital	0.0%	0.0%	100.0%
Owner's Contingency (% of EPC Costs)	30.0%	50.0%	20.0%
Development Fee (% of EPC Costs)	0.0%	30.0%	70.0%
Start-up (% of EPC Costs)	0.0%	0.0%	100.0%
Initial Debt Reserve Fund	70.0%	30.0%	0.0%
Owner's Cost (in thousand dollars) COMBINED WITH DEVELOPMENT AND S/U TO = 10%	0.0%	30.0%	70.0%
Interest During Construction	100.0%	0.0%	0.0%
Financing Fee	0.0%	30.0%	70.0%
Additional Financing Cost #1	0.0%	30.0%	70.0%
Plant Ramp-up Option (Yes or No)	Yes		

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

Year 2, First Quarter	90.9%
Year 2, Second Quarter	90.9%
Year 2, Third Quarter	90.9%
Year 2, Fourth Quarter	90.9%
Year 2 Average Capacity %	90.9%

Air-Blown and Oxygen-Blown Cases

Air-Blown Case

Oxygen-Blown Case

Figure D.1 Coal Handling System – Air- and Oxygen-Blown Cases – Simplified Flow Diagram

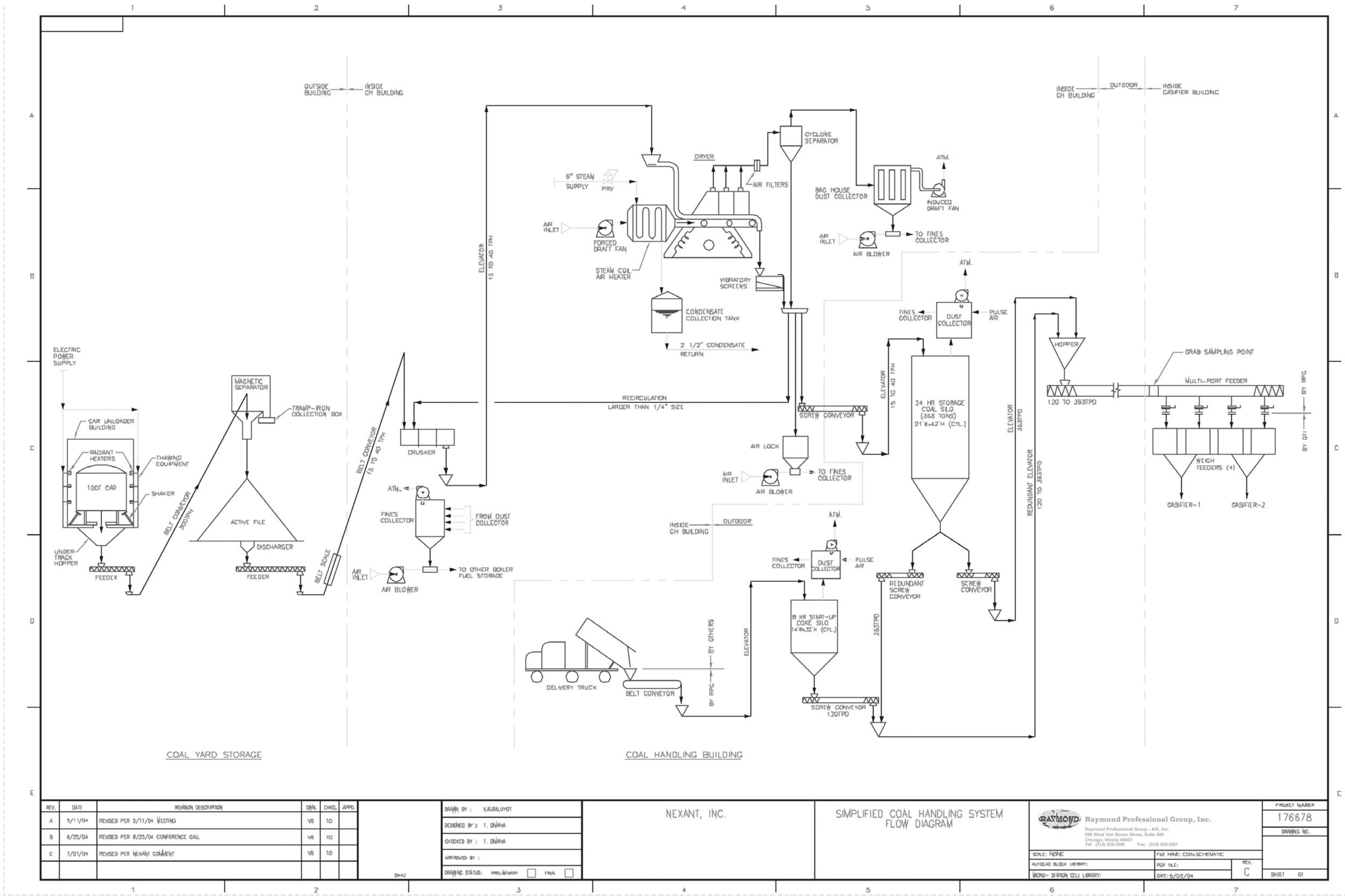
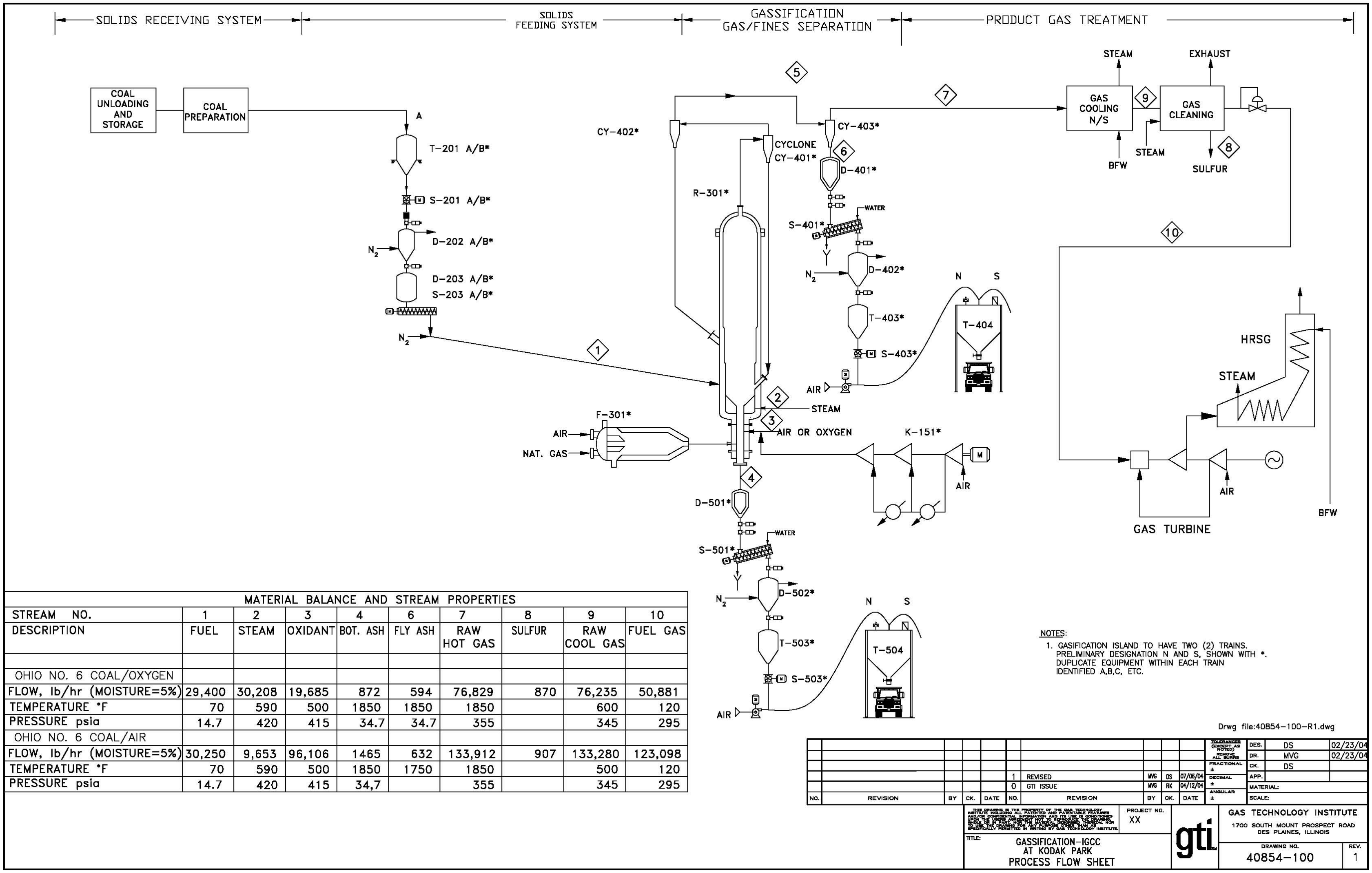


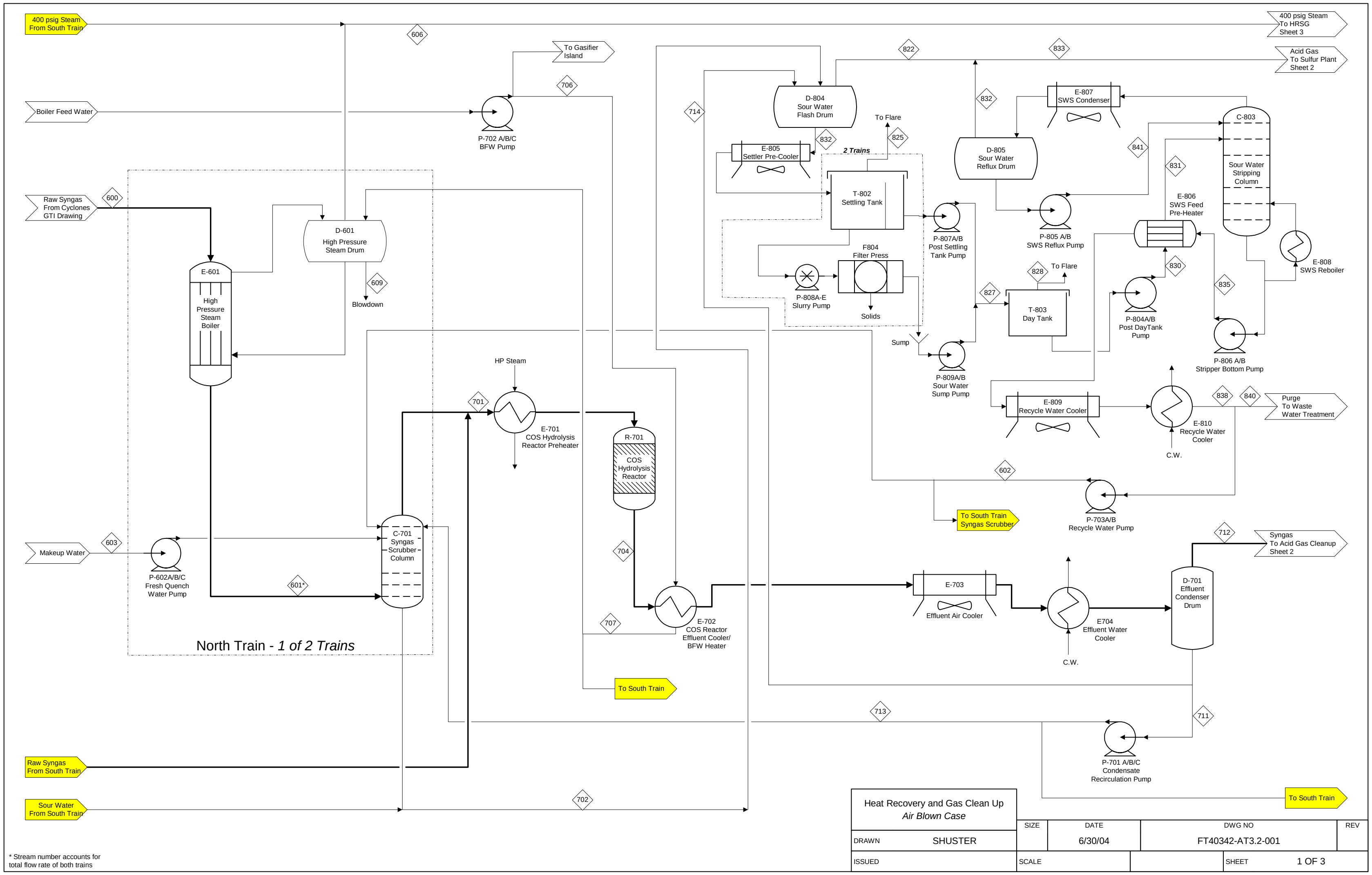
Figure D.2 Gasification – Air- and Oxygen-Blown Cases – IGCC Process Flow Sheet



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/OXYGEN									
FLOW, lb/hr (MOISTURE=5%)	29,400	30,208	19,685	872	594	76,829	870	76,235	50,881
TEMPERATURE °F	70	590	500	1850	1850	1850		600	120
PRESSURE psia	14.7	420	415	34.7	34.7	355		345	295
OHIO NO. 6 COAL/AIR									
FLOW, lb/hr (MOISTURE=5%)	30,250	9,653	96,106	1465	632	133,912	907	133,280	123,098
TEMPERATURE °F	70	590	500	1850	1750	1850		500	120
PRESSURE psia	14.7	420	415	34,7		355		345	295

Figure D.3 Air-Blown Case – Heat Recovery and Gas Clean Up Process Flow Sheet



Heat Recovery and Gas Clean Up *Air Blown Case*	SIZE	DATE	DWG NO	REV
DRAWN SHUSTER		6/30/04	FT40342-AT3.2-001	
ISSUED	SCALE		SHEET 1 OF 3	

Figure D.4 Air-Blown Case – Mercury and Acid Gas Removal Process Flow Sheet

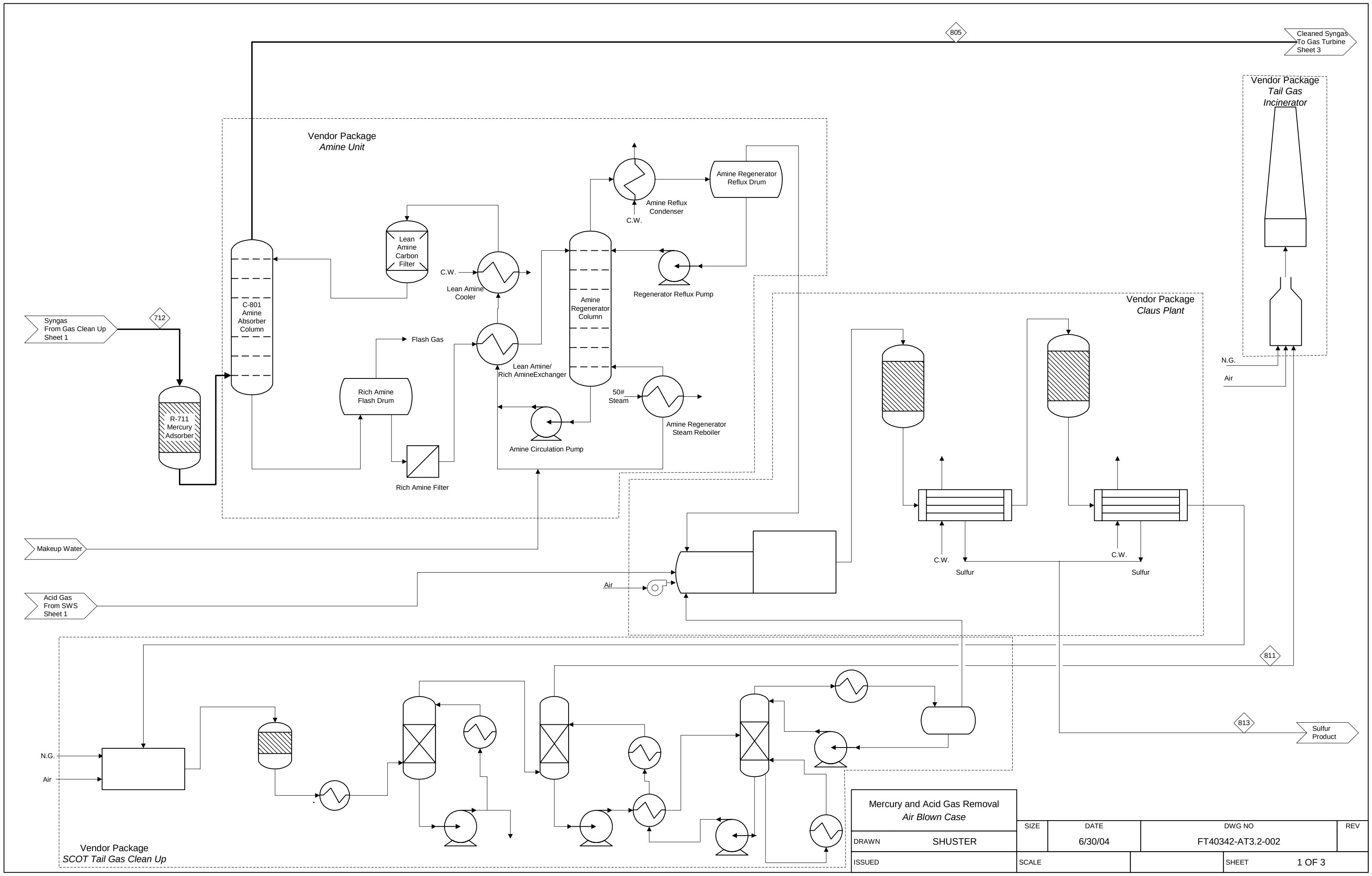


Figure D.5 Air-Blown Case – Gas Turbine & Gas Recovery Steam Generation Process Flow Sheet

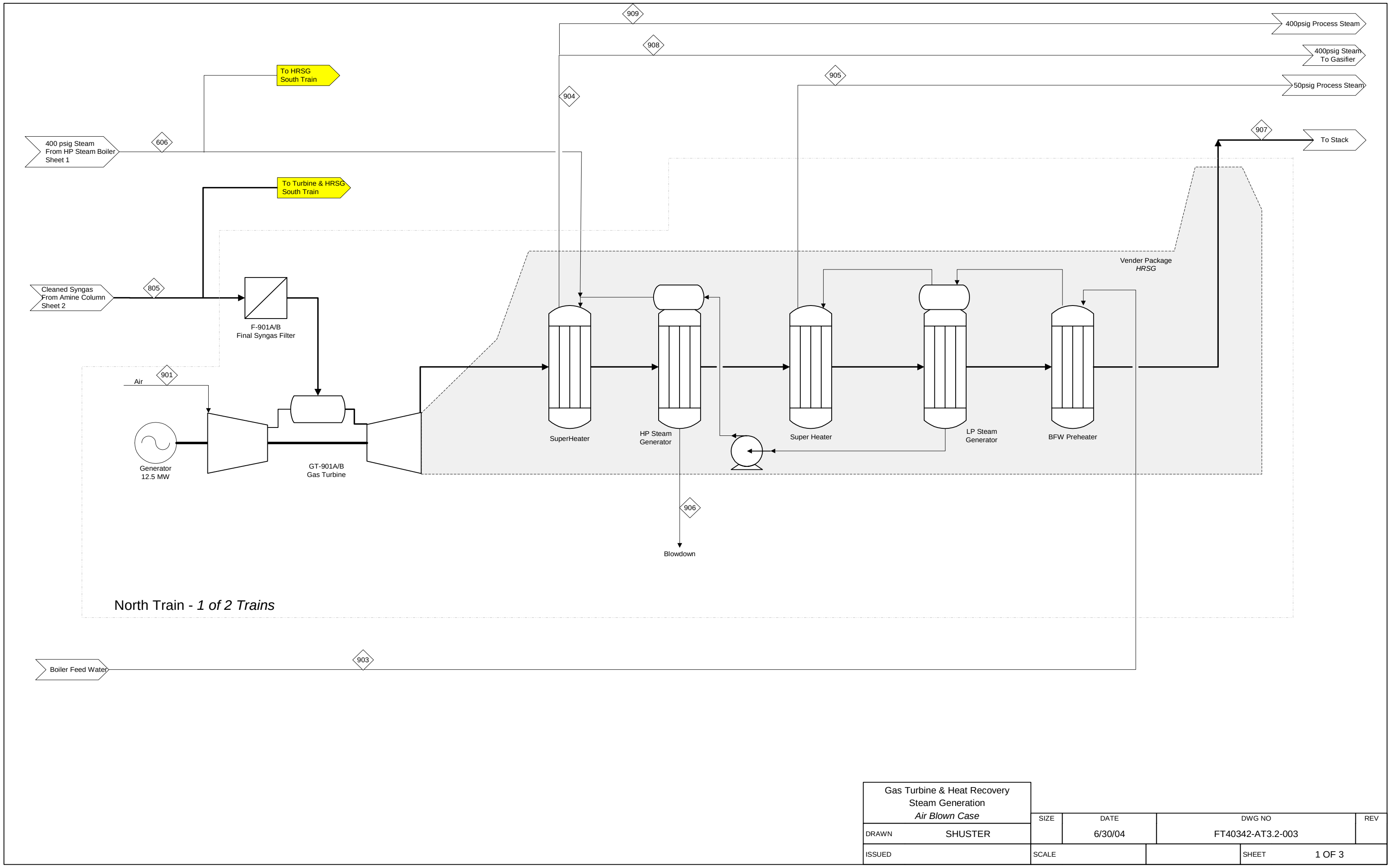


Table D.1 Air-Blown Case – Gasifier Island Material and Energy Balance

Stream No.		**1**	**2**	**3**	**4**	**6**
Stream Description		**Coal**	**Steam**	**Oxidant**	**Bottom Ash**	**Fly Ash**
Stream Composition, lb/h						
	CO					
	CO2					
	H2					
	H2O	1,513	9,653	906		
	CH4					
	H2S					
	COS					
	NH3					
	HCN					
	N2			73,335		
	O2			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 D.2 Air-Blown Case – Gas Cooling & Cleaning Material and Energy Balance

	600	601	602	603	606	701	702	704	706	707	711	712	713	805	811	813
	Raw Syngas	Cooled Syngas	From SWS	Makeup Water	Saturated Steam	Syngas to COS	Sour Water to SWS	Syngas Dwnstrm COS	Inlet BFW	BFW to Steam Drum	Condensate	Syngas To Amine	Condensate To Scubber	Syngas To GT	Vent Gas	Sulfur
Temperature F	1750	600	110	80	451	265	265	275	151	250	110	110	110	120	500	
Pressure psia	355	345	435	415	430	340	340	320	450	440	300	300	300	295	15	
Vapor Frac	1.00	1.00	0.00	0.00	1.00	1.00	0.00	1.00	0	0.00	0.00	1.00	0.00	1.00	1.00	
Mass Flow lb/hr	133,280	133,280	24,000	15,354	50,048	139,519	39,355	139,519	51,622	51,622	12,481	127,038	6,241	123,098		907
Volume Flow cuft/hr	360,362	178,082	459	290	55,621	129,865	822	140,010	1,004	1,054	404	101,918	202	103,727		
Enthalpy MMBtu/hr	-85.22	-137.65	-163.81	-105.30	-282.06	-186.88	-261.30	-186.42	-350	-344.50	-83.04	-123.12	-41.52	-111.18		
Density lb/cuft	0.37	0.75	52.31	52.92	0.90	1.07	47.86	1.00	51	48.99	30.89	1.25	30.89	1.19		
Mass Flow lb/hr																
CO	31158	31158	0	0	0	31157	1	31157	0	0	0	31157	0	31157	0	0
CO2	16221	16221	0	0	0	16324	90	16354	0	0	385	15969	192	12836	4269	0
H2	1309	1309	0	0	0	1309	0	1309	0	0	0	1309	0	1309	0	0
H2O	6183	6183	24000	15354	50048	12304	39185	12291	51622	51622	11904	387	5952	517	1238	0
CH4	3650	3650	0	0	0	3650	0	3650	0	0	0	3650	0	3632	0	0
H2S	943	943	0	0	0	939	26	962	0	0	44	918	22	0	0	0
COS	42	42	0	0	0	42	0	0	0	0	0	0	0	0	0	0
H3N	135	135	0	0	0	157	52	157	0	0	148	9	74	9	0	0
CHN	17	17	0	0	0	17	0	17	0	0	0	17	0	17	0	0
N2	73623	73623	0	0	0	73622	1	73622	0	0	0	73622	0	73622	6527	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	451	0
Mole Flow lbmol/hr	5365.8	5365.8	1332.2	852.3	2778.1	5709.0	2181.1	5709.0	2865.5	2865.5	679.5	5029.5	339.8	4937.4	412.8	28.4
CO	1112	1112	0	0	0	1112	0	1112	0	0	0	1112	0	1112	0	0
CO2	369	369	0	0	0	371	2	372	0	0	9	363	4	292	97	0
H2	649	649	0	0	0	649	0	649	0	0	0	649	0	649	0	0
H2O	343	343	1332	852	2778	683	2175	682	2865	2865	661	21	330	29	69	0
CH4	228	228	0	0	0	227	0	227	0	0	0	227	0	226	0	0
H2S	28	28	0	0	0	28	1	28	0	0	1	27	1	0	0	0
COS	1	1	0	0	0	1	0	0	0	0	0	0	0	0	0	0
H3N	8	8	0	0	0	9	3	9	0	0	9	1	4	1	0	0
CHN	1	1	0	0	0	1	0	1	0	0	0	1	0	1	0	0
N2	2628	2628	0	0	0	2628	0	2628	0	0	0	2628	0	2628	233	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	14	0
Mass Flow lb/hr																0
ASH	632	632	0	0	0	0	632	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4.45	0
Total	133912	133912	24000	15354	50048	139519	39987	139519	51622	51622	12481	127038	6241	123098	12488	907

Table D.3 Air-Blown Case – Sour Water Stripper Material and Energy Balance

	702	714	822	823	824	825	826	827	828	829	830	831	832	833	834	835	836	837	838	839	840	841
Description	Sour water from water wash	Process condensate	Overhead from flash	Sour water to cooler	Cooled water to settling tank	Vent to Flare	Settling tank to Pump	Pump to day tank	Vent to flare	Day tank to pump	Pump to stripper preheater	To sour water stripper	Overhead from stripper column	Mixed vapor stream to sulfur plant	Stripped water from stripper column	Stripped water to stripper preheater	Stripped water to air fin cooler	Stripped water to water cooler	Stripped water from cooling train	Recycle water to water wash	Purge stream to water treatment	Liquid Reflux
Temperature F	265	110	240	240	186		186	186		186	186	250	186	210	267	267	201	140	110	110	110	186
Pressure psia	340	300	38	38	15	15	15	35	15	15	43	38	35	30	40	70	65	60	55	55	55	60
Vapor Frac	0.00	0.00	1.00	0.00	0.00		0.00	0.00		0.00	0.00	0.00	1.00	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole Flow lbmol/hr	2,181.1	339.8	12.0	2,508.8	2,508.8	0.0	2,508.8	2,508.8	0.0	2,508.8	2,508.8	2,508.8	13.6	25.6	2,495.2	2,495.2	2,495.2	2,495.2	2,495.2	1,332.2	1,163.0	492.6
Mass Flow lb/hr	39,355	6,241	313	45,283	45,283	0	45,283	45,283	0	45,283	45,283	45,283	331	643	44,952	44,952	44,952	44,952	44,952	24,000	20,952	9,927
Volume Flow cuft/hr	675	99	2343	768	750	0	750	749	0	750	749	1830	2661	6073	771	771	748	732	727	388	339	171
Enthalpy MMBtu/hr	-260.54	-41.44	-1.41	-300.58	-303.05		-303.05	-303.05		-303.05	-303.05	-300.03	-0.89	-2.30	-298.22	-298.21	-301.22	-303.97	-305.32	-163.01	-142.31	-53.24
Liquid Density lb/cuft	58.31	62.86		58.93	60.43		60.43	60.43		60.43	60.43	58.62			58.31	58.31	60.12	61.37	61.87	61.87	61.87	57.93
Mass Flow lb/hr																						
CO	0.5	0.0	0.4	0.1	0.1	0.0	0.1	0.1	0.0	0.1	0.1	0.1	0.1	0.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	89.8	192.3	153.9	128.2	128.2	0.0	128.2	128.2	0.0	128.2	128.2	128.2	128.2	282.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1533.7
H2	0.1	0.0	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	39185.3	5952.2	142.8	44994.7	44994.7	0.0	44994.7	44994.7	0.0	44994.7	44994.7	44994.7	44.6	187.4	44950.1	44950.1	44950.1	44950.1	44950.1	23998.9	20951.2	6405.5
CH4	0.3	0.0	0.1	0.2	0.2	0.0	0.2	0.2	0.0	0.2	0.2	0.2	0.2	0.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1
H2S	26.0	22.1	11.7	36.4	36.4	0.0	36.4	36.4	0.0	36.4	36.4	36.4	36.4	48.1	0.1	0.1	0.1	0.1	0.1	0.0	0.0	493.6
COS	0.1	0.0	0.0	0.1	0.1	0.0	0.1	0.1	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	0.1
H3N	51.8	73.9	2.4	123.3	123.3	0.0	123.3	123.3	0.0	123.3	123.3	123.3	121.3	123.7	2.0	2.0	2.0	2.0	2.0	1.1	0.9	1493.4
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	0.0	0.0	0.0	0.0	0.0	0.0	0.2
N2	1.2	0.0	1.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.2	0.0	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	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	2.0	4.4	3.5	2.9	2.9	0.0	2.9	2.9	0.0	2.9	2.9	2.9	2.9	6.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	34.8
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	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	2175.1	330.4	7.9	2497.6	2497.6	0.0	2497.6	2497.6	0.0	2497.6	2497.6	2497.6	2.5	10.4	2495.1	2495.1	2495.1	2495.1	2495.1	1332.1	1163.0	355.6
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	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.8	0.6	0.3	1.1	1.1	0.0	1.1	1.1	0.0	1.1	1.1	1.1	1.1	1.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	14.5
COS	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H3N	3.0	4.3	0.1	7.2	7.2	0.0	7.2	7.2	0.0	7.2	7.2	7.2	7.1	7.3	0.1	0.1	0.1	0.1	0.1	0.1	0.1	87.7
CHN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	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	0.0	0.0	0.0	0.0	0.0	0.0	0

Table D.4 Air-Blown Case – GT/HRSG Material and Energy Balance

	901	903	904	905	906	907	908	909
Description	Air	BFW	400 PSIG Super Heated Steam	50 PSIG Super Heated Steam	400 PSIG Evaporator Blowdown	Stack Exhaust	400 PSIG Super Heated Steam - Internal Use (Gasifier)	400 PSIG Super Heated Steam - Export
Temperature F	65	150	548	353	449	240	548	548
Pressure psia	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
Mole Flow lbmol/hr	24,843.2	6,786.2	7,256.4	2,128.4	179.1	28,733.1	2,218.1	5,038.3
Mass Flow lb/hr	711,396.0	122,256.0	130,726.0	38,344.0	3,227.2	834,494.0	39,960.0	90,766.0
Volume Flow cuft/hr	9,486,220.0	2,376.4	173,660.8	258,547.3	77.2	14,675,060.0	53,084.2	120,576.6
Enthalpy MMBtu/hr	-55.5	-829.3	-729.8	-216.7	-20.8	-393.3	-223.1	-506.7
Liquid Phase Density lb/cuft	53.18	51.45			41.80			
Mass Flow lb/hr								
H2O	8869.9	122256.0	130726.0	38344.0	3227.2	25914.9	39960.0	90766.0
N2	548553.1	0.0	0.0	0.0	0.0	606654.4	0.0	0.0
CO2	231.8	0.0	0.0	0.0	0.0	72122.3	0.0	0.0
SO2	0.0	0.0	0.0	0.0	0.0	1.3	0.0	0.0
AR	6554.5	0.0	0.0	0.0	0.0	9082.5	0.0	0.0
O2	147186.7	0.0	0.0	0.0	0.0	120718.6	0.0	0.0
Mole Flow lbmol/hr								
H2O	492.4	6786.2	7256.4	2128.4	179.1	1438.5	2218.1	5038.3
N2	19581.8	0.0	0.0	0.0	0.0	21655.8	0.0	0.0
CO2	5.3	0.0	0.0	0.0	0.0	1638.8	0.0	0.0
SO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
AR	164.1	0.0	0.0	0.0	0.0	227.4	0.0	0.0
O2	4599.8	0.0	0.0	0.0	0.0	3772.6	0.0	0.0

Figure D.6 Oxygen-Blown Case – Heat Recovery and Gas Clean Up Process Flow Sheet

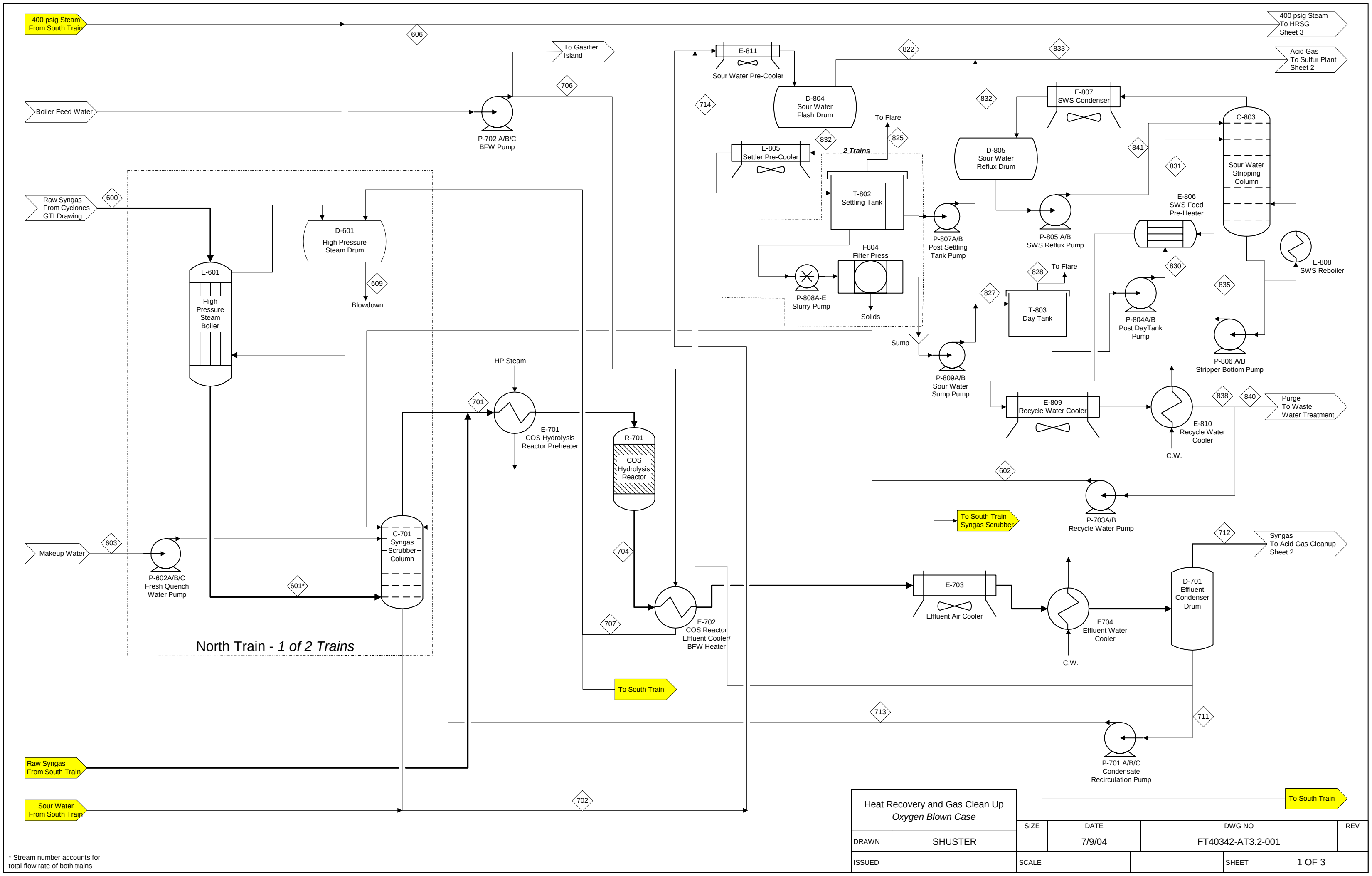


Figure D.7 Oxygen-Blown Case – Mercury and Acid Gas Removal Process Flow Sheet

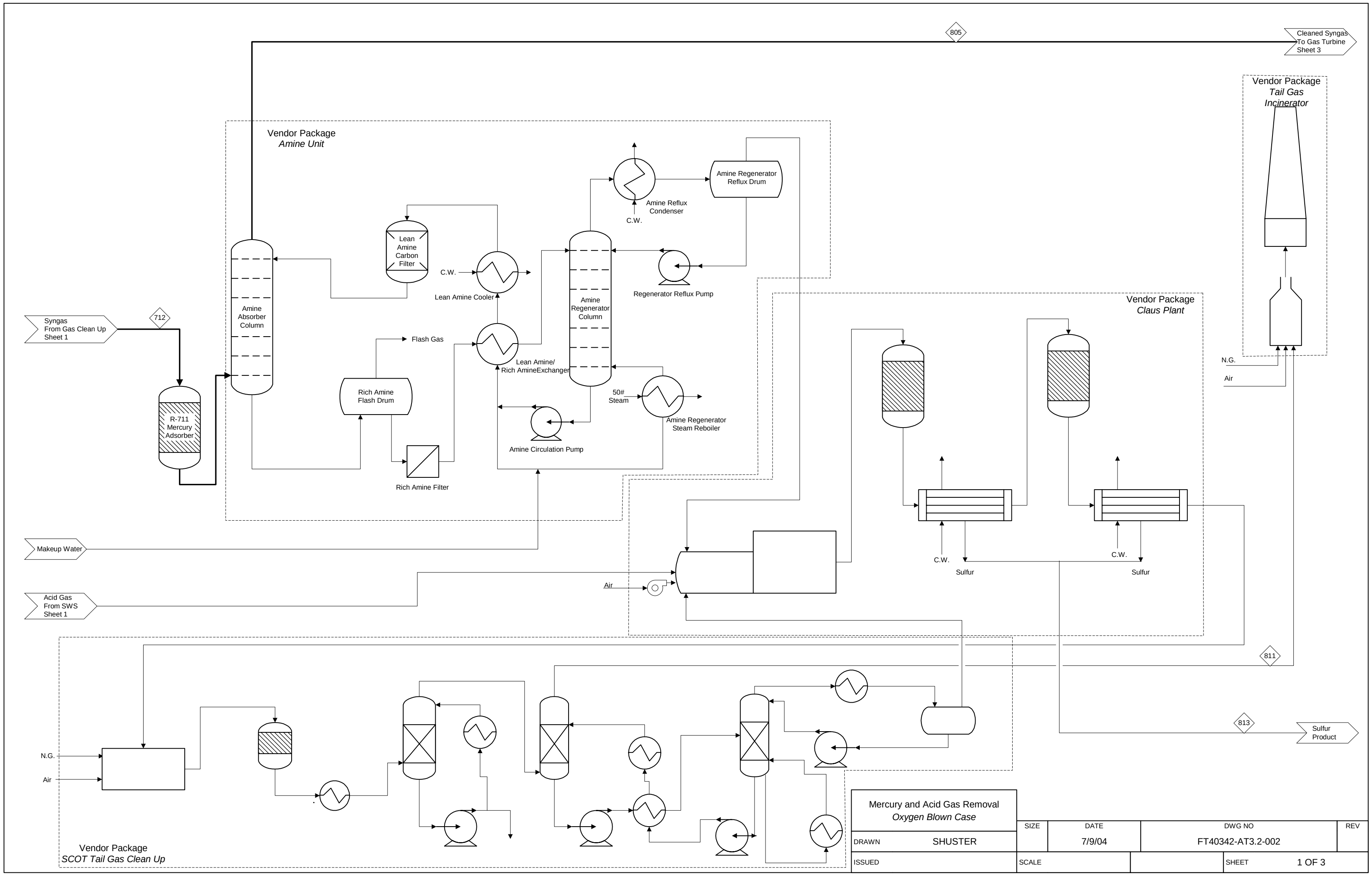


Figure D.8 Oxygen-Blown Case – Gas Turbine & Gas Recovery Steam Generation Process Flow Sheet

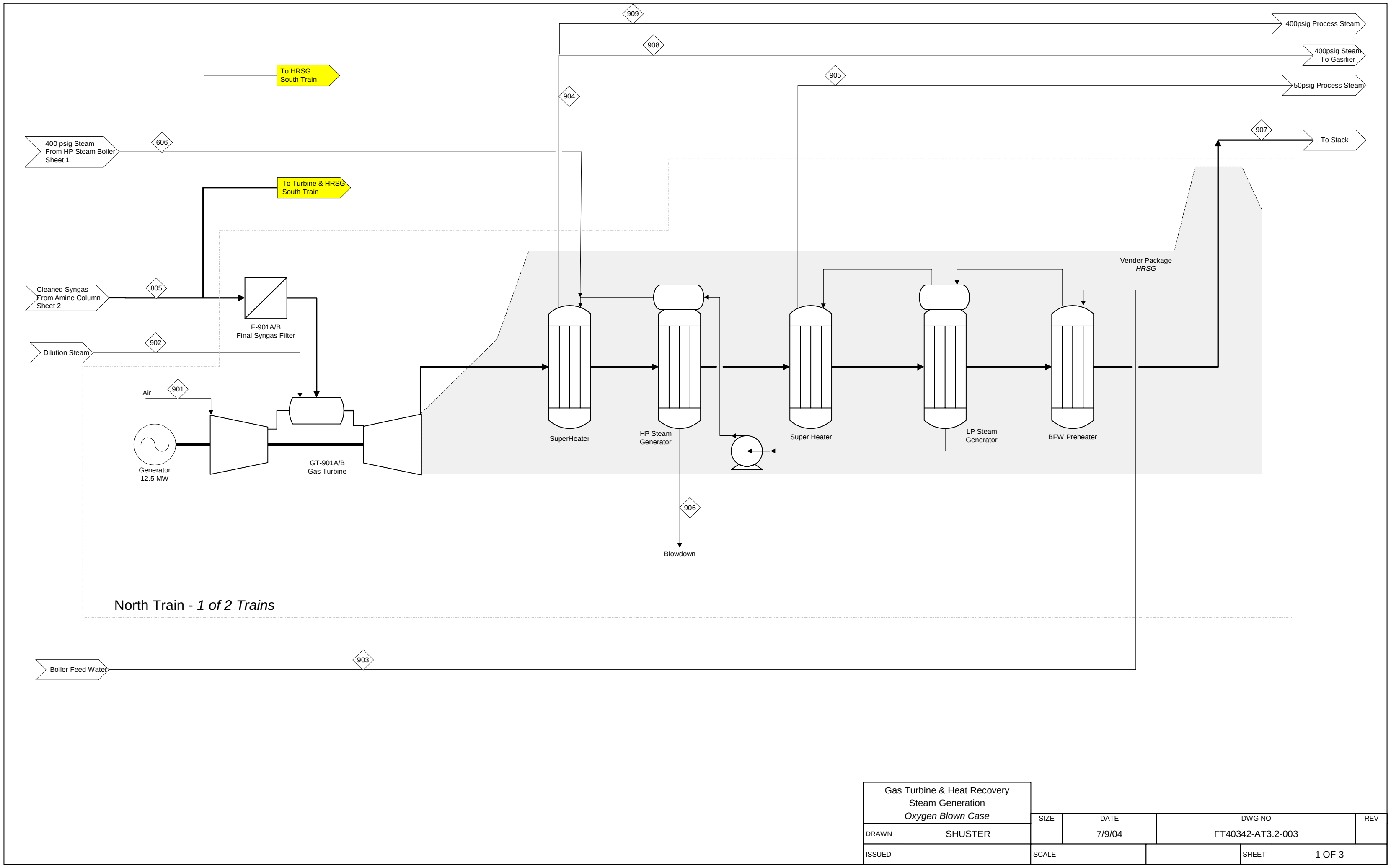


Table D.5 Oxygen-Blown Case – Gasifier Island Material and Energy Balance

Stream No.	1	2	3	4	6
Stream Description	**Coal**	**Steam**	**Oxidant**	**Bottom Ash**	**Fly Ash**
Stream Composition, lb/h					
CO					
CO2					
H2					
H2O	1,420	30,208	168		
CH4					
H2S					
COS					
NH3					
HCN					
N2			860		
O2			18,657		
Coal/residue [1]	25,385			15	225
Mineral Matter/Ash	1,595			857	369
Total, lb/h	**28,400**	**30,208**	**19,685**	**872**	**594**
Temperature, F	70	550	500	1850	1850
Pressure, psia	14.7	420	415	14.7	14.7

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

Table D.6 Oxygen Blown Case – Gas Cooling & Cleaning Material and Energy Balance

	600	601	602	603	606	701	702	704	706	707	711	712	713	805	811	813
	Raw Syngas	Cooled Syngas	From SWS	Makeup Water	Saturated Steam	Syngas to COS	Sour Water to SWS	Syngas Dwnstrm COS	Inlet BFW	BFW to Steam Drum	Condensate	Syngas To Amine	Condensate To Scubber	Syngas To GT	Vent Gas	Sulfur
Temperature F	1750	600	110	80	451	265	265	275	151	250	110	110	110	120	500	
Pressure psia	355	345	435	415	430	340	340	320	450	440	300	300	300	295	15	
Vapor Frac	1.00	1.00	0.00	0.00	1.00	1.00	0.00	1.00	0.00	0.00	0.01	1.00	0.01	1.00	1.00	
Mass Flow lb/hr	76,235	76,235	90,000	35,111	39,179	63,140	141,532	63,140	40,200	40,200	6,652	56,488	3,326	50,881	17,492	870
Volume Flow cuft/hr	254,863	124,779	1,720	664	43,542	69,783	2,991	75,303	782	821	183	54,760	91	54,097		
Enthalpy MMBtu/hr	-199.47	-240.54	-614.29	-240.80	-220.81	-177.87	-940.12	-177.59	-272.60	-268.28	-44.71	-143.73	-22.36	-125.41		
Density lb/cuft	0.30	0.61	52.31	52.92	0.90	0.90	47.31	0.84	51.44	48.99	36.38	1.03	36.38	0.94		
Mass Flow lb/hr																
CO	24109	24109	0	0	0	24107	3	24107	0	0	0	24107	0	24107	0	0
CO2	25201	25201	0	0	0	24869	394	24898	0	0	125	24773	63	19912	5982	0
H2	2036	2036	0	0	0	2035	1	2035	0	0	0	2035	0	2035	0	0
H2O	19313	19313	90000	35111	39179	6693	140966	6682	40200	40200	6471	211	3235	341	1734	0
CH4	3337	3337	0	0	0	3335	2	3335	0	0	0	3335	0	3319	0	0
H2S	905	905	0	0	0	865	54	887	0	0	27	859	14	0	0	0
COS	40	40	0	0	0	39	0	0	0	0	0	0	0	0	0	0
H3N	126	126	0	0	0	29	112	29	0	0	28	0	14	0	0	0
CHN	16	16	0	0	0	16	0	16	0	0	0	16	0	16	0	0
N2	1152	1152	0	0	0	1152	0	1152	0	0	0	1152	0	1152	9145	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	631	
Mole Flow lbmol/hr	3799.7	3799.7	4995.8	1949.0	2174.8	3084.1	7842.5	3084.1	2231.4	2231.4	364.5	2719.7	182.2	2590.2	578.4	27.2
CO	861	861	0	0	0	861	0	861	0	0	0	861	0	861	0	0
CO2	573	573	0	0	0	565	9	566	0	0	3	563	1	452	136	0
H2	1010	1010	0	0	0	1010	0	1010	0	0	0	1010	0	1010	0	0
H2O	1072	1072	4996	1949	2175	372	7825	371	2231	2231	359	12	180	19	96	0
CH4	208	208	0	0	0	208	0	208	0	0	0	208	0	207	0	0
H2S	27	27	0	0	0	25	2	26	0	0	1	25	0	0	0	0
COS	1	1	0	0	0	1	0	0	0	0	0	0	0	0	0	0
H3N	7	7	0	0	0	2	7	2	0	0	2	0	1	0	0	0
CHN	1	1	0	0	0	1	0	1	0	0	0	1	0	1	0	0
N2	41	41	0	0	0	41	0	41	0	0	0	41	0	41	326	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	19.73323	0
Mass Flow lb/hr																0
ASH	594	594	0	0	0	0	594	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	6.238876	0
Total	76829	76829	90000	35111	39179	63140	142125	63140	40200	40200	6652	56488	3326	50881	17499	870

Table D.7 Oxygen-Blown Case – Sour Water Stripper Material and Energy Balance

	702	714	820	821	822	823	824	825	826	827	828
Description	Sour water from water wash	Process condensate	Mixed input	Cooled input	Overhead from flash	Sour water to cooler	Cooled water to settling tank	Vent to Flare	Settling tank to Pump	Pump to day tank	Vent to flare
Temperature F	265	110	262	171	171	171	161		161	161	
Pressure psi	340	300	300	15	15	15	15	15	15	35	15
Vapor Frac	0.00	0.00	0.00	0.00	1.00	0.00	0.00		0.00	0.00	
Mole Flow lbmol/hr	7,842.6	182.2	8,024.9	8,024.9	8.8	8,016.1	8,016.1	0.0	8,016.1	8,016.1	0.0
Mass Flow lb/hr	141,532	3,326	144,858	144,858	264	144,594	144,594	0	144,594	144,594	0
Volume Flow cuft/hr	2426	54	2480	6398	4019	2379	2371	0	2371	2371	0
Enthalpy MMBtu/hr	-937.35	-22.27	-959.62	-972.83	-1.07	-971.76	-973.20		-973.20	-973.20	
Liquid Phase Density lb/cuft	58.31	61.93	58.43	60.80		60.80	60.99		60.99	60.99	
Mass Flow lb/hr											
CO	3.0	0.0	3.0	3.0	1.8	1.2	1.2	0.0	1.2	1.2	0.0
CO2	394.0	63.0	457.0	457.0	179.2	277.8	277.8	0.0	277.8	277.8	0.0
H2	1.0	0.0	1.0	1.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	140966.0	3235.0	144201.0	144201.0	66.2	144134.8	144134.8	0.0	144134.8	144134.8	0.0
CH4	2.0	0.0	2.0	2.0	0.3	1.7	1.7	0.0	1.7	1.7	0.0
H2S	54.0	14.0	68.0	68.0	15.2	52.8	52.8	0.0	52.8	52.8	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
H3N	112.0	14.0	126.0	126.0	0.1	125.9	125.9	0.0	125.9	125.9	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
N2	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	0.1	0.0	0.1	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0
CO2	9.0	1.4	10.4	10.4	4.1	6.3	6.3	0.0	6.3	6.3	0.0
H2	0.5	0.0	0.5	0.5	0.5	0.0	0.0	0.0	0.0	0.0	0.0
H2O	7824.8	179.6	8004.4	8004.4	3.7	8000.7	8000.7	0.0	8000.7	8000.7	0.0
CH4	0.1	0.0	0.1	0.1	0.0	0.1	0.1	0.0	0.1	0.1	0.0
H2S	1.6	0.4	2.0	2.0	0.4	1.5	1.5	0.0	1.5	1.5	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
H3N	6.6	0.8	7.4	7.4	0.0	7.4	7.4	0.0	7.4	7.4	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
N2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

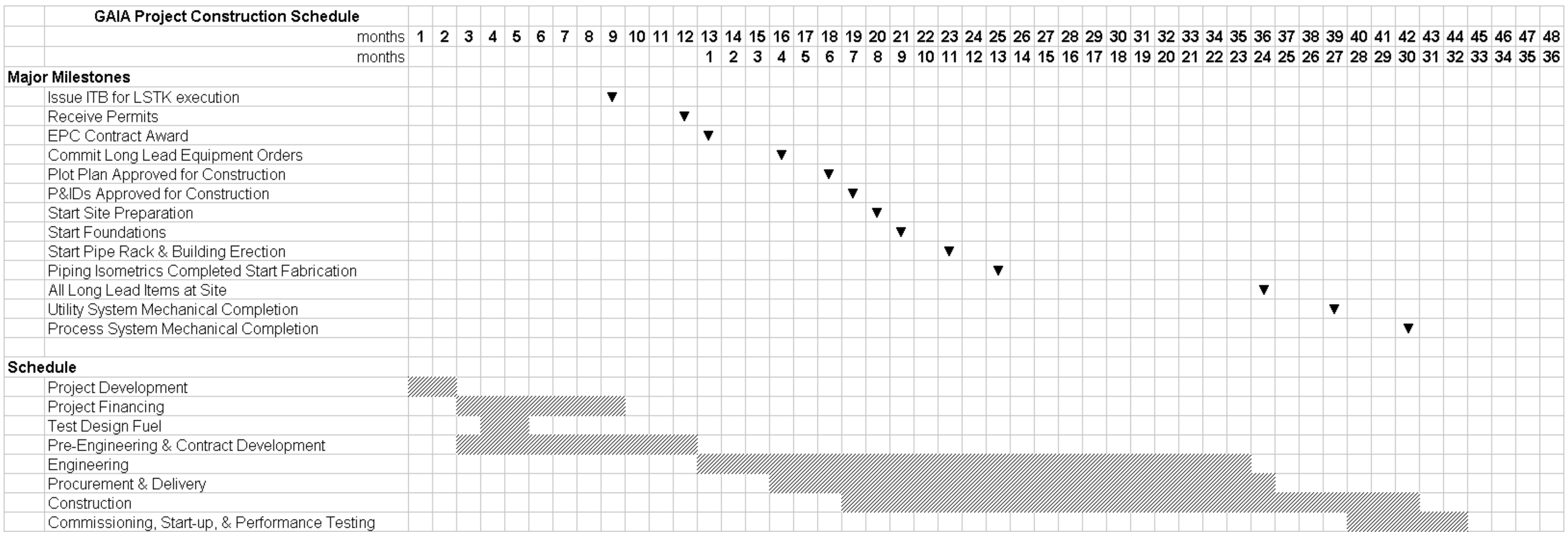
	829	830	831	832	833	834	835	836	837	838	839	840	841
Description	Day tank to pump	Pump to stripper preheater	To sour water stripper	Overhead from stripper column	Mixed vapor stream to sulfur plant	Stripped water from stripper column	Stripped water to stripper preheater	Stripped water to air fin cooler	Stripped water to water cooler	Stripped water from cooling train	Recycle water to water wash	Purge stream to water treatment plant	Liquid Reflux
Temperature F	161	161	250	207	197	267	267	176	140	110	110	110	207
Pressure psi	15	38	38	35	30	40	70	70	65	60	60	60	60
Vapor Frac	0.00	0.00	0.00	1.00	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole Flow lbmol/hr	8,016.1	8,016.1	8,016.1	22.8	31.6	7,993.3	7,993.3	7,993.3	7,993.3	7,993.3	4,995.8	2,997.5	948.1
Mass Flow lb/hr	144,594	144,594	144,594	594	857	144,001	144,001	144,001	144,001	144,001	90,000	54,001	17,808
Volume Flow cuft/hr	2371	2371	5644	4617	7356	2470	2470	2374	2346	2327	1455	873	310
Enthalpy MMBtu/hr	-973.20	-973.18	-959.93	-2.01	-3.08	-955.31	-955.29	-968.55	-973.75	-978.06	-611.29	-366.78	-107.94
Liquid Phase Density lb/cuft	60.99	60.99	58.74			58.31	58.31	60.68	61.37	61.87	61.87	61.87	57.50
Mass Flow lb/hr													
CO	1.2	1.2	1.2	1.2	3.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	277.8	277.8	277.8	277.8	457.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1065.8
H2	0.0	0.0	0.0	0.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	144134.8	144134.8	144134.8	140.3	206.5	143994.5	143994.5	143994.5	143994.5	143994.5	89996.2	53998.3	14951.6
CH4	1.7	1.7	1.7	1.7	2.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.7
H2S	52.8	52.8	52.8	52.6	67.8	0.2	0.2	0.2	0.2	0.2	0.1	0.1	383.2
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
H3N	125.9	125.9	125.9	119.9	120.0	6.0	6.0	6.0	6.0	6.0	3.8	2.3	1406.8
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
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
Mole Flow lbmol/hr													
CO	0.0	0.0	0.0	0.0	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO2	6.3	6.3	6.3	6.3	10.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	24.2
H2	0.0	0.0	0.0	0.0	0.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	8000.7	8000.7	8000.7	7.8	11.5	7992.9	7992.9	7992.9	7992.9	7992.9	4995.5	2997.4	829.9
CH4	0.1	0.1	0.1	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2S	1.5	1.5	1.5	1.5	2.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	11.2
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
H3N	7.4	7.4	7.4	7.0	7.0	0.4	0.4	0.4	0.352	0.352	0.22	0.132	82.605
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
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

Table D.8 Oxygen-Blown Case – GT/HRSG Material and Energy Balance

	901	902	903	904	905	906	907	908	909
Description	Air	Dilution Nitrogen	BFW	400 PSIG Super Heated Steam	50 PSIG Super Heated Steam	400 PSIG Evaporator Blowdown	Stack Exhaust	400 PSIG Super Heated Steam - Internal Use (Gasifier)	400 PSIG Super Heated Steam - Export
Temperature F	62	120	150	548	353	449	240	548	548
Pressure psia	14.7	295.0	80.0	415.0	70.0	420.0	14.7	415.0	415.0
Vapor Frac	1.00	1.00	0.00	1.00	1.00	0.00	1.00	1.00	1.00
Mole Flow lbmol/hr	25,021.2	2,220.8	6,847.1	6,103.0	2,761.7	157.1	28,736.9	2,991.9	3,111.1
Mass Flow lb/hr	717,022.0	62,211.0	123,352.0	109,948.0	49,752.0	2,830.8	830,136.0	53,900.0	56,048.0
Volume Flow cuft/hr	9,501,160.0	46,807.2	2,397.7	146,058.6	335,469.6	67.7	14,676,540.0	71,602.6	74,456.0
Enthalpy MMBtu/hr	-51.2	0.5	-836.8	-613.8	-281.2	-18.2	-404.2	-300.9	-312.9
Liquid Phase Density lb/cuft	53.24		51.45			41.80			
Mass Flow lb/hr									
H2O	8046.8	0.0	123352.0	109948.0	49752.0	2830.8	31018.5	53900.0	56048.0
N2	553627.2	62211.0	0.0	0.0	0.0	0.0	601115.0	0.0	0.0
CO2	234.0	0.0	0.0	0.0	0.0	0.0	67274.0	0.0	0.0
SO2	0.0	0.0	0.0	0.0	0.0	0.0	1.3	0.0	0.0
AR	6614.9	0.0	0.0	0.0	0.0	0.0	9161.9	0.0	0.0
O2	148499.1	0.0	0.0	0.0	0.0	0.0	121565.3	0.0	0.0
Mole Flow lbmol/hr									
H2O	446.7	0.0	6847.1	6103.0	2761.7	157.1	1721.8	2991.9	3111.1
N2	19762.9	2220.8	0.0	0.0	0.0	0.0	21458.1	0.0	0.0
CO2	5.3	0.0	0.0	0.0	0.0	0.0	1528.6	0.0	0.0
SO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
AR	165.6	0.0	0.0	0.0	0.0	0.0	229.3	0.0	0.0
O2	4640.8	0.0	0.0	0.0	0.0	0.0	3799.1	0.0	0.0

Figure E-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 E.1 Project Construction Schedule



F.1 SUMMARY

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 both the Subtask 3.2 IGCC co-production (export power and steam) plant designs to account for forced and scheduled outages to determine expected annual revenue and expense cash flows. Based on these cash flows, financial analyses were performed to evaluate the comparative economics of the various plant configurations and alternate design options.

The effect of sparing (back-up equipment or parallel trains of reduced capacity) can have a significant affect on the capacity factor (availability) of a plant depending upon the amount of spare equipment or parallel trains that are present. Sparing is most effective in increasing the overall plant availability when those portions of the plant with the lowest on-stream factors are replicated. Because operability is key to the Subtask 3.2 design, sparing played an important role in the design development to provide optimum on-stream capacity while also attempting to maintain economic viability. Availability analysis for the Subtask 3.2 air-blown and oxygen-blown cases result in annual average on-stream factors of 85.67% and 82.65% respectively (including scheduled maintenance). The availability of the air separation unit is the reason for the lower availability of the oxygen-blown case.

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

Attachment A, Availability Nomenclature, contains definitions of availability related terms as proposed by the Gasification Technology Council. This table is supplemented with additional terms as used in this study.

F.2 AVAILABILITY ANALYSIS

Common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period all are dependent on the project cash flow. The net cash flow is the sum of all project revenues and expenses. Depending upon the detail of the financial analysis, the cash flow streams usually are computed on 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 that influence it include scheduled maintenance, forced outages, equipment reliability, and redundancy. This addendum describes the results of the availability analyses that were conducted to calculate the annual average production rates (capacity factors) for the air-blown and oxygen-blown cases. The calculations are based on the availability data for the individual plant sections (as shown in Table F.1) that were observed at the Wabash River Repowering Project during the demonstration period and for estimates developed for coal preparation and handling, gasifier island, and mercury control. This information was then used for the basis of the calculations for the specific plant configurations. Based on published mathematical formulae that account for parallel trains, spare equipment, equipment reliability, and scheduled maintenance, average annual production rates were calculated. Thus, these calculations allowed the effects of various train and equipment sparing configurations on the annual production rates to be examined.

In a subsequent financial analysis, these production rates were then used as the basis for calculating the annual revenue streams. These financial analyses and their results are discussed in the main portion of this report.

Attachment A, Availability Nomenclature, contains definitions of availability related terms as proposed by the Gasification Technology Council. This table is supplemented with additional terms as used in this study.

F.3 AVAILABILITY ANALYSIS BASIS

F.3.1 Gasifier Block

The following analysis is developed based on the assumption that the equipment has been "proven" in design and that failure is due to normal wear and tear on the equipment. It is assumed that chronic failure of components due to improper application is not a factor. *The overall availability of each Gasifier Island train is anticipated to be 97.49% based on an unscheduled maintenance estimate of approximately 9 days per year.* The basis for this assumption is summarized herein.

F.3.1.1 Overall System

It is assumed that there will be a 3 week planned maintenance outage performed on an entire gasification train once per year. During this period, routine maintenance on pumps, valves (seals, etc.), pressurized feed components, and inspection and repair of refractory will be performed. Instrumentation will be recalibrated, safety valves reset, etc.

Planned Maintenance = 21 days per year

F.3.1.2 Gasifier Lockhopper System

The lockhopper system components include:

- T-201 Weigh Hopper
- S-201 Rotary Feeder
- D-202 Lockhopper
- D-203 Surgehopper
- S-203 Feed Metering Screw
- Lockhopper Valves (designed for solid applications with diaphragm seals)
- Gas Lockhopper Valves (designed for feed and discharge of inert gas to lockhopper)
- Instrumentation (differential pressure transmitters, purge rotometers, etc.)

The fuel feed lockhopper system will function reliably if the fuel is delivered within specification with minimal surface moisture. The presence of moisture can cause the coal to “stick” to itself and the walls of the lockhopper system. This analysis assumes that the fuel preparation system (area 100) is functioning normally.

The fuel feed system is redundant. Two 100% capacity feed lockhopper systems are provided to ensure an uninterrupted flow of fuel to the gasifier. This should ensure the feed to each gasifier has a reliability of greater than 99%.

Areas for unexpected “forced” maintenance on a feed system train failure include the following items. Note that only a simultaneous failure of components on both trains would cause a gasifier outage. All this equipment can be maintained during operation:

- Failure of the diaphragm seals on a lockhopper valve. These can be replaced in about 4 hours. Failure is not assumed to be less than 3 – 4 times a year.
- Failure of the pressure seals on S-203. This could require 1 – 2 hours to repair and is assumed no more than once per year.
- Failure of the bearings on S-203. This could require 6 – 12 hours to repair and is assumed no more than once per year.

- Assorted random instrumentation failures, accounting for 2-4 hours to repair occurring up to 3-5 times/year.

Total maintenance for each lockhopper = 1 - 2½ days per year

> *Note that the coincidence of forced maintenance on both lockhoppers occurring is assumed to be 20%, thus a forced outage on the gasifier from the coal lockhopper system is estimated to be ½ - 1 days per year*

F.3.1.3 Gasifier Reactor

Improper operation of the gasifier can result in plugging of the bottom grid. This will require a complete shutdown to clean the system, make repairs, if necessary, and restart the gasifier. This should occur infrequently, but the likelihood of this occurrence is about 50% in any given year. Such an outage will require about 7 – 10 days of downtime.

Maintenance for the gasifier plugging = 3½ - 5 days per year

F.3.1.4 Startup Heater

The startup heater is a package unit that is used to preheat the gasifier after an outage prior to admitting solid fuel. The startup heater is used typically 3 – 5 times/year depending on the overall reliability of the gasifier island. The heater should be very reliable since it is used infrequently. If there are problems, they will result in a longer than anticipated startup. The only likely problem would be failure of the fireye burner detection system. This system can be assumed to be 99.95% reliable.

Maintenance for the startup heater = 1/8 days per year

Dus*t* Cyclones

Two cyclones, CY-401 and CY-402, are used to recycle dust back to the gasifier to improve carbon conversion. Abnormal operation can cause the cyclones to lose their seal legs, resulting in aberrant conditions in the gasifier. This does not require maintenance, but can result in poor gasifier performance that could force diversion of gas to the flare. This could happen once or twice per year and result in off-spec gas for a period of 1 – 2 hours. There is about a 15% possibility that a “bridge” can form once per year in the cyclones or related piping. Should this occur, a 6 - 10 day outage will be required to clean and restart the gasifier.

Maintenance for off-spec syngas = 1/8 days per year

Maintenance for the cyclone bridge = ½ - 1½ days per year

The dust from cyclone CY-403 is removed from the process by a lockhopper system consisting of the following equipment:

- D-401 Surge Hopper
- Twin Safety Valves for equipment isolation
- S-401 Rotary Screw Feeder
- D-402 Lockhopper
- T-203 Transport hopper
- S-403 Rotary Screw
- Lockhopper Valves (designed for solid applications with diaphragm seals)
- Gas Lockhopper Valves (designed for feed and discharge of inert gas to lockhopper)
- Instrumentation (differential pressure transmitters, purge rotometers, etc.)

Areas for unexpected “forced” maintenance in the lockhopper system train could include the following items. All this equipment can be maintained without a cold shutdown and depressurization of the gasifier:

- Failure of the diaphragm seals on a lockhopper valve. These can be replaced in about 4 hours. Failure is not expected more than 3 – 4 times a year.
- Failure of the pressure seals on S-401. This could require 1 – 2 hours to repair and is assumed to occur no more than once per year.
- Failure of the bearings on S-401. This could require 6 – 12 hours to repair and is anticipated no more than once per year.
- Assorted random instrumentation failures could occur 3-5 times/year, each requiring about 2-4 hours for repair.

Maintenance for the dust lockhoppper = 1 - 2½ days per year

F.3.1.5 Ash Discharge System

Ash is removed from the gasifier by a lockhopper system consisting of the following equipment:

- D-501 Surge Hopper
- Twin Safety Valves for equipment isolation
- S-501 Rotary Screw Feeder

- D-502 Lockhopper
- T-503 Transport hopper
- S-503 Rotary Screw
- Lockhopper Valves (designed for solid applications with diaphragm seals)
- Gas Lockhopper Valves (designed for feed and discharge of inert gas to lockhopper)
- Instrumentation (differential pressure transmitters, purge rotometers, etc.)

Areas for unexpected "forced" maintenance on the lockhopper system train include the following items. All this equipment can be maintained without shutdown and depressurization of the gasifier:

- Failure of the diaphragm seals on a lockhopper valve. These can be replaced in about 4 hours. Failure is not anticipated more than 3 – 4 times a year.
- Failure of the pressure seals on S-401. This could require 1 – 2 hours to repair and is anticipated no more than once per year.
- Failure of the bearings on S-401. This could require 6 – 12 hours to repair and is anticipated no more than once per year.
- Assorted random instrumentation failures could occur 3-5 times/year, each requiring about 2-4 hours to repair.

Maintenance for the ash lockhoppper = 1 - 2½ days per year

F.3.2 Gas Clean-Up and Balance of Plant Operations

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. For this analysis, most operations of Subtask 3.2, exclusive of the gasifier island, are fundamentally similar to those of the Wabash River Repowering Project. In addition to the gasifier block, two other blocks, 1.) coal preparation and handling and 2.) mercury removal are not represented in the Wabash River final report. Availability estimates for those operations are estimated based on the conceptual design and are not based on actual operating experience. Additionally, availability estimates for the combustion turbine are based on the GE 7F advanced combustion turbine design used at Wabash River. The turbine used for this subtask, the GE 10, is not currently available for use with coal-derived syngas. Therefore, demonstrated on-stream performance is not available for a syngas application. Availability estimates used for this analysis are summarized in Table F.1.

Table F.1. Availability Estimates, Air-Blown and Oxygen-Blown Cases

Plant Section	Availability	
	Air-Blown Case	Oxygen-Blown Case
Coal Prep/Oxidant		
Coal Preparation and Handling	99.95%	99.95%
Air Compressor / ASU	99.84%	96.32%
Gasifier Block		
Gasifier Island	97.49%	97.49%
High Temperature Heat Recovery	97.96%	97.96%
Water Scrubber	99.87%	99.87%
Gas Clean-up		
COS System	100%	100%
Acid Gas Removal	99.72%	99.72%
Sulfur Recovery	99.94%	99.94%
Sour Water Treatment	100%	100%
Mercury Removal	100%	100%
Power Block		
Combustion Turbine/Generator	98.19%	98.19%
Heat Recovery Steam Generator	97.40%	97.40%

As can be seen in Table F.1, unit level availabilities range from 96.32% for the air separation unit to 100% for the COS system, sour water treatment, and mercury removal. The availability is a function of component mean time between failures (MTBF) and mean down time (MDT). For relatively simple systems (e.g., carbon beds for mercury removal) or for systems with significant redundancy or back-up storage (e.g., day tank for the sour water treatment system), availability is expected to be high. For more complex systems such as the ASU or gasifier operations, availability is expected to be slightly lower.

Based on the availability estimates in Table F.1, analyses were calculated using the EPRI recommended procedure. This procedure calculates availabilities based only on two plant states, operating at design capacity or not operating. For a single train plant with all the units in a series configuration (i.e., no redundancy), the overall plant availability simply is the product of the availability of all the individual unit availabilities. For multiple trains (or for plant sections with spare units), the EPRI report presents mathematical formulae based on a probabilistic approach for predicting the availability of all trains or combinations thereof, such as 1 of 2, 2 of 3, 1 of 3, etc. Appropriate combinations of these formulae are used to represent plants with some sections containing multiple trains or spare equipment, and other sections being single trains.

Since the objective of this availability study is to determine the projected annual revenue stream, this study does not differentiate between forced and scheduled outages. In other words, 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. Consequently, the annual availabilities reported here probably will be different than those from studies which do not consider forced and scheduled outages in such a rigorous manner.

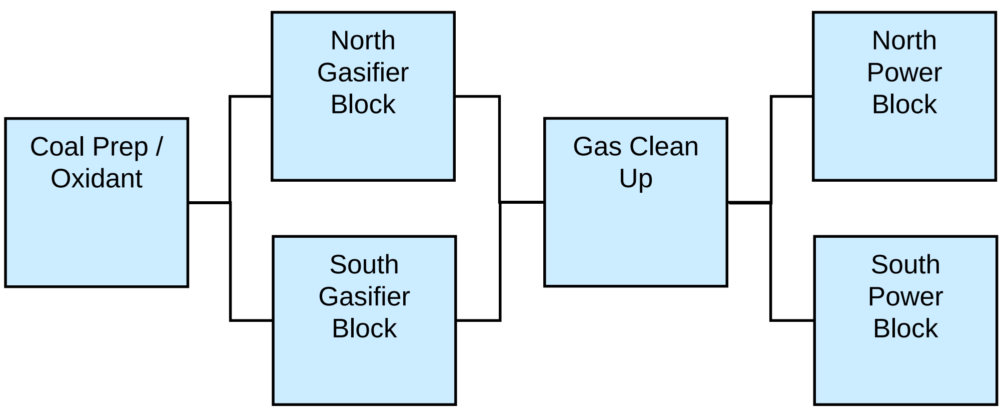
F.4 USE OF NATURAL GAS

Due to limited availability of natural gas at the site, it is not considered for use as a back-up fuel during any gasifier outage.

F.5 AVAILABILITY BLOCK DIAGRAM

Figure F.1 represents the block flow of the gasification and power block used for the availability analysis. The figure also illustrates the combination of parallel and series configuration. Make-up of the individual blocks as well as availabilities of the component units are presented in Table F.1 previously.

Figure F.1. Availability Block Diagram



F.6 AVAILABILITY CALCULATIONS

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

Table F.2 Calculated Availabilities

		Air-Blown Case	Oxygen-Blown Case
Syngas Operations[1]			
	2 of 2	90.47%	87.28%
	1 of 2	99.24%	95.74%
Power Block[2]			
	2 of 2	91.46%	91.46%
	1 of 2	99.81%	99.81%
Equivalent Availability			
	w/o Scheduled Maintenance	90.90%	87.69%
	w/ Scheduled Maintenance	85.67%	82.60%

Notes:
1. Represents coal preparation and handling through final gas cleaning and includes sulfur recovery and sour water treatment.
2. 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 product (power and steam). Three operating states were considered for this study and are presented in Table F.3.

Table F.3 Operating State Statistics

Syngas Operations[1]	Power Block[2]	Net Product Output[3]	Equivalent Availability[4]
Air Case			
2 of 2	2 of 2	100%	77.99%
	1 of 2	50%	85.10%
1 of 2[5]	1 of 2	50%	93.35%
Oxygen Case			
2 of 2	2 of 2	100%	75.24%
	1 of 2	50%	82.10%
1 of 2[5]	1 of 2	50%	99.05%

Notes:
1. Represents coal preparation and handling through final gas cleaning and includes sulfur recovery and sour water treatment.
2. Includes combustion turbine, generator, and heat recovery steam generation.
3. Represents gross power and steam output minus internal power and steam demand.
4. Includes scheduled outage of 21 days/year
5. Represents both gasifiers operating at 50% turndown.

The equivalent availabilities included in Table F.2 are a result of weighted equivalent availabilities from Table F.3. For this study, internal power and steam demands are assumed proportional to syngas generation. For periods when 1 of 2 gasifiers are available, internal power and steam demands are 50% of full load. For periods when 2 of 2 gasifiers are in-service and only 1 of 2 CT/HRSG units is available, both gasifiers are assumed operating at 50% of full load, with internal power and steam demands also at 50% of full load. The lower equivalent availability of the oxygen-blown case when compared to the air-blown case is a direct result of the lower availability of the air separation unit compared to the air compressor.

F.7 UNCERTAINTIES

The availability estimates provided here are the best available information based on the gasifier and process unit designs, including published operating data of the Wabash River Repowering Project. The estimate of gasifier island availability is based on an average of a range (96.37% to 98.04%, exclusive of scheduled maintenance) developed based on component MTBF and MDT. A higher or lower on-stream factor can change equivalent availability. Also, differences between the two processes (Wabash River and U-GAS®), most significantly the size of the process equipment at Wabash River compared to the smaller scale industrial application studied here, may affect equipment availabilities. Finally, the availability estimate for the turbine is based on published data from Wabash River for the GE 7FA turbine, with demonstrated performance on coal-based syngas. The turbine used for this study is the much smaller GE 10 turbine that is not currently available for syngas applications.

Attachment A

Availability Nomenclature

The following table of availability nomenclature and definitions is based on material prepared by a working group of the Gasification Technology Council (GTC).[1] They have been supplemented by terms used in this study.

Availability - The yearly production of the unit or a portion thereof divided by the design production, expressed as a percentage. When expressed on a time basis, the percent of time the unit(s) is operating at a useable capacity.

Average Daily Production - The yearly production divided by 365.

Capacity Factor - The yearly production of the unit divided by the design production, expressed as a percentage.

Design Production - The maximum production that the unit would produce at the design rate over the calendar year when operated in an integrated manner. Calculated by multiplying the average annual daily design rate by 365. Note that the Design Production can change over time as the plant is debottlenecked.

Equivalent Availability - Similar to availability. Average annual daily production rate divided by the design production rate, expressed as a percentage.

Forced Outage Rate - Defined as the time during which the downstream unit or customer did not receive product divided by the time during which they expected product, expressed as a percentage.

On-Stream - Percent of the year the unit was operating and supplying product in a quantity useful to the downstream unit or customer.

Planned Outages - Percent of the year that the unit is not operated due to outages which were scheduled at least one month in advance. Includes yearly planned outages as well as maintenance outages with more than one month notice.

Product Not Required - Percent of the year that the product from the unit was not required, and therefore, the unit was not operated. The unit was generally available to run and not in a planned or forced outage.

Unplanned Outages - Percent of the year the unit was not operated due to forced outages which had less than one month notice. Includes immediate outages as well as maintenance outages with less than one month notice.

[1] James M. Childress, email entitled “Gasification Plant Availability Reporting Guidelines, Oct. 4, 2001.

Yearly Production - The total amount of product actually produced from the unit in a calendar year. For gasification units, the GTC prefers to have production reported on the basis of total fuel LHV.

G.1 INTRODUCTION

The objective of these Design Bases is to define the process units and process support units including plant configuration for Subtask 3.2. This section includes the design basis and criteria for the subsequent engineering study and capital cost estimates. Subtask 3.2 is the base cases for the later optimized cases defined in Subtasks 3.3 and 3.4. Subtask 3.2 is defined as follows:

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

G.2 SUBTASK 3.2, PRELIMINARY DESIGN FOR EASTERN COAL CASE

G.2.1 Plant Description

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

- Unit 100: Coal Prep/Handling
- Unit 150: Air Separation or Compression Unit
- Units 200, 300 & 400: Gasification Island
- Unit 500: Ash Handling
- Unit 600: High Temperature Heat Recovery
- Unit 700: Water Scrubber, COS Reactor, Low Temperature Heat Recovery and Mercury Removal
- Unit 800: Amine Unit, SWS, Sulfur Plant, Tail Gas Clean-up
- Unit 900: Power Block including the gas turbines (GT) and heat recovery steam generator (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)

A block flow diagram of the plant is shown in Figure G.1 (Section G.13).

G.2.2 Site Selection

The upstate New York site is a large industrial site of over 1800 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 has been prepared.

G.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. We will use an existing analysis of a Southeastern Ohio coal as representative of the coals from this region. Seeking fuel bids and mine analysis at this time is not practical for the study. The coal analysis to be used for this study is shown in Section G.4 of this addendum.

Coal delivery to the site is by rail. Drying facilities will be designed to handle up to 15% moisture and to produce dried coal to the gasifier with a maximum of 5% moisture.

G.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.

G.2.5 Configuration

The plant has two 50% gasifier vessels operating in parallel each with a 345.7 and 336.2 TPD (dry) of coal capacity for the air-blown and oxygen-blown cases, respectively. The gasifiers will operate at 340 psig.

G.2.6 Gasification Unit

Although a one gasifier plant has a lower capital cost than a two gasifier plant, GTI has determined that it is not practical to feed two (2) gas turbines from a single gasifier on a regular basis, as requested by the client. This is because it places too large a turndown demand on the equipment. GTI will design the system for two gasifiers feeding two turbines. In the event one gasifier becomes unavailable, the remaining gasifier can be "turned-up" to its maximum capability and possibly supplemented with natural gas to fuel the second turbine.

Downstream of the gasifier each reactor will have a 3 stage cyclone system followed by a high temperature heat removal system generating 400 psig steam, a low temperature heat removal system, and a water scrubber.

The two trains will then be combined into one train for sulfur removal. This will include a COS hydrolysis unit, acid gas removal using an amine system, and a Claus unit for sulfur recovery followed by a tail gas cleanup unit.

G.2.7 Air Separation or Compression Unit

Both an oxygen-blown gasifier and an air-blown gasifier will be evaluated. The former will require an air separation unit, while the latter will only require an air compression. There will be no nitrogen, oxygen or argon export.

G.2.8 Power Block

Two (2) gas turbines (GE-10) are specified with a nominal rating of 12.5 MW each for a total of 25 MW.

G.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

G.4 FEEDSTOCKS

Ultimate Analysis	Dry Basis	As Determined	As Received	ASTM Method
C	74.65	72.38	68.38	D3176
H	5.79	5.95	5.3	D3176
N	1.54	1.49	1.41	D3176
S	3.32	3.22	3.04	D4239
Ash	5.91	5.73	5.41	D3176
O	8.79	11.23	8.06	D3176
Total	100	100	91.6	
Proximate Analysis				
Residual Moisture		3.04		D3173
Total Moisture			8.4	D3302
Ash	5.91	5.73	5.41	D3174
Volatile Matter	43.24	41.92	39.6	D3175
Fixed Carbon	50.85	49.31	46.59	D3172
Total	100	100	100	
Air-Dry Loss			5.53	D2013
Sulfur	3.32	3.22	3.04	D4239
Gross Caloric Value (Btu)	13,590	13,177	12,448	D1898
Dry, Ash Free	14,443			
Pounds SO2/MBtu	4.88			

Ash Fusion	Reducing
IT	1974
ST	2025
HT	2049
FT	2067
Coal Ash Analysis	%
SiO2	33.3
Al2O3	29.6
Fe2O3	29.3
TiO2	0.6
CaO	2.9
MgO	0.7
Na2O	0.4
K2O	0.5
SO3	2.1
P2O5	< 0.1
BaO	< 0.1
Mn2O3	< 0.1
SrO	< 0.1
Total	99.4

G.5 ELECTRIC POWER

	Air-Blown Case	Oxygen-Blown Case
Export Power, MW (actual)	22.25	23.86
Voltage, kV	230	

G.6 EXPORT STEAM PRODUCTION

	Air-Blown Case	Oxygen-Blown Case
Medium Pressure Steam		
Flow Rate, MLb/Hr (actual)	101.7	26.8
Pressure at Delivery, psig	400	
Temperature at Delivery, ^{o}F	550	

G.7 WATER MAKE-UP

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

G.8 NATURAL GAS

HHV, Btu/SCF	1050
LHV, Btu/SCF	960

G.9 BY-PRODUCTS

	Air-Blown Case	Oxygen-Blown Case
Ash, tpd (actual)	2097	1466
Sulfur, tpd (actual)	907	870

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

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

G.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
Construction interest rate	10%
Tax rate	40%
Debt-to-equity ratio	2:1
Cost of capital	8%

G.12 ANNUAL ESCALATION

- at 3%, with the exception of coal (2%)

G.13 BLOCK FLOW DIAGRAM

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

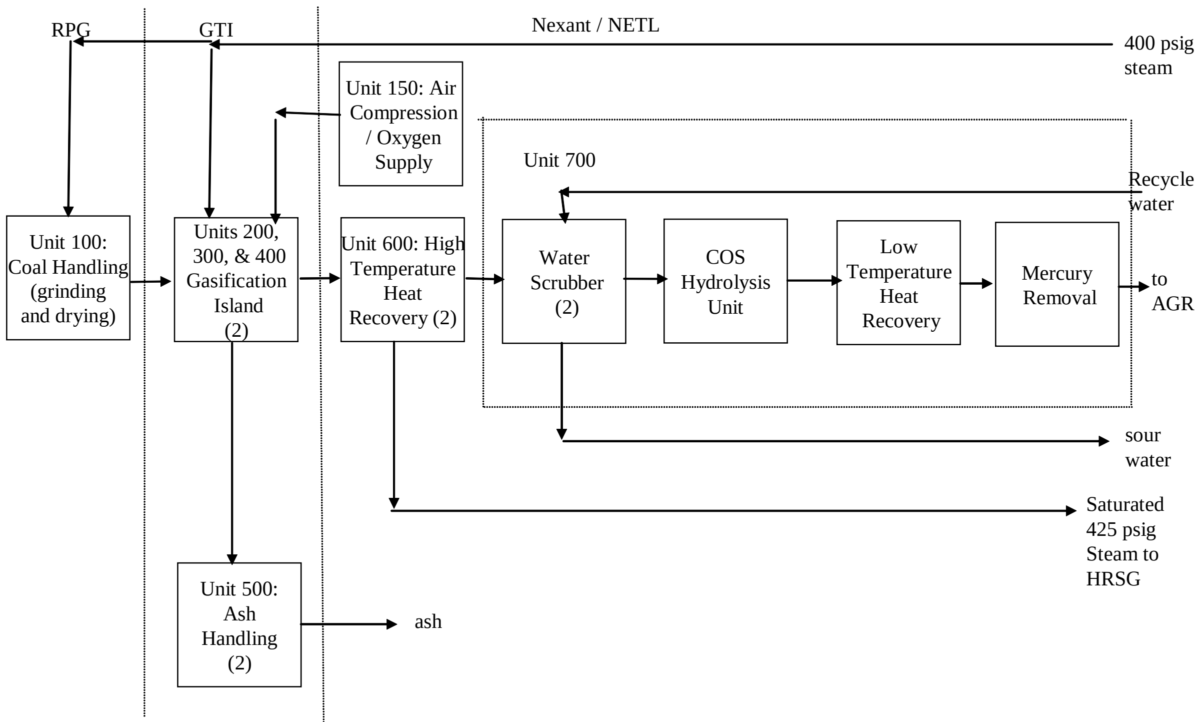


Figure G.2 Block Flow Diagram - Sulfur Removal and Recovery, Sour Water Stripper and Power Block