Prepared for:
United States Department of Energy National Energy Technology Laboratory
Contract No. DE-AC26-99FT40342

Gasification Plant Cost and Performance Optimization
Task 2 Topical Report Coke/Coal Gasification With Liquids Coproduction

Volume 1

Submitted By:

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Global Energy

Nexant

September 2003

U. S. Department of Energy
National Energy Technology Laboratory (NETL)

Gasification Plant Cost and Performance Optimization (Contract No. DE-AC26-99FT40342)

Task 2 Topical Report
Coke/Coal Gasification With Liquids Coproduction

Volume 1 of 2

September 2003

Work Performed from September 1999 to September 2003

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DISCLAIMER

PREFACE

This report was prepared as an account of the Task 2 work on Department of Energy Contract DE-AC26-99FT40342 by Bechtel, Global Energy, and Nexant. Since all the technical work under this contract was completed, ConocoPhillips acquired the proprietary gasification technology from Global Energy Inc. on August 7, 2003. Thus, the patents and intellectual property associated with the E-GAS^{TM} technology for gasification now are the property of ConocoPhillips who should be contacted for further information concerning the technology.

Abstract

This report describes Task 2 of a Department of Energy sponsored study (DOE contract DE-AC26-99FT40342) that extended the investigation of petroleum coke and coal fueled IGCC power plants to those that that co-produce liquid transportation fuel precursors using Fischer-Tropsch hydrocarbon synthesis technology. Task 2 is divided into three subtasks.

In Task 1, Bechtel, Nexant and Global Energy, Inc. developed optimized designs for several coal and petroleum coke IGCC power and coproduction projects. The as-built design and actual operating data from the DOE sponsored Wabash River Coal Gasification Repowering Project provided a firm starting point. Optimized designs were developed for:

- A petroleum coke fueled IGCC power plant that co-produces hydrogen and steam for an adjacent power plant (Subtask 1.3 Next Plant)
- An advanced single-train coal fueled IGCC power plant (Subtask 1.4)
- A single-train coal fueled IGCC power plant (Subtask 1.5A)
- A single-train petroleum coke fueled IGCC power plant (Subtask 1.5B)
- A four train, nominal 1,000 MW coal fueled IGCC power plant (Subtask 1.6)
- A single-train coal to hydrogen plant (Subtask 1.7)

Starting from the Subtask 1.3 Next Plant, Subtask 2.1 developed a petroleum coke gasification power plant with hydrocarbon liquids coproduction by eliminating the export steam and hydrogen production facilities and replacing them with a single-train, once through Fischer-Tropsch hydrocarbon synthesis plant. This plant produces 617 MW of export power and 4,125 bpd of liquid fuel precursors from slightly less petroleum coke (5,376 vs. 5,417 dry tpd) than the Subtask 1.3 Next Plant. On a higher heating value (HHV) basis, this plant has a thermal efficiency 47.9%. It cost 818 MM mid-year 2000 dollars which is about 31 MM mid-2000 dollars more than the Subtask 1.3 Next Plant.[1]

Subtask 2.2 optimized the previous coproduction plant design by maximizing liquids production at the expense of power production. This plant produces 10,450 bpd of liquid fuel precursors and 367 MW of export power from 5,417 tpd of dry petroleum coke. It has an EPC cost of 735 MM mid-year 2000 dollars. On a higher heating value basis, this plant has a thermal efficiency 56.7%.

The Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction was developed from the Subtask 1.6 and Subtask 2.2 plants. This plant produces 12,377 bpd of liquid fuel precursors, 675.9 MW of export power, and 237 tpd of sulfur from 9,266 tpd of dry Illinois No. 6 coal. This plant has an EPC cost of 1,159 MM mid-year 2000 dollars which is about 72 MM mid-year 2000 dollars less than the Subtask 1.6 plant. On a higher heating value basis, this plant has a thermal efficiency 53.4% which is lower than that of the Subtask 2.2 petroleum coke plant because, on a relative basis, it produces less liquid fuel and more power than the coke plant.

Adding hydrocarbon liquids coproduction can improve the return of an IGCC power plant when oil prices are relatively high. This is especially true for a coke coproduction plant because besides providing a refinery with a means of disposing of the low-value byproduct coke, it makes liquids, which can be upgraded in the refinery to high-value liquid transportation fuels.

As more coal and coke IGCC plants are built, further improvements can be expected which should lead to additional cost reductions and improved availability that will make IGCC the preferred option for new base-load power plants.

[1] All costs are mid-year 2000 costs. They are presented here to show the relative differences between the cases. Current cost estimates should be developed for any proposed application.

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Executive Summary

This report describes Task 2 of a Department of Energy sponsored study (DOE contract DE-AC26-99FT40342) that extended the investigation of petroleum coke and coal fueled IGCC power plants to those that that co-produce liquid transportation fuel precursors using Fischer-Tropsch hydrocarbon synthesis technology. Task 2, which is divided into three subtasks, showed that adding hydrocarbon liquids coproduction to an IGCC power plant can be cost effective when oil prices are relatively high.

Task 1 of this "Gasification Plant Cost and Performance Optimization" project examined the current state-of-the-art of coal gasification to provide baseline optimized design cases from which the Department of Energy can measure future progress towards commercialization of gasification processes and achievement of the Vision 21 program goals. This optimization focus or metric was to minimize the cost of electric power produced by IGCC plants primarily by reducing plant capital cost, increasing efficiency, increasing overall system availability, co-producing products, and reducing operating and maintenance costs.

The Vision 21 concept is the approach being developed by the U. S. Department of Energy to promote energy production from fossil fuels in the 21st century. The objective is to integrate advanced concepts for high efficiency power generation and pollution control into a new class of fuel-flexible facilities capable of co-producing electric power, process heat, high value liquid fuels, and chemicals with virtually no emissions of air pollutants. Also, this facility will be capable of a variety of configurations to meet different marketing needs, including both distributed and central power generation.

Gasification systems are inherently clean, relatively efficient, and commercially demonstrated for converting inexpensive fuels such as coal and petroleum coke into electric power, steam, hydrogen, liquid fuels and chemicals. However, the gasification system also is relatively complex and costly. Optimization should allow IGCC to become the preferred low cost power generation option.

Starting from the DOE sponsored Wabash River Coal Gasification Repowering Project (at Terre Haute, Indiana), a design and mid-year 2000 cost were developed for a grass-roots plant equivalent to the Wabash River facility. This case updates the then current Wabash River plant by including all modifications and improvements that were made since the initial startup. The mid-year 2000 cost of the grass-roots plant was developed based on the actual construction cost of the Wabash River facility and subsequent modifications; thereby providing a sound cost basis for the subsequent cases.

Task 1 was divided into nine basic subtasks. Subtasks 1.1 and 1.2 developed non-optimized designs for coal and coke IGCC power and coproduction plants. Subtasks 1.3 through 1.7 and 1.3 Next Plant developed optimized designs for coal and coke IGCC power and coproduction plants. Subtask 8 performed a review of warm gas cleanup systems. Subtask 1.9 documented the availability analysis study (and results) that was performed as part of the Value Improving Practices portion of the optimization efforts.

Subtask 1.1 started from the DOE sponsored Wabash River Coal Gasification Repowering Project (at Terre Haute, Indiana) and developed a design and mid-year 2000 cost for a grass-roots plant equivalent to the Wabash River facility. This case updates the then current Wabash River plant by including all modifications and improvements that were made since the initial startup. The mid-year 2000 cost of the grass-roots plant was developed based on the actual construction cost of the Wabash River facility and subsequent modifications; thereby providing a sound cost basis for the subsequent cases.

Subsequent subtasks developed optimized designs for coal and petroleum coke IGCC power plants with and without coproduction. Significant reductions were achieved. For example, on a $/kW basis, the cost of the 416 MW advanced Subtask 1.4 single-train IGCC power plant was reduced by 34% compared to the 269 MW Wabash River base case (1,116 $/kW vs. 1,681 $/kW)[2]. The required power selling price for a 12% after tax ROI was reduced by about 41% to 39.8 $/MW-hr using a conservative economic scenario.[3] Further improvements that are as yet undeveloped have the potential to further reduce the plant cost, to increase the thermal efficiency, and to lower the cost of electric power.

Task 2 was divided into three subtasks. These subtasks dealt with converting two of the optimized plants developed during Task 1 into IGCC power plants with liquid fuels coproduction. Table ES-1 summarizes the three Task 2 plant designs and the relevant Task 1 designs from which they were developed.

Subtask 1.3 Next Plant developed an optimized design, cost estimate and economics for a Petroleum Coke IGCC Coproduction Plant processing about 5,417 tpd of dry petroleum coke and producing about 80 MMscfd of hydrogen and 980,000 lb/hr of industrial-grade steam (750^{o}F/700 psig) in addition to electric power. The Subtask 1.3 Next plant produced 474 MW of export power and 373 tpd of sulfur. It has an EPC cost of 787 MM mid-year 2000 dollars.[1]

Starting from the Subtask 1.3 Next Plant, Subtask 2.1 developed a non-optimized design for a petroleum coke gasification power plant with hydrocarbon liquids coproduction by eliminating the export steam and hydrogen production facilities and replacing them with a single-train, once through Fischer-Tropsch hydrocarbon synthesis plant. A once through system eliminates the cost of the expensive recycle system which requires recycle gas purification facilities in addition to the recycle compressor. The energy that was used to produce the export steam now is used to generate additional power. This plant produces 617 MW of export power and 4,125 bpd of liquid fuel precursors from slightly less petroleum coke (5,376 vs. 5,417 dry tpd) than the Subtask 1.3 Next Plant. On a higher heating value (HHV) basis, this plant has a thermal efficiency 47.9% when the heating value of the byproduct sulfur is included. It cost 818 MM mid-year 2000 dollars.[1]

Subtask 2.2 developed an optimized design for a petroleum coke gasification power plant with hydrocarbon liquids coproduction by maximizing the liquid fuels production at the expense of power production. In this design, about 92% of the syngas goes through the once-through slurry-bed F-T hydrocarbon synthesis reactor. The unconverted syngas and light hydrocarbons from the F-T area are mixed with the remaining 8% of the syngas, compressed, and sent to the single gas turbine for power generation. This plant produces 10,450 bpd of liquid fuel precursors and 367 MW of export power from 5,417 tpd of dry petroleum coke. It has an EPC cost of 735 MM mid-year 2000 dollars. On a higher heating value basis, this plant has a thermal efficiency 56.7% when the heating value of the byproduct sulfur is included and 54.9% when the byproduct sulfur is not included. With 27

[2] All costs are mid-year 2000 costs. They are presented here to show the relative differences between the cases. Current cost estimates should be developed for any proposed application.
[3] All power costs are current year 2000 power costs which increase at 1.7%/year.

$/MW-hr and 30$/bbl liquids, this plant has a 18.2% ROI, and the Subtask 2.1 plant only has a 9.50% ROI. (Both cases assume an 80% loan rate at 10% annual interest.

Subtask 1.6 developed a current day optimized design, cost estimate and financial analysis for a nominal 1,000 MW coal fed IGCC power plant using four gasifiers and four GE 7FA+e combustion turbines. The plant consumes 9,266 tpd of dry Illinois No. 6 coal and generates 1,155 MW of export power. It cost 1,231 MM mid-year 2000 dollars (1,066 $/kW) and can export power at 44.4 $/MW-hr without natural gas backup while producing a 12% ROI. With 2.60 $/MMBtu backup natural gas, the required power selling price for a 12% ROI drops to 40.2 $/MW-hr. On a higher heating value (HHV) basis, this plant has a thermal efficiency 42.4% when the heating value of the byproduct sulfur is included.

Subtask 2.3 developed a design for an Optimized Coal Gasification Power Plant with Liquids Coproduction from the Subtask 1.6 plant using the design approach adopted for the optimized Subtask 2.2 coke plant. The coal gasification capacity of the plant was kept the same as Subtask 1.6. F-T liquids production was maximized, and power production was reduced to only one power block train consisting of two combustion turbines, two HRSGs, and a single steam turbine. The unconverted syngas and light hydrocarbons from the F-T synthesis section is compressed and combined with the 18% of syngas bypassing the F-T reactors to provide fuel for the two combustion turbines.

The plant produces 12,377 bpd of liquid fuel precursors, 675.9 MW of export power, and 237 tpd of sulfur from 9,266 tpd of dry Illinois No. 6 coal. This plant has an EPC cost of 1,159 MM mid-year 2000 dollars. On a higher heating value (HHV) basis, the plant has a thermal efficiency 53.4% when the heating value of the byproduct sulfur is included. This thermal efficiency is lower than that of the Subtask 2.2 optimized petroleum coke coproduction plant because this plant produces less liquid fuel and more power on a relative basis that the coke plant. With 30 $/bbl liquids and 2.60 $/MMBtu natural gas, this plant requires a power selling price of 42 $/MW-hr to produce a 12% ROI whereas the Subtask 1.6 plant requires a power selling price of only 40.2 $/MW-hr.

Enlarging the gasification train capacity of the coal plant by 33% so that the plant would have three operating trains and a spare gasification train to make it similar to that of the petroleum coke case, would improve the ROI by about 6 to 8%. With 30 $/bbl liquids, the plant still would require power selling prices of 40 plus $/MW-hr to justify building the facility.

As more IGCC plants, either with or without coproduction facilities, are built and operated, availability should improve which will improve the plant ROI at given power price, or lower the required product selling prices for a given ROI. At low power prices relative to oil prices, IGCC power plants with liquid fuels coproduction will be favored, and conversely when power prices are high relative to oil prices, IGCC power only power plants will be preferred.

Based on the above results, in order for a gasification power plant with liquids coproduction to have a better ROI than a conventional IGCC power plant, the plant design must be balanced. Some features that contribute to this balanced design include

- The use of large, cost efficient gasification trains to minimize cost
- Inclusion of a spare gasification train for maximum availability
- The syngas should have high CO and H_2 contents and a low methane content to allow the F-T area to produce an offgas with a minimal Btu content.

- High conversion in the F-T section so that it can produce an offgas with a high CO_2 content for NOx control
- The ability to process all, or almost all, of the syngas in the F-T reactors
- A large, efficient combustion turbine that is correctly sized to process all the fuel gas with minimum additional steam dilution for NOx control

The Subtask 2.2 Optimized Petroleum Coke IGCC Power Plant with Liquids Coproduction does a good job of satisfying most of the above criteria. However, the Subtask 2.3 coal plant produces a syngas with a methane content that is about 2.6 times greater than the syngas produced by the gasification of coke because of the higher volatiles content of the coal. As a result, the F-T offgas has a higher Btu content and requires more steam dilution for NOx control. Furthermore, the total amount of F-T offgas contains too much energy for one GE 7FA+e turbine, and not enough for two turbines. Consequently, about 18% of the syngas has to be bypassed around the F-T reactors to fully load the two GE 7FA+e turbines. This significantly reduces the liquids production. Ideally, a single larger turbine [or two smaller turbines] that would require bypassing only very little, if any, syngas around the F-T reactors would result in a better balanced plant that could have a better return on investment.

The balanced approach in which the gas turbine fuel gas is diluted with CO_2 to a level where only minimal or no additional steam dilution for NOx control also could be applied to an ICGG power plant that co-produces hydrogen (instead of liquid fuels) for power generation with fuel cells. In such a plant, CO_2 production by the shift reaction that is in excess of that needed for NOx control would be captured for possible sequestration.

Gasification is viewed as the environmentally superior process for power generation from coal. The Wabash River facility demonstrated the superior environmental performance of gasification in terms of SOx, NOx, and particulate emissions. In a carbon-constrained environment, the CO_2 easily can be captured for sequestration or other uses. Even without CO_2 capture, CO_2 emissions are reduced because gasification plants are more efficient than conventional coal power plants.

With low coal and coke prices and high oil prices, the return of a gasification power plant can be improved by adding hydrocarbon liquids coproduction. This is especially true for a coke plant associated with a petroleum refinery because besides providing a means of disposing of the byproduct coke, the plant can convert it into liquid hydrocarbons, which when upgraded in the refinery become the main refinery products, liquid transportation fuels.

As natural gas and power prices increase and environmental constraints for coal fired generation plants tighten, coal IGCC will further penetrate the power market. As more coal and coke IGCC plants are built, further improvements can be expected which should lead to additional cost reductions and improved availability that will make IGCC the preferred option for new base-load power plants.

Table ES-1

Task 1 and 2 Coal and Coke IGCC Case Summaries

Case Description	Subtask 1.1 Wabash River Greenfield	Subtask 1.3 Next Optimized Pet Coke IGCC Coproduction Plant	Subtask 1.5 Single Train Power 1.5A Coal		Subtask 1.6 1,000 MW Coal IGCC Power Plant	Subtask 2.1 Petroleum Coke to Liquids and Power	Subtask 2.2 Optimized Pet Coke to Liquids and Power	Subtask 2.3 Optimized Coal to Liquids and Power
			1.5A Coal	1.5B Coke				
Configuration								
Plant Location	Midwest	Gulf Coast	Gulf Coast	Gulf Coast	Midwest	Gulf Coast	Gulf Coast	Midwest
Number of Air Separation Units	1	2	1	1	3	2	2	3
Number of Gas Turbines	1	2	1	1	4	2	1	2
Number of Gasification Trains	1	3	1	1	4	3	3	4
Number of Gasification Vessels	2	3	2	2	4	3	3	4
No of Syngas Processing Trains	1	2	1	1	2	2	2	2
Number of 50% H2 trains	0	2	0	0	0	0	0	0
Number of F-T Liquid Trains	0	0	0	0	0	1	1	1
Design Feed Rates								
Feedstock Type	Coal	Pet Coke	Coal	Pet Coke	Coal	Pet Coke	Pet Coke	Coal
Coal or Coke, TPD as received	2,642	5,692	2,754	2,077	10,837	5,649	5,684	10,837
Coal or Coke, TPD dry	2,259	5,417	2,355	1,977	9,266	5,376	5,417	9,266
Feed, MMBtu HHV/hr	2,400	6,703	2,481	2,446	9,844	6,652	6,703	9,844
Feed, MMBtu LHV/hr	2,311	6,567	2,389	2,397	9,478	6,518	6,567	9,478
Flux, TPD	0	110.6	0	40.3	0	109.7	110.6	0
Water, gpm	2,790	5,223	2,840	2,525	9,752	6,472	5,693	7,403
Condensate, Mlb/hr	---	686	---	---	---	---	---	---
Oxygen, TPD of 95% O2	2,130	5,954	2,015	2,143	8,009	5,919	5,877	7,919
Oxygen, TPD of O2	2,009	5,615	1,900	2,021	7,553	5,582	5,542	7,468
Design Product Rates								
Electric Power, MW	269.3	474.0	284.6	291.3	1,154.6	617.0	366.9	675.9
Steam (750^{o}F/700 psig), lb/hr	---	980.0	---	---	---	---	---	---
Hydrogen, MMscfd	---	80.0	---	---	---	---	---	---
Sulfur, TPD	57	373	60	136	237	371	373	237
Slag (@ 15% water), TPD	356	195	364	71	1,423	194	195	1,423
Fuel Gas, MMBtu HHV/hr	---	0	---	---	---	---	---	---
Solid Waste to Disposal, TPD (4)	---	---	---	---	---	0.95	1.31	1.72
Liquid Hydrocarbons, bpd	---	---	---	---	---	4,125	10,450	12,377
Gas Turbine								
Type	GE 7FA	GE 7FA+e	GE 7FA+e	GE 7FA+e	GE 7FA+e	GE 7FA+e	GE 7FA+e	GE 7FA+e
Fuel Input, Mlb/hr	411.4	1,016.8	447.0	426.7	1,741.6	1,092.8	1,000.8 (5)	1,303.0
Heat Input, MMBtu/hr LHV	1,675	3,592	1,796	1,796	7,184	3590	1,763.3	3,532
Steam Injection, Mlb/hr	111.0	395.7	246.8	272.3	1,037.8	531.6	0	510.5
Gross Power Output, MW	192	420	210	210	840	420	199.4	416
Cold Gas Efficiency (HHV), %	76.9	77.5	77.8	77.4	78.0	77.5	77.7	78.3
Steam Turbine Power, MW	118	164.3	113	121	465.2	307.0	274.9	403.6
Internal Power Use, MW	41	110	38.4	40.7	151	110.0	107.4	118.8
Heat Rate, HHV Btu/kW-hr	8,912	NA	8,717	8,397	8,526	NA	NA	NA
Thermal Efficiency, % HHV (1)	38.3	NA	39.1	40.6	40.0	46.0	54.9	52.6
Emissions								
SOx as SO2, lb/hr	312	350	142	119	438	321	276	329
NOx as NO2, lb/hr	161	166	69	69	275	136	94	166
CO, lb/hr	49	89	33	34	131	66	37	65
Sulfur Removal, %	96.7	99.4	98.5	99.4	98.9	99.5	99.6	100
Performance Parameters								
Tons 02 / Ton of Dry Feed	0.889	1.037	0.807	1.022	0.815	1.038	1.023	0.806
Gross MW / Ton of Dry Feed	0.137	0.108	0.137	0.168	0.141	0.135	0.088	0.088
Net MW / Ton of Dry Feed	0.119	0.088	0.121	0.147	0.125	0.115	0.068	0.073
Emissions								
SOx (SO2) as lb/MW-hr	1.159	0.738	0.499	0.409	0.379	0.520	0.752	0.487
SOx (SO2) as lb/MMBtu (HHV)	0.130	0.052	0.057	0.049	0.044	0.048	0.041	0.033
NOx (NO2) as lb/MW-hr	0.598	0.350	0.242	0.237	0.238	0.220	0.256	0.246
NOx (NO2) as lb/MMBtu (HHV)	0.067	0.025	0.028	0.028	0.028	0.020	0.014	0.017
CO, lb/MW-hr	0.182	0.188	0.116	0.117	0.113	0.107	0.101	0.096
CO, lb/MMBtu (HHV)	0.020	0.013	0.013	0.014	0.013	0.010	0.006	0.007
Daily Average Feed/Product Rates with Backup Natural Gas (Subtask 1.1 is without Backup Natural Gas. Subtask 2.2 purchases power.)								
Coal or Coke, TPD dry	1,705	4,842	1,826	1,546	7,018	4,805	4,984	6,929
Coal or Coke, % of design	75.5%	89.4%	77.5%	78.2%	75.7%	89.4%	92.0%	74.8%
Power, MW	203.2	448.4	264.4	269.4	1,081	572.5	316.4	613.7
Power, % of design	75.5%	94.6%	92.9%	92.5%	93.6%	92.8%	86.2%	90.8%
Steam, lbs/hr	---	974.6	---	---	---	---	---	---
Steam, % of design	---	99.4%	---	---	---	---	---	---
Hydrogen, MMscfd	---	78.8	---	---	---	---	---	---
Hydrogen, % of design	---	99.4%	---	---	---	---	---	---
Fuel Gas, MMBtu HHV/hr	---	0	---	---	---	---	---	---
Fuel Gas, % of design	---	---	---	---	---	---	---	---
Natural Gas, Mscfd	NA	9,059	6,929	6,929	34,960	8,856	0	26,466
Liquid Hydrocarbons, bpd	---	---	---	---	---	3,938	9,702	10,397
Liquid Hydrocarbons, % of design	---	---	---	---	---	95.5%	92.8%	84.0%
Plant Cost, MM mid-2000 $ (2)	452.6	787.3	375.0	367.0	1,231.3	817.9	735.3	1159.1
Plant Cost, $/design kW	1,681	NA	1,318	1,260	1,066	NA	NA	NA
Required Electricity Selling Price for a 12% ROI, $/MW-hr (3)								
Without Natural Gas Backup	67.5	---	53.9	43.9	44.4	28.8	19.5	48.1
With Natural Gas Backup	---	30.0	48.9	40.6	40.2	29.0	17.7	42.0

NA = Not Applicable
July 31, 2003

1. Without including the sulfur byproduct, but including the F-T liquid fuels, when appropriate.
2. All costs are mid-year 2000 EPC costs which exclude contingency, taxes, fees and owners costs. They are presented here to show the relative differences between cases. Current cost estimates should be developed for any proposed applications.
3. Power selling prices are presented to show a relative comparison between cases. Based on a natural gas price of $2.60 $/MMBtu and a liquids price of 30 $/bbl. Subtask 2.2 purshases power at the power selling price rather than natural gas.
4. Used COS hydrolysis catalyst, Used ZnO sulfur sorbent, and used F-T catalyst, all on a dry, hydrocarbon free basis. The used activated carbon in Subtasks 2.2 and 2.3 is mixed with the gasifier feed and converted to syngas and slag.
5. Includes 57.8 Mlbhr of steam is added to the fuel to get a net heating value of 147.1 Btu/scf. No additional steam is needed for NOx control.

Chapter I

Introduction

The *Vision 21* concept is the approach being developed by the U. S. Department of Energy to energy production from fossil fuels in the 21st century. The objective is to integrate advanced concepts for high efficiency power generation and pollution control into a new class of fuel-flexible facilities capable of co-producing electric power, process heat, and high value fuels and chemicals with virtually no emissions of air pollutants. Hopefully, it will be capable of a variety of configurations to meet different marketing needs, including both distributed and central power generation.

Vision 21 builds on technology advancements being made in the Energy Department's Fossil Energy Program. It will integrate ongoing research and development in advanced coal and biomass gasification and combustion with next-generation fuel cells, high-performance turbine technology, and advanced coal conversion systems.

A *Vision 21* plant will be capable of using a variety of fuels, including coal and natural gas, perhaps mixed with petroleum coke, biomass, or municipal wastes. In contrast to today's single product energy facilities, a *Vision 21* plant could produce a multiple slate of products: electricity, liquid and/or gaseous fuels, and industrial-grade heat and/or steam.

In the Department of Energy's Fossil Energy Program. *Vision 21* will serve as a "roadmap" for future electric power and fuels research and development efforts. Key technologies will be developed as modules with the goal of combining them into highly flexible energy complexes. The *Vision 21* roadmap will establish technical specifications for integrating these modules. It will focus on the engineering challenges of reliability and operability of an integrated "energyplex." Furthermore, it will identify the research and development objectives that are needed to establish the technological foundation for an entirely new fleet of energy facilities that could be deployed in the 2010-2030 timeframe.

Specifically, the *Vision 21* goals are:

> **Power:** Generating efficiencies greater than 60% using coal and greater than 75% using natural gas. For comparison, current coal technology is 33 to 35% efficient, and current natural gas technology is 45 to 55% efficient.
>
> **Combined Heat and Power:** Overall thermal efficiencies of 85 to 90%.
>
> **Enviromental:** Near zero emissions for all traditional pollutants, including smog- and acid rain-forming pollutants.
>
> **Greenhouse Gas Reduction:** Carbon dioxide emissions reduced by 40 to 50% through efficiency improvements: reduced to zero (net) if coupled with carbon sequestration.
>
> **Coproducts:** Clean, affordable transportation quality fuels at costs equivalent to an oil price of 20 $/barrel or less in 1998 dollars; also industrial-grade heat and/or steam and the potential for fuel-grade gas production.

Vision 21 will not be a single configuration. It will be a series of interconnected modules. Future designers will integrate these modules to meet specific market needs. A *Vision 21* plant might serve as the hub of an industrial complex, providing steam and/or heat in addition to electric power. Another *Vision 21* configuration might co-produce high-value chemicals or fuel gases for neighboring manufacturing facilities. Or it might be a power plant-coal refinery combination, producing electricity and liquid transportation fuels.

One of the core technologies in the Department of Energy's *Vision 21* program is coal gasification because it produces a gas stream that can be used as a source of

- energy to produce electric power, or
- hydrogen for fuel cells or chemical processes, or
- carbon and hydrogen for making high-value chemicals, or
- carbon and hydrogen for making high-quality liquid transportation fuels, or
- energy as a fuel gas for industrial plants.

This "Gasification Plant Cost and Performance Optimization" project, contract number DE-AC26-99FT40342, examines the current state-of-the-art of coal gasification to provide baseline design cases from which the Department of Energy can measure future progress towards achieving the *Vision 21* goals. This study also illustrates how advanced engineering design tools, previous design work, and operating experience acquired from the coal gasification demonstration plant can lower the plant cost and improve the overall project economics. Additional sensitivity cases were developed to demonstrate that petroleum coke gasification with hydrogen and steam coproduction is commercially ready and competitive. Operating experience from these commercial petroleum coke gasification plants will reduce the technical risk and the capital and operating costs of future coal gasification plants.

The Wabash River Repowering Project was the starting point for this study. The Wabash River project repowered an existing steam turbine by the addition of a Global Energy gasifier processing a nominal 2,500 tons/day of coal producing clean syngas for a General Electric MS 7001 7FA gas turbine and steam for powering the existing steam turbine.

This project originally was divided into three tasks. Task 1 is work that primarily deals with gasification optimization using either coal or petroleum coke as fuel. The Optimized Coal IGCC Plant will only produce electric power. The Optimized Petroleum Coke IGCC Coproduction Plant will produce hydrogen and industrial-grade steam in addition to electric power. Task 2 will study coal and petroleum coke gasification plants that will produce liquid transportation fuel precursors in addition to electric power. If implemented, Task 3 will examine conceptual designs for advanced gasification plants including the integration with fuel cells and/or the addition of carbon dioxide control technologies.

The primary objective of Task 1 was to develop optimized engineering designs and costs for five Integrated Gasification Combined Cycle (IGCC) plant configurations. 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.

- Non-optimized Petroleum Coke IGCC Coproduction Plant

- Optimized Petroleum Coke IGCC Coproduction Plants that will produce hydrogen and industrial-grade steam in addition to electric power (Subtasks 1.3 and 1.3 Next Plant – four cases)

- A future optimized Coal IGCC 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)

Figure I.1 shows the chronological development of the above gasification plant designs.

In addition there are two other subtasks. Subtask 1.8 has the objective to develop a review of various warm gas cleanup methods that are applicable to IGCC systems. The Subtask 1.8 cases cover a variety of processes and provide a look at future syngas cleanup methods. Subtask 1.9 documents 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.[1]

Task 2 has the objectives of developing optimized designs, cost estimates and economics for a petroleum coke gasification power plant with liquid fuel precursors coproduction and a coal gasification power plant with liquid 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 fuels coproduction

- An optimized petroleum coke IGCC power plant with liquid fuels coproduction

- An optimized coal IGCC power plant with liquid fuels coproduction

The starting point for these design was the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant design for the two coke cases and the Subtask 1.6 1,000 MW Coal IGCC Power Plant design for the coal case. Building on these previous cases provides common bases for comparison economics and ROI.

This report is the Topical Report for Task 2. It summarizes the three individual task reports (which are included as appendices) and discusses the overall purpose, results and potential of this work. It is divided into the following chapters.

[1] "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.

Chapter	Title
I	Introduction
II	Study Objectives, Basis, Background and Overview
III	Petroleum Coke Cases
IV	Coal Cases
V	Market Potential and Future Applications
VI	Summary and Recommendations
VII	Acknowledgements

Chapter II presents the objectives of this study, describes the study basis, briefly reviews the results of Task1, and presents an overview of the Task 2 investigation.

Chapter III summarizes the Subtask 2.1 and Subtask 2.2 petroleum coke-fueled gasification power plants with liquid fuels coproduction.

Chapter IV summarizes the Subtask 2.3 optimized coal-fueled gasification power plant with liquid fuels coproduction.

Chapter V discusses the market potential and future application of gasification power plants with liquid fuels coproduction.

Chapter VI briefly summaries the Task 2 work and provides recommendations for further work

Chapter VII acknowledges the contributions of others.

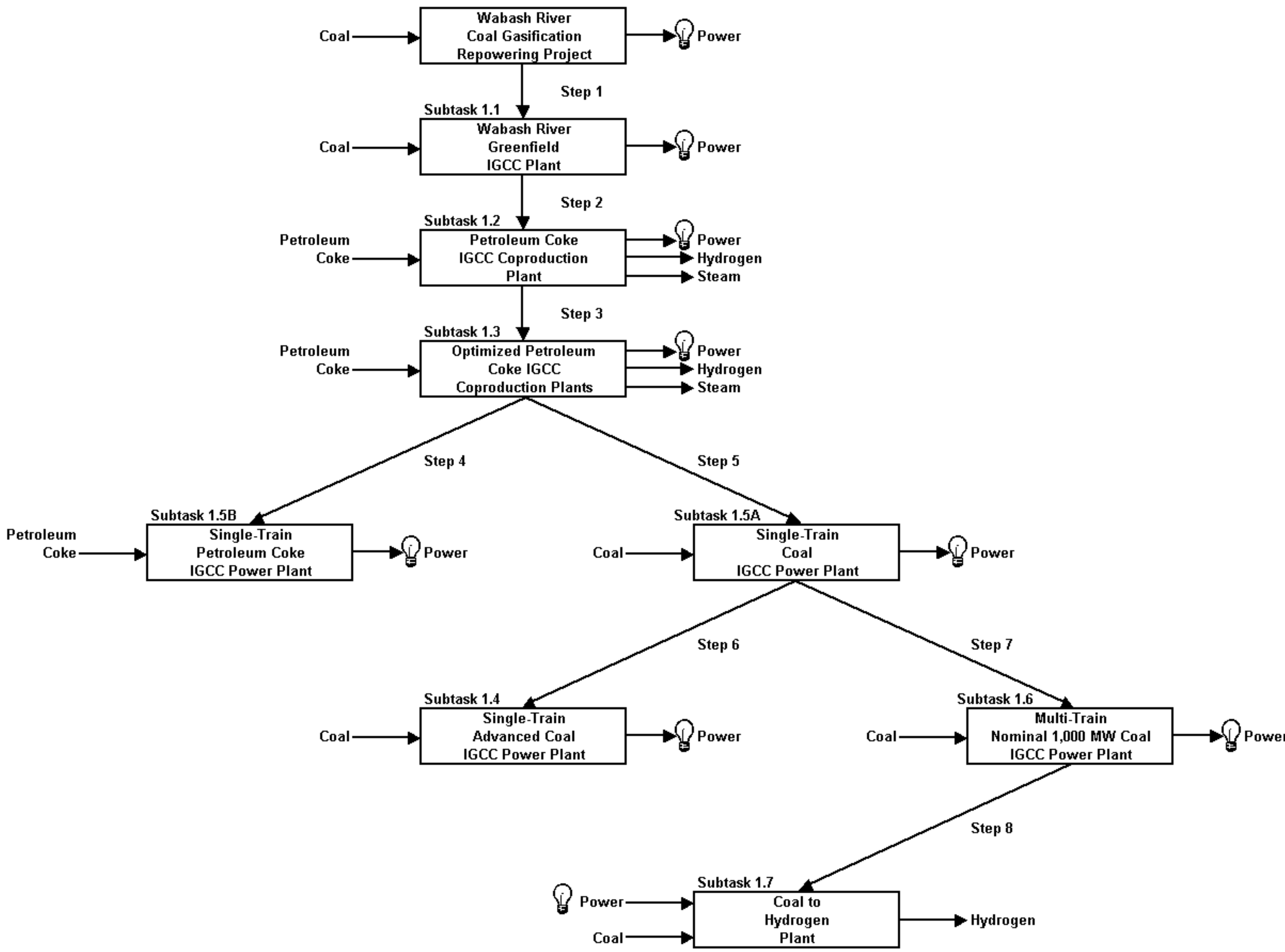
In addition this report contains the following Appendices.

Appendix	Title
A	Subtask 2.1 – A Coke Gasification Power Plant with Liquid Fuels Coproduction
B	Subtask 2.2 – An Optimized Coke Gasification Power Plant with Liquid Fuels Coproduction
C	Subtask 2.3 – An Optimized Coal Gasification Power Plant with Liquid Fuels Coproduction

Because this report describes plant designs that are based on proprietary information, some key details are omitted. However, this report contains sufficient information to allow the reader to assess the performance of Global Energy'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 PFD's and heat and material balances developed by the project team for each subtask. Design development included line sizings and marked up P&IDs for piping takeoffs. This information can be used to check the overall mass, carbon, and energy balances for the gasification plant and the power block, and possibly to adapt these to new cases. However, the project team, particularly Global Energy, would prefer to generate project specific mass and energy balances under a secrecy agreement. Such an agreement will allow Global Energy to provide additional details and to share confidential information.

Figure I.1

**Schematic Diagram Showing the
Chronological Development of the Task 1 Gasification Plant Designs**



Chapter II

Study Objectives and Methodology

II.1 Study Objectives

The objectives of this project are to examine the current state-of-the-art of coal gasification and to develop designs that will reduce the cost of power generated by IGCC plants by reducing their capital and operating costs, increasing their efficiency, and making them less polluting. Cases using a petroleum coke feedstock and co-producing hydrogen and steam or liquid fuel precursors also were developed 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 is to provide baseline cases from which the Department of Energy can measure future progress towards achieving their *Vision 21* goals.

The work is divided into two tasks. Task 1 was concerned with gasification optimization using either coal or petroleum coke as fuel. The Optimized Coal IGCC Plant only produced electric power. The Optimized Petroleum Coke IGCC Coproduction Plant produced hydrogen and industrial-grade steam in addition to electric power. Task 2 studied coal and petroleum coke gasification plants that produced liquid transportation fuel precursors in addition to electric power.

Task 1 of this project had the objective to develop optimized engineering designs and costs for four Integrated Gasification Combined Cycle (IGCC) plant configurations and a coal to hydrogen plant. Starting from the as-built design, operation, and cost information from the commercially proven Wabash River Coal Gasification Repowering Project, the following optimized cases were developed:

1. Optimized Petroleum Coke IGCC Coproduction Plants that will produce hydrogen and industrial-grade steam in addition to electric power (Subtasks 1.3 and 1.3 Next Plant – four cases)

2. A Coal IGCC Plant producing only power using a next generation gas turbine (Subtask 1.4)

3. Single-train Coal and Coke IGCC Power Plants (Subtask 1.5A [coal] and 1.5B [coke])

4. A Nominal 1,000 MW Coal IGCC Power Plant (Subtask 1.6)

5. A Coal to Hydrogen Plant (Subtask 1.7)

In addition there were two other subtasks which did not involve developing the design of an optimized plant. They are:

1. Subtask 1.8 – Reviewed the status of warm gas clean-up technology as applicable to coal and/or coke fueled IGCC power and coproduction plants. The objective was to evaluate developing technologies that operate in the 300 to 750°F temperature range, preferably closer to 750°F, and to determine their potential economic benefit.

2. Subtask 1.9 – Discuss the Value Improving Practices availability and reliability design optimization program. Starting from historic Wabash River Repowering Project data, this subtask discussed how the availability analysis and design considerations, such as the expected annual coke consumption, influenced plant performance and sparing philosophy.

The results of the Task 1 study have been previously reported in a Topical Report.[1]

Task 2 consisted of three subtasks. These subtasks are devoted to converting an optimized coke and an optimized coal IGCC power (or coproduction) plant into optimized power plants that co-produce liquid fuel precursors. These three subtasks and the appendices in which they are documented are:

3. Subtask 2.1 – Develop a design, cost estimate, and economics for a petroleum coke gasification power plant with liquid fuels coproduction starting from previous coal liquefaction studies and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. (Appendix A)

4. Subtask 2.2 – Develop a design, cost estimate, and economics for an optimized petroleum coke gasification power plant with liquid fuels coproduction starting from the Subtask 2.1plant. (Appendix B)

5. Subtask 2.3 – Develop a design, cost estimate, and economics for an optimized coal gasification power plant with liquid fuels coproduction starting from the Subtask 2.2 optimized plant and the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant. (Appendix C)

II.2 Background and Methodology

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 was completed in January 2000.[2,3]

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1 "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.

2 Topical Report No. 20, "The Wabash River Coal Gasification Repowering Project – An Update," U. S. Department of Energy, September 2000.

The participants jointly developed, separately designed, constructed, owned, and operated the integrated coal gasification combined-cycle power plant, using Destec's coal gasification technology to repower the oldest of the six units at PSI's Wabash River Generating Station in West Terre Haute, Indiana. The 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 an additional 104 MW of power. Parasitic power is 34 MW giving a total 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 increase availability. The net result of these changes has been a substantial improvement in plant operations. Furthermore, in addition to operation on Indiana and Illinois basin 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 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 representing the present day plant configuration that was the basis for developing appropriate models for the subsequent subtasks. As-built cost information was obtained and provided the cost basis for the cost estimates. Because the cost estimates are based on actual equipment purchases and construction labor use, the resulting 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.

The optimization studies for the Subtask 1.3 and Subtask 1.4 plants were done using the structured Value Improving Practices Program promoted by Independent Project Analysis, Inc. Subtasks 1.5A (coal), 1.5B (coke), 1.6, and 1.7 were developed from the basic optimized designs of Subtasks 1.3 and 1.4 with appropriate modifications.

Task 2 deals with converting the best coke and coal IGCC power plants into coproduction plants producing liquid fuel precursors. Subtask 2.1 developed a design for a [non-optimized] coke gasification power plant with liquid fuels coproduction. Subtask 2.2 improved this design to develop an optimized coke gasification power plant with liquid fuels coproduction. Subtask 2.3 developed an optimized coal gasification power plant with liquid fuels coproduction based on the Subtask multi-train power plant 1.6 and the Subtask 2.2 coke plant. The liquid products are sent to a conventional petroleum refinery for upgrading into fungible liquid transportation fuels because it would not be economic to build dedicated upgrading facilities for the small amount of liquid hydrocarbons that are produced by these coproduction plants (less than 12,500 bpd).

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

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), cost of electricity, and sensitivities.

Global Energy's operating personnel developed the operating and maintenance costs based on Wabash River experience. This is proprietary information.

II.3 Value Improving Practices

Value Improving Practices (VIPs) are focused activities aimed at removing unnecessary investment from a project scope.

Eleven industry standard VIPs were benchmarked by Independent Project Analysis, Inc. (IPA). Eight of these were selected for this project. In addition, a ninth item was added, Plant Layout Optimization which encompasses schedule optimization and some aspects of constructability. These nine items are described in detail in the Task 1 Topical report.[1]

1. Technology Selection
2. Process Simplification
3. Classes of Plant Quality
4. Value Engineering
5. Availability (Reliability) Modeling
6. Design-to-Capacity
7. Plant Layout Optimization
8. Schedule (Construction and Procurement) Optimization
9. Operating and Maintenance Savings

Value Improving Practices have proven to very successful over the years for reducing the cost of facilities, improving their efficiency, conserving raw materials, and being beneficial in many other ways. They generally are implemented in the project development stage when there is time pressure to get the project completed, and therefore, only a specific amount of time is allowed for the VIP procedures. In many of these situations, the full benefit of the VIP procedures is not realized. Because of this, there are advantages of doing the VIP procedures "off-line" where there no time pressure for completion in order to maintain the project schedule. It is in this spirit that the VIPs were applied to Global Energy's IGCC process to develop substantially improved and optimized designs.

The detailed results of the entire VIP exercise for the Subtask 1.3 and 1.4 IGCC plants are documented in a confidential VIP report.

After completion of the design of the Subtask 2.1 [Non-optimized] Petroleum Coke Gasification Power Plant with Liquid Fuels Coproduction, a Value Improving Practices meeting was held to develop ideas for optimization of the design. Representatives of Global Energy, Bechtel and Nexant attended this meeting.

II.4 Fischer-Tropsch Hydrocarbon Synthesis

Fischer-Tropsch hydrocarbon synthesis process is an old process in which synthesis gas or syngas (carbon monoxide and hydrogen) react over a catalyst to produce aliphatic hydrocarbons (principally normal paraffins and straight chain 1-olefins). It was used by Germany during the Second World War to make liquid fuels for military use. Subsequent cost reductions may have made F-T processes competitive in certain situations. Currently, there is a lot of interest in using the F-T process for monetizing remote natural gas by converting it into an easily transportable synthetic crude oil that can be upgraded to liquid transportation fuels.

In general, the F-T hydrocarbon synthesis reactions for olefins and normal paraffins can be written as

$$n\,CO + 2n\,H_2 \rightarrow C_nH_{2n} + n\,H_2O$$

$$n\,CO + (2n+2)\,H_2 \rightarrow H\text{-}(CH_2)_n\text{-}H + n\,H_2O$$

As seen from the above reaction stoichiometry, the ideal syngas composition is just over 2 moles of hydrogen for each mole of carbon monoxide.

The reaction is very exothermic. Traditionally, at a large scale the reaction has been performed over solid catalyst that is placed in small diameter tubes immersed in a cooling medium (such as boiling water) to remove the heat of reaction. The hydrocarbon product yield distribution can be characterized by a Schultz-Flory distribution in which the molar ratio of a component containing n carbon atoms to one with n+1 carbon atoms is a constant called alpha (α). As the reaction temperature increases, the yield distribution shifts to lighter hydrocarbons; i.e., the α parameter gets smaller. As time has progressed, more sophisticated mathematical yield models using multiple α parameters have been developed to represent the F-T reaction yields.

In the 1950s, the slurry-bed reactor was developed in which fine catalyst particles are suspended in a liquid, and the reactant syngas is bubbled up through the catalyst/liquid mixture. Steam is generated within cooling coils immersed in the slurry-bed to remove the heat of reaction. Iron-based catalysts promote the water gas shift reaction which produces hydrogen from carbon monoxide and water; whereas cobalt catalysts generally do not

In the early 1990s, Bechtel developed several designs for indirect coal liquefaction plants using Fischer-Tropsch technology).[4,5] The Baseline plant consumes 20,323 tpd of ROM Illinois No. 6 coal (8.6 wt% water) and 3,119 bpsd of normal butane to produce at total of 50,491 bpsd of petroleum products (1,921 bpsd of C3 LPG, 23,915 bpsd of gasoline, and 24,655 bpsd of distillate fuels). Appendix A contains a complete description of this plant.

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[4] "Topical Report – Volume I, Process Design – Illinois No. 6 Coal Case with Conventional Refining", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.

"Topical Report – Volume IV, Process Flowsheet (PFS) Models", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.

[5] "Topical Report VI – Natural Gas Fischer-Tropsch Case, Volume II, Plant Design and Aspen Process Simulation Model", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, August, 1996.

II.5 Availability Analysis

The common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period, all are dependent on the project cash flow. The net cash flow is the sum of all project revenues and expenses. Depending upon the detail of the financial analysis, the cash flow streams usually are computed on 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. In order to predict the annual revenue stream, an availability analysis that considers all of the above factors must be performed to predict the annual production and annual revenue streams to develop a meaningful financial analysis.

On this basis, an availability analysis was performed on each of the cases considered in Task 1 of this study to determine the applicable revenue streams and the ROI.

Appendix J of the Task 1 Topical Report contains a more detailed explanation of the availability analysis studies.[1]

II.5.1 Availability Analysis Basis

In Table 5.0A of the Final Report for the Wabash River Repowering Project, Global Energy reported downtime and an availability analysis of each plant system for the final year of the Demonstration Period.[3] This information is summarized in Table 1 of Appendix J of the Task 1 Topical Report. 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 three adjustments, this data was used to estimate the availability of the Task 1 coal and petroleum coke IGCC plant designs. The first adjustment increased the availability of the air separation unit (ASU) from the observed availability of 96.32% to the industry average availability of 98%. The second adjustment was to improve the availability of the first gasification stage by negating the impact of a slag tap plugging problem caused by an unexpected change in the coal blend to the gasifier. For the Subtask 1.2 and 1.3 plants, this adjustment is justified since a dedicated petroleum coke plant would be very unlikely to experience this problem. The third adjustment eliminated a short outage that was caused by a service interruption in the water treatment facility because sufficient treated water storage will be available to handle this type of outage.

Subsequently, Clifton Keeler has reported improved availability of the Wabash River Repowering Project.[6] However, for consistency with the Task 1 results, the same availability basis was used for Task 2. Thus, the reported financial results for both tasks are on the conservative side.

Based on the reported Wabash River data, availability analyses were calculated using the EPRI recommended procedure.[7] 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 formulas based on a probabilistic approach for predicting the availability of all trains or of 1 of 2, 2 of 3, 1 of 3, etc. Appropriate combinations of these mathematical formulas are used to represent plants with some portions containing multiple trains or spare equipment and other portions 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 or refractory replacement. Consequently, the annual availabilities reported in this study will be lower than those studies which do not consider scheduled outages.

II.5.2 Use of Natural Gas

To improve the yearly power output from single train gasification plants, backup natural gas is used to fire the gas turbine to make power when syngas is unavailable. Thus, for most of the year power is made from the lower cost coal, but for those times when the syngas generation portion of the plant is unavailable and the economics are favorable, power can be produced from higher priced natural gas. Multiple train power plants can be operated in a similar manner when insufficient syngas is available to fully load all the gas turbines.

The situation with the Subtask 1.3 Next Plant and Subtask 2.1 petroleum coke coproduction plants is somewhat different. The gasification train in Subtask 1.3 is sized so that one train has sufficient capacity to provide the design amounts of hydrogen and steam to the adjacent petroleum refinery at the expense of export power production. The situation is similar in Subtask 2.1 in that one gasification train can fully feed the F-T liquid synthesis area. However, when only one gasification train is operating, insufficient syngas is available to fully fire even one combustion turbine. Thus, in this case, natural gas can be used to supplement the syngas and fire one or both of the combustion turbines. When this situation occurs, the power output from the combustion turbines is reduced.

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

[7] 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.

Appendix A discusses several other situations for Subtask 2.1 when natural gas can be used to fire the combustions turbine(s) which leads to three possible operating scenarios, a full backup gas case, a minimal natural gas use case, and an maximum power case. .

In Subtask 2.2, most of the syngas goes to F-T liquids production area, and the unconverted syngas and light hydrocarbons from the F-T area provide most of the fuel for the combustion turbine. When one gasifier is out of service, essentially two options are available. The first is to maximize F-T liquids production by sending all the gas through the F-T area and supplement the F-T off gas with natural gas. The second option is to reduce the F-T liquids production by bypassing more syngas around the F-T liquids synthesis area to fully load the turbine. When the combustion turbine is out of service, the plant will purchase power from the grid to maintain F-T liquids production. Appendix B describes these two options in more detail.

For Subtasks 1.6 and 2.3, the average daily natural gas rates were calculated as part of the availability analysis and are shown later in this report.

In all cases, natural gas usage during startup and during maintenance operations, such as for curing refractory, are not considered in the availability analysis calculations, but are included in the operating and maintenance costs during the financial analysis.

II.6 Commodity Pricing

At the start of this project in early 2000, an economic and financial environment for the discounted cash flow evaluations of this project was assumed based on reasonable future projections. This set of economic conditions was used for all the discounted cash flow financial analyses performed in this study. Table II.1 contains a list of most of these economic assumptions. The commodity prices are based on long term projections for the U. S. Gulf Coast (except the coal price which is a Mid-West price). In this price structure, the hydrogen and steam prices were set based on their cost of production from 2.60 $/MMBtu natural gas. Also, an in-house combined cycle model predicts a required electricity price of about 35 $/MW-hr for a 12% after tax ROI with natural gas at 2.60 $/MMBtu. The inflation rates generally are based on the Energy Information Administration's *Annual Energy Outlook 2001.*[8]

However, since the time when these commodity prices were set, the economic scenario has changed. Natural gas prices have spiked to 9-10 $/MMBtu, dropped to below 3.00 $/MMBtu, and now are almost 6.00 $/MMBtu.[9] Oil prices also have had wide fluctuations over the past few years as a result of the economic slowdown, OPEC actions, and the political situation in the Middle East. Now they are in the 25 to 30 $/bbl range. Studies have shown that the F-T liquids can be more valuable than crude oil. The specific amount can range from only a couple of $/bbl up to 10 $/bbl depending upon the refinery configuration, the crude oils being replaced, and the required refinery product mix.[10]

[8] U. S. Department of Energy, Energy Information Administration, "Annual Energy Outlook 2001 with Projections to 2020", December 2000, www.eia.doe.gov/oiaf/aeo.

[9] Oil and Gas Journal, page 6, Sept 10, 2001, and Houston Chronicle, page 7D, June 15, 2003.

[10] Marano, J. J., Rogers, S., Choi, G. N., and Kramer, S. J., "Product Valuation of Fischer-Tropsch Derived Fuels," ACS National Meeting, Washington, D. C., August 21-6, 1994.

Interest rates in the United States are the lowest that have been in over 40 years. Electricity deregulation is occurring and its effect on the utility market is unknown. The *Annual Energy Outlook 2001* shows a current industrial power price of about 40 $/kW-hr and an average residential power price of about 84 $/kW/hr with the average to all users being about 60 $/kW-hr. Furthermore, over the next 20 years the Energy Information Administration predicts a 0.5%/year decrease in power prices (on a current dollar basis). This study inflated the cost of electricity at 1.7%/year which is 2.3% less than the general inflation rate. On a constant dollar basis this is a 0.6% annual decrease. Thus, the economic projections used in the study may be slightly conservative.

The following assumed economic and financial environment was reasonable when this study started, it may not be suitable for evaluating a specific project. Each project should be evaluated using a project specific economic scenario that is appropriate for its situation. For example, one project developer may place a low value on the coke, and another may place a high value on it because they have a well developed local market.

Table II.1
Basic Economic Parameters

Feeds	Price	Inflation, %/yr
Petroleum Coke, $/ton	0 $/ton	1.2
Coal	22.0 $/ton	1.2
Flux, $/ton	5.0 $/ton	1.7
Natural Gas, HHV	2.6 $/MMBtu	3.9
Products		
Electric Power	Calculated*	1.7
Hydrogen	1.3 $/Mscf	3.1
Fischer-Tropsch Liquids	30.0 $/bbl	3.1
Steam	5.6 $/ton	3.1
Fuel Gas	2.6 $/MMBtu	3.9
Sulfur	30.0 $/ton	0
Slag	0 $/ton	0
Other Financial Parameters		
General Inflation	2.3 %/year	
Loan Amount	80%	
Loan Interest Rate	10 %/year	
Loan Financing Fee	3%	
Owner's Contingency	5 % of EPC cost	
Development Fee	1.2 % of EPC cost	
Start-up Cost	1.5 % of EPC Cost	
Income Tax Rate	40%	

* Electric power prices are calculated to yield a given return on investment. They are reported on a current day cost; i.e., the cost at the time when construction begins.

II.7 Financial Analysis

For all cases a financial analysis was performed using a discounted cash flow (DCF) model that was developed by Bechtel Technology and Consulting (now Nexant Inc.) for the DOE as part of the Integrated Gasification Combined Cycle (IGCC) Economic and Capital Budgeting Practices Task.[11] This model calculates investment decision criteria used by industrial end-users and project developers to evaluate the economic feasibility of IGCC projects.

The required input information to the DCF financial model is organized into two distinct input areas that are called the Plant Input Sheet and the Scenario Input Sheet. The Plant Input Sheet contains data that are directly related to the specific plant as follows.

Data Contained on the Plant Input Sheet

- Project summary information
- Plant output and operating data
- Capital costs
- Operating costs and expenses
- Contingency, fees, owners cost, and start up expenses.

The Scenario Input Sheet contains data that are related to the general economic environment that is associated with the plant as well as some data that are plant related. The data on the Scenario Input Sheet are shown below.

Data Contained on the Scenario Input Sheet

- Financial and economic data
- Fuel data
- Tariff assumptions
- Construction schedule data
- Startup information

For all cases, the EPC spending pattern was adjusted to reflect forward escalation during the construction period since the EPC cost estimate is an "overnight" cost estimate based on mid-year 2000 costs.

Finally, items that were excluded in the cost estimate, such as spares, owners cost, contingency, and risk are included in the financial analysis.

The appendices contain filled in data input sheets for the discounted cash flow financial model for most of the cases. However, in all cases, the operating and maintenance cost information has been omitted because it is considered proprietary and highly confidential.

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

Chapter III

Study Basis and Overview

III.1 Study Basis

Global Energy's experience in the design, construction, and operation of the Wabash River Coal Gasification Repowering Project is the primary input that forms the foundation or basis for this study.[1] This project involved the repowering of a 1953 steam turbine with a Global Energy gasifier and a General Electric MS 7001 FA gas turbine. The design, construction, cost and operational information from this commercial facility were the starting point from which the subsequent Task 1 designs were developed. The Task 2 studies extended this work to IGCC coproduction producing liquid fuel precursors.

III.2 Coproduction Optimization Philosophy

The starting point for optimization of the liquids coproduction scheme is the clean syngas after acid gas removal. Thus the upstream, optimized coke or coal gasification system, which is a large part of the plant cost, was assumed to be the same for these liquid fuel coproduction studies. Modifications were made only as required for integration with the liquid fuels production area. The coproduction system optimization focused on dividing the clean syngas between the production of F-T liquids and power generation. Comparison of Subtasks 2.1 and 2.2 showed that maximizing F-T liquids production at the expense of power production improved the Return on Investment (ROI). However, power generation efficiency should be sub-optimized to maximize the revenue and ROI within the combined cycle power block. Ideally, to maximize F-T liquids and combined cycle ROI, the design should approach a once-through F-T system where all of the F-T offgas is consumed in a single large (maximum size) gas turbine/combined cycle power generation block. The F-T offgas is an ideal gas turbine fuel because it contains mostly methane and carbon dioxide along with unconverted carbon monoxide, hydrogen, and byproduct C2+. This offgas is a low BTU fuel gas which when used in the gas turbine requires minimal steam injection for NOx control. Also, the HRSG section of the combined cycle plant should recover sufficient energy to provide hot BFW to the gasification island and to superheat the steam from the gasification block and the F-T liquids synthesis area. Subtask 2.2 is close to this ideal case in that over 92% of the syngas production goes to the F-T synthesis area with the remaining 8% going directly to the turbine. In comparison, Subtask 2.3 approaches this ideal case, but 18% of the available syngas has to bypass the F-T area and go directly to the gas turbine. (A larger gas turbine consuming 40% more syngas would allow the coal case to approach ideal optimization.) Trying to use a single GE 7FAe+ gas turbine with some duct firing to superheat the process steam slightly improves the ROI at low power prices (changes the slope of the ROI versus the power price curve).

The results of each of the Task 2 subtasks are described in detail in separate appendices. Table III.1 summarizes the results of the three Subtask 2 gasification power plants with liquid fuels coproduction and some Task 1 plants. The Task 1 plants are given to provide a reference for measuring the performance of the Task 2 plants. This table is presented here

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

to provide an overview of the cases and to be used as a reference for the following chapters.

III.3 Heat Integration

Integrated Gasification Combined Cycle (IGCC) or IGCC with coproduction (IGCP), as the name implies, is the integration of two primary process blocks, gasification and combined cycle power generation. Integration refers to the sharing of energy between the various process blocks. The optimum use of heat has been extensively studied.[2] Figure III.1 shows the overall input streams, output streams, and integration streams between the gasification block, hydrogen production facilities, and the combined cycle power block for the Subtask 1.3 Next Plant. Figure III.2 only shows the interconnecting energy streams between the gasification block, combined cycle power block and the F-T hydrocarbon synthesis area. As shown, the heat integration scheme for the Subtask 2.2 plant, although similar, includes additional streams involving the hydrocarbon synthesis area. It is the efficiency of the individual pieces and the sharing of energy between the pieces that determines the overall plant output and thermal efficiency. From the overall energy balance and the information in the individual subtask reports (Appendices A through C), it can be shown that most of the fuel (coal or coke) energy is used to make power and F-T liquids. There also is a significant amount of energy that is recovered as medium and high-pressure steam, which after superheating, is used for power production. Most of the low level energy is used effectively for syngas moisturization, to heat the syngas going to F-T synthesis reactor(s), or to heat the F-T offgas going to the combustion turbine. Very little low level energy is recovered in the steam bottoming cycle or is rejected to the cooling water system.

Global Energy's two-stage gasifier at Wabash River has a relatively high cold gas efficiency of above 77% when operating on either subituminous coal or petroleum coke. Carbon conversion efficiency is about 99%. When combined with high temperature heat recovery, heat integration, and steam extraction for process and gas turbine diluent use, high plant thermal efficiencies of 40% or greater can be achieved. As shown in Table III.1, the coproduction of F-T liquids increases the overall thermal efficiency of the plant.

The inclusion of a Fischer-Tropsch liquid fuels coproduction section in the plant introduces other options for heat integration between it and the rest of the facility. The F-T slurry-bed reactor(s) generate large amounts of 440º/375 psia steam within coils inside the reactor. Most of this steam goes to the power block where it is superheated in the HRSG and used to generate power in the steam turbine. Higher temperature and higher pressure steam from either the gasification and/or power blocks can be used for heating in the in the F-T area. Low-pressure steam from the power block is used in the F-T area for regenerating the activated carbon beds that are used to remove the residual sulfur from the syngas going to the F-T reactor.

Because of the various stream interactions between the different sections in the plant, there are numerous opportunities for improving the heat integration and to increase the thermal efficiency. The Value Improving Practices exercise generated numerous Value Engineering ideas in this area. However, the objective of this study was to lower the cost of electricity and not to design plants with the highest thermal efficiency at any cost. Thus, economic

[2] Geosits, R. F. and Y. Mohammad-zadeh, "Optimization of Air and Heat Integration for IGCC Plants", presented at Power-Gen Americas '95, Anaheim, CA, December 7, 1995.

viability provided the criteria for incorporating improvements. Depending upon the relative costs of fuel, products and equipment, the optimal plant thermal efficiency can change. For example, a plant using a low cost feedstock, such as the Subtask 2.2 coke fueled plant, may have a better return on investment at a lower thermal efficiency than one that uses a higher priced coal feedstock, such as the Subtask 2.3 plant.

Global Energy's gasification technology appears to have some design flexibility (e.g., the Wabash River design vs. full slurry quench (FSQ) vs. full slurry vaporization (FSV)). In the Wabash River design, temperature control at the second stage outlet is maintained by injection of cooler syngas. With full slurry quench, the slurry feed is distributed between the first and second stages with the amount entering the second stage being manipulated to control the second stage outlet temperature. Wabash River is moving to this type of operation. With full slurry feed vaporization, the temperature control criterion is eliminated and all the fresh feed enters the second stage. Slurry feed vaporization theoretically provides the maximum conversion of feed to chemical energy and the lowest oxygen demand (tons of O_2 per ton of feed), resulting in the highest cold gas efficiency. It also produces more methane in place of CO and H_2. This design was evaluated in Subtask 1.4 along with the next generation advanced "G/H" combustion turbine. As shown in the Task 1 Topical Report, these changes dramaticly improve plant ROI. However these advancements are not likely to occur in the near future, and therefore, were not considered for Task 2.

Fuel cost per unit of production is inversely proportional to the efficiency except for the coke cases in this study where the coke is assumed to have a net zero cost. More importantly, increasing the cold gas efficiency will shift energy to the F-T and combined cycle sections which hopefully will increase the liquids production and power output (and efficiency).

III.4 Cost Drivers

The primary objective of this study was to reduce the cost of power and liquid hydrocarbons from gasification based power plants and/or increase their return on investment. The following items were identified as the most important cost drivers.

1. Total Installed Cost
2. Plant and/or Train Size
3. Product Slate (Power vs. Coproduction)
4. Revenue Generating Capacity (Availability)
5. Operating and Maintenance Costs
6. Economic and Financial Environment
7. Project Specific Requirements

The plant designers can have an influence over the first five of the above cost drivers within technological limits. The sixth and seventh items are the ballpark in which the designers must work. The financial environment is ever changing. For example, a natural gas fired combined cycle power plant will look good when natural gas prices are low, but when they are high, many gas fired power plants may have to be shut down if possible because the revenue generated by their power sales is less that the cost of the natural gas used to produce it. The same situation also can exist for the liquid hydrocarbon product. Over the past couple of years, crude oil prices have ranged between 10 and 30 $/bbl which probably would cause a similar variation in the liquid product values. For these reasons, any contemplated project should be evaluated under the present and various likely future economic environments to determine if it is viable.

The total installed cost is the predominant cost driver over which the plant designer has the most control. For this reason, this study concentrated on reducing the plant cost. The Value Improving Practices procedures that were used in this study of Process Simplification (PS), Classes of Plant Quality (CPQ), Design-to-Capacity (DTC), Plant Layout Optimization (PL), Constructability Reviews (C), and Technology Selection (TS) all are related to reducing the total installed cost of the plant. Task 1 applied these procedures which resulted in the

1. Elimination of the redundant and/or duplicate equipment, such as unnecessary spare pumps (PS)
2. Reduction in the size of equipment by eliminating spare capacity or extra capacity for possible expansion (DTC)
3. Removal of things that would be “nice to have” but are not required (CPQ)
4. Deleting unnecessary flexibility by removing extra capacity in some plant sections in case a different feedstock may be used (CPQ)
5. Shrinkage in the plant site without sacrificing accessibility during construction or for maintenance to save piping and site preparation costs (PL)
6. Selection of the most cost effective technology (TS)
7. Improved scheduling for shorter construction times (C)
8. Increased output or increased efficiency

The main focus of the above VIPs was cost reduction and optimization with considerations given to the costs of cold gas efficiency improvements and additional heat recovery.

By application of the above procedures, Task 1 and Task 2 achieved significant cost reductions, and it is expected that more cost reductions will be achieved in the future.

Cost reductions per unit of material processed can be achieved by using larger train sizes until the maximum size of a critical (or expensive) piece of equipment is reached. Generally equipment costs increase by the 0.6 to 0.7 power of the capacity. This means that the plant cost on a unit of material processed basis decreases as the plant size increases; i. e., the economies of scale effect. Because of this, the larger 210 MW GE 7FA+e combustion turbine, that was used in Subtasks 1.2 though 1.6, also was used in the Task 2 plants.

A plant that is shut down is not producing any revenue. Therefore, care was taken in the plant designs to minimize the amount of scheduled downtime, to increase reliability, and to facilitate maintenance access. Availability analyses based on operating data from the Wabash River Repowering Project which were used to predict the availability of the plant designs. Thus, the Subtask 2.1 and 2.2 coke power plants with liquid fuels coproduction are based on the Subtask 1.3 Next Plant which contains a spare gasification train to achieve high syngas availability.

Any operating and maintenance (O&M) cost reductions fall directly to the bottom line. Although the specific details are considered proprietary, Global Energy personnel were included as part of the VIP team to develop and examine specific ideas for reducing the O&M costs of any new facility. If they were economic, the design changes were implemented, as required, to generate long term O&M savings. As a result of this effort, significant O&M savings based on Wabash River operations were achieved.

III.5 Plant Size

For IGCC plants, the capital cost is the largest component of the electricity cost. Table 13 on page 75 of the EIA *Annual Energy Outlook 2001* estimates the cost of producing electricity from an advanced coal plant of conventional design with a 36.9% thermal efficiency at 43.2 $/MW-hr.[3] About 72% of this cost is attributable to the capital cost of the plant, about 18% to the fuel cost, and about 10% to the operating and maintenance costs. This clearly shows that the plant cost is the dominant factor, and must be decreased in order to significantly reduce the cost of electricity. At the moment, IGCC plants are more expensive on a per unit of export power than conventional pulverized coal power plants, but they have a higher efficiency and very low emissions. Thus, the capital cost component of the electricity cost is larger for IGCC plants.

As noted above, the cost of production decreases as the plant size increases. The general relationship between capacity and plant cost is that the plant cost increases with the capacity raised to the 0.6 to 0.7 power. This relationship holds until the maximum size of a critical or expensive piece of equipment is reached, and any further capacity increases only can be achieved by replicating that piece of equipment.

The costs of utilities and off site facilities also follow the same exponential relationship. The cost of production from multiple train plants also is lower than that from single train plants because the costs of the utilities and offsite facilities can be shared between trains. However, the reduction is not as great because the utilities and offsite facilities are not major components of the plant cost.

Based on the above logic, the current gasifier capacity could be expanded by 40% to 50% to take advantage of the economies of scale, whenever appropriate. Global Energy believes this can be accomplished with their current design. The Subtask 1.3 Next Plant and Subtask 2.2 liquids coproduction plant each use two air separation units with capacities just just under 3,000 tpd. New plants can have capacities of 3,500 tpd and above. For Subtasks 1.6 and 2.3, larger gasifiers only are required if spare capacity is desired to allow the facility to produce design output with only 3 of 4 gasifiers operating.

III.6 Study Perceptions and Strategic Marketing Considerations

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

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

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[3] U. S. Department of Energy, Energy Information Administration, "Annual Energy Outlook 2001 with Projections to 2020," December 2000, www.eia.doe.gov/oiaf/aeo.

Precision – Using cost information from the as-built Wabash River facility and Bechtel's Power Line™ plants allowed the cost estimates to have a high degree of confidence or, expressed differently, a minimum amount of uncertainity.

Potential – This study addresses the potential of Global Energy's gasification technology along with coproduction of F-T liquids to reduce the cost and improve the efficiency of IGCC plants. Further cost savings ideas are under investigation.

Price – The above mentioned cost savings significantly reduced the cost of electricity to the point where under certain situations IGCC is competitive.

Product (or Market Penetration) – Currently coke fueled IGCC plants have the advantage over coal fueled ones because of the lower feedstock cost. The initial application of coke IGCC plants will further develop IGCC technology leading to improved designs, reduced costs, and increased efficiencies.

Place (Location) – The U. S. Gulf coast location, especially if it is on a waterway, seems to be the best location for coke fueled IGCC plants because it is likely close to the source of the refineries that produce the coke. A coke coproduction plant should be located adjacent to a petroleum refinery to minimize transportation costs and allow sharing of support facilities.

Proliferation - As more IGCC plants are built using either coke and coal. Their costs will decrease leading to the construction of additional IGCC plants.

Preferred Design – The Subtask 2.2 F-T liquids coproduction plant is the preferred design for a coke IGCC coproduction plant. It includes a two-stage dry particulate removal system. However, during the study wet particulate filtration tests showed better than expected results. Therefore, Global Energy also is considering pursuing the development of a wet filtration system to determine if additional cost savings are possible. In any case, as capital costs continue to decrease and fuel prices (especially natural gas prices) increase, large coal fueled IGCC facilities, similar to the Subtask 1.6 case, should become the preferred design for coal power plants because of their higher ROI and lower emissions.

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 IGCC by demonstrating that starting with the Wabash River design and applying VIP optimization techniques, it is possible to build a low cost IGCC and F-T liquids coproduction plants 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 adequately quantify future revenues.

Table III.1

Task 1 and 2 Coal and Coke IGCC Case Summaries

Case Description	Subtask 1.1 Wabash River Greenfield	Subtask 1.3 Next Optimized Pet Coke IGCC Coproduction Plant	Subtask 1.5 Single Train Power 1.5A Coal	Subtask 1.5 Single Train Power 1.5B Coke	Subtask 1.6 1,000 MW Coal IGCC Power Plant	Subtask 2.1 Petroleum Coke to Liquids and Power	Subtask 2.2 Optimized Pet Coke to Liquids and Power	Subtask 2.3 Optimized Coal to Liquids and Power
Configuration								
Plant Location	Midwest	Gulf Coast	Gulf Coast	Gulf Coast	Midwest	Gulf Coast	Gulf Coast	Midwest
Number of Air Separation Units	1	2	1	1	3	2	2	3
Number of Gas Turbines	1	2	1	1	4	2	1	2
Number of Gasification Trains	1	3	1	1	4	3	3	4
Number of Gasification Vessels	2	3	2	2	4	3	3	4
No of Syngas Processing Trains	1	2	1	1	2	2	2	2
Number of 50% H2 trains	0	2	0	0	0	0	0	0
Number of F-T Liquid Trains	0	0	0	0	0	1	1	1
Design Feed Rates								
Feedstock Type	Coal	Pet Coke	Coal	Pet Coke	Coal	Pet Coke	Pet Coke	Coal
Coal or Coke, TPD as received	2,642	5,692	2,754	2,077	10,837	5,649	5,684	10,837
Coal or Coke, TPD dry	2,259	5,417	2,355	1,977	9,266	5,376	5,417	9,266
Feed, MMBtu HHV/hr	2,400	6,703	2,481	2,446	9,844	6,652	6,703	9,844
Feed, MMBtu LHV/hr	2,311	6,567	2,389	2,397	9,478	6,518	6,567	9,478
Flux, TPD	0	110.6	0	40.3	0	109.7	110.6	0
Water, gpm	2,790	5,223	2,840	2,525	9,752	6,472	5,693	7,403
Condensate, Mlb/hr	---	686	---	---	---	---	---	---
Oxygen, TPD of 95% O2	2,130	5,954	2,015	2,143	8,009	5,919	5,877	7,919
Oxygen, TPD of O2	2,009	5,615	1,900	2,021	7,553	5,582	5,542	7,468
Design Product Rates								
Electric Power, MW	269.3	474.0	284.6	291.3	1,154.6	617.0	366.9	675.9
Steam (750^{o}F/700 psig), lb/hr	---	980.0	---	---	---	---	---	---
Hydrogen, MMscfd	---	80.0	---	---	---	---	---	---
Sulfur, TPD	57	373	60	136	237	371	373	237
Slag (@ 15% water), TPD	356	195	364	71	1,423	194	195	1,423
Fuel Gas, MMBtu HHV/hr	---	0	---	---	---	---	---	---
Solid Waste to Disposal, TPD (4)	---	---	---	---	---	0.95	1.31	1.72
Liquid Hydrocarbons, bpd	---	---	---	---	---	4,125	10,450	12,377
Gas Turbine								
Type	GE 7FA	GE 7FA+e	GE 7FA+e	GE 7FA+e	GE 7FA+e	GE 7FA+e	GE 7FA+e	GE 7FA+e
Fuel Input, Mlb/hr	411.4	1,016.8	447.0	426.7	1,741.6	1,092.8	1,000.8 (5)	1,303.0
Heat Input, MMBtu/hr LHV	1,675	3,592	1,796	1,796	7,184	3590	1,763.3	3,532
Steam Injection, Mlb/hr	111.0	395.7	246.8	272.3	1,037.8	531.6	0	510.5
Gross Power Output, MW	192	420	210	210	840	420	199.4	416
Cold Gas Efficiency (HHV), %	76.9	77.5	77.8	77.4	78.0	77.5	77.7	78.3
Steam Turbine Power, MW	118	164.3	113	121	465.2	307.0	274.9	403.6
Internal Power Use, MW	41	110	38.4	40.7	151	110.0	107.4	118.8
Heat Rate, HHV Btu/kW-hr	8,912	NA	8,717	8,397	8,526	NA	NA	NA
Thermal Efficiency, % HHV (1)	38.3	NA	39.1	40.6	40.0	46.0	54.9	52.6
Emissions								
SOx as SO2, lb/hr	312	350	142	119	438	321	276	329
NOx as NO2, lb/hr	161	166	69	69	275	136	94	166
CO, lb/hr	49	89	33	34	131	66	37	65
Sulfur Removal, %	96.7	99.4	98.5	99.4	98.9	99.5	99.6	100
Performance Parameters								
Tons 02 / Ton of Dry Feed	0.889	1.037	0.807	1.022	0.815	1.038	1.023	0.806
Gross MW / Ton of Dry Feed	0.137	0.108	0.137	0.168	0.141	0.135	0.088	0.088
Net MW / Ton of Dry Feed	0.119	0.088	0.121	0.147	0.125	0.115	0.068	0.073
Emissions								
SOx (SO2) as lb/MW-hr	1.159	0.738	0.499	0.409	0.379	0.520	0.752	0.487
SOx (SO2) as lb/MMBtu (HHV)	0.130	0.052	0.057	0.049	0.044	0.048	0.041	0.033
NOx (NO2) as lb/MW-hr	0.598	0.350	0.242	0.237	0.238	0.220	0.256	0.246
NOx (NO2) as lb/MMBtu (HHV)	0.067	0.025	0.028	0.028	0.028	0.020	0.014	0.017
CO, lb/MW-hr	0.182	0.188	0.116	0.117	0.113	0.107	0.101	0.096
CO, lb/MMBtu (HHV)	0.020	0.013	0.013	0.014	0.013	0.010	0.006	0.007
Daily Average Feed/Product Rates with Backup Natural Gas (Subtask 1.1 is without Backup Natural Gas. Subtask 2.2 purchases power.)								
Coal or Coke, TPD dry	1,705	4,842	1,826	1,546	7,018	4,805	4,984	6,929
Coal or Coke, % of design	75.5%	89.4%	77.5%	78.2%	75.7%	89.4%	92.0%	74.8%
Power, MW	203.2	448.4	264.4	269.4	1,081	572.5	316.4	613.7
Power, % of design	75.5%	94.6%	92.9%	92.5%	93.6%	92.8%	86.2%	90.8%
Steam, lbs/hr	---	974.6	---	---	---	---	---	---
Steam, % of design	---	99.4%	---	---	---	---	---	---
Hydrogen, MMscfd	---	78.8	---	---	---	---	---	---
Hydrogen, % of design	---	99.4%	---	---	---	---	---	---
Fuel Gas, MMBtu HHV/hr	---	0	---	---	---	---	---	---
Fuel Gas, % of design	---	---	---	---	---	---	---	---
Natural Gas, Mscfd	NA	9,059	6,929	6,929	34,960	8,856	0	26,466
Liquid Hydrocarbons, bpd	---	---	---	---	---	3,938	9,702	10,397
Liquid Hydrocarbons, % of design	---	---	---	---	---	95.5%	92.8%	84.0%
Plant Cost, MM mid-2000 $ (2)	452.6	787.3	375.0	367.0	1,231.3	817.9	735.3	1159.1
Plant Cost, $/design kW	1,681	NA	1,318	1,260	1,066	NA	NA	NA
Required Electricity Selling Price for a 12% ROI, $/MW-hr (3)								
Without Natural Gas Backup	67.5	---	53.9	43.9	44.4	28.8	19.5	48.1
With Natural Gas Backup	---	30.0	48.9	40.6	40.2	29.0	17.7	42.0

NA = Not Applicable
July 31, 2003

1. Without including the sulfur byproduct, but including the F-T liquid fuels, when appropriate.
2. All costs are mid-year 2000 EPC costs which exclude contingency, taxes, fees and owners costs. They are presented here to show the relative differences between cases. Current cost estimates should be developed for any proposed applications.
3. Power selling prices are presented to show a relative comparison between cases. Based on a natural gas price of $2.60 $/MMBtu and a liquids price of 30 $/bbl. Subtask 2.2 purshases power at the power selling price rather than natural gas.
4. Used COS hydrolysis catalyst, Used ZnO sulfur sorbent, and used F-T catalyst, all on a dry, hydrocarbon free basis. The used activated carbon in Subtasks 2.2 and 2.3 is mixed with the gasifier feed and converted to syngas and slag.
5. Includes 57.8 Mlbhr of steam is added to the fuel to get a net heating value of 147.1 Btu/scf. No additional steam is needed for NOx control.

Figure III.1

Interconnecting Streams for the Subtask 1.3 Next Plant

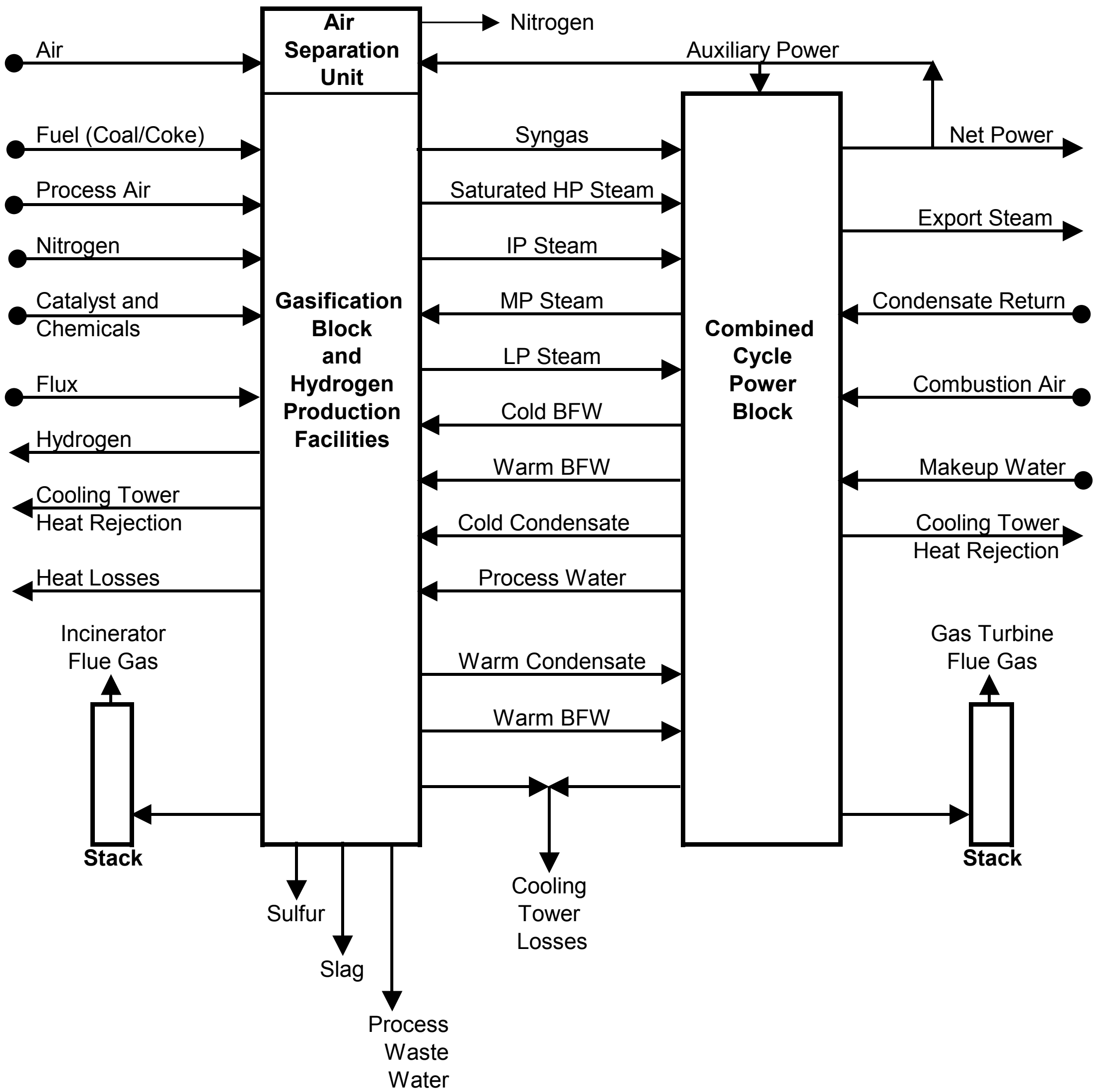
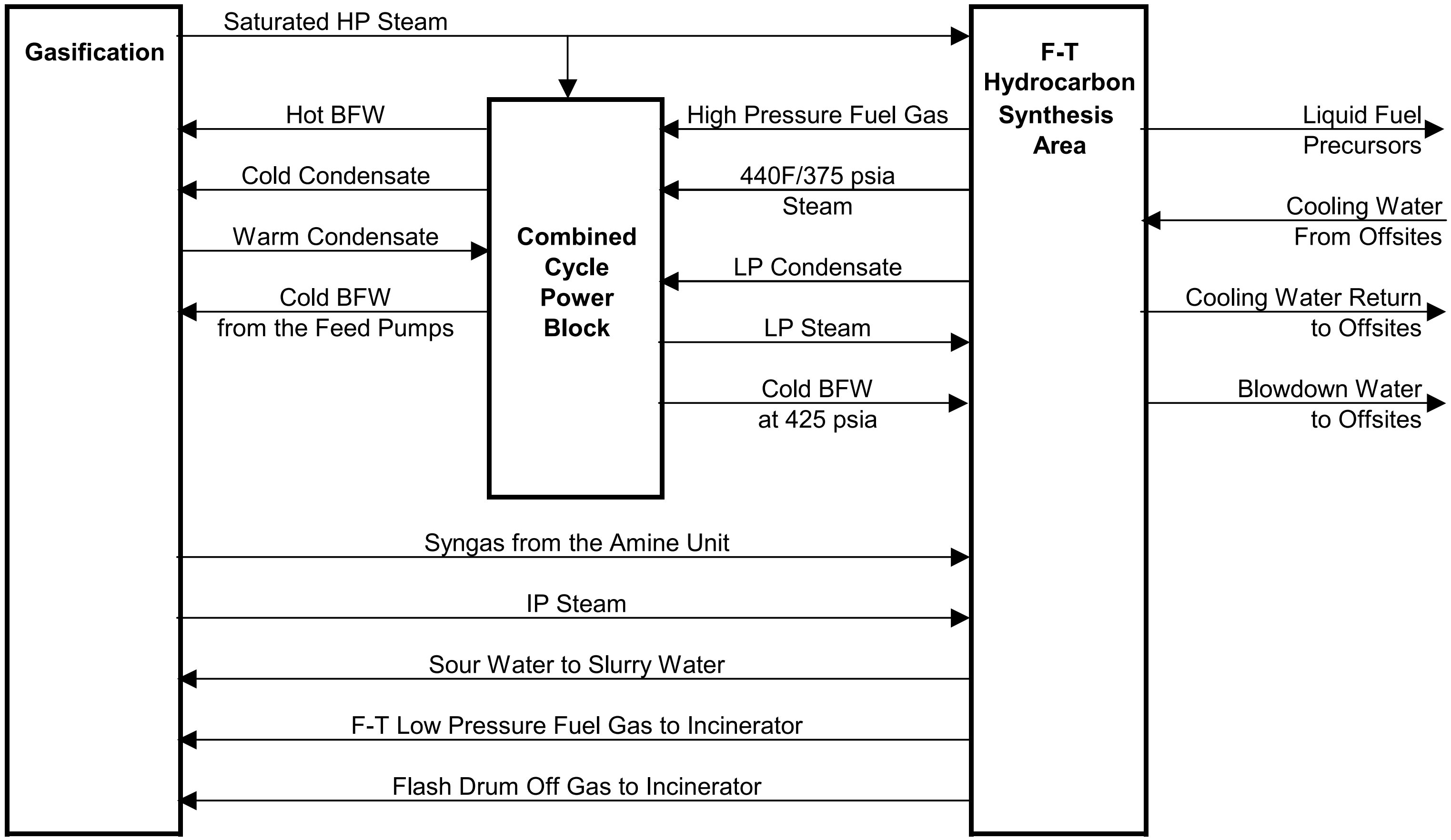


Figure III.2

Interconnecting Energy Streams Between the Gasification Block, Power Block and the F-T Synthesis Area



Chapter IV

Petroleum Coke Cases

IV.1 Introduction

The designs for two petroleum coke IGCC gasification power plants with liquid fuels coproduction were developed in Task 2. The Subtask 2.1 non-optimized plant design is based on the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. In the Subtask 2.1 plant, the F-T hydrocarbon synthesis area essentially replaced the hydrogen production area of the Subtask 1.3 plant, and the steam that previously was exported now is used for power production. This facility essentially is a power plant with a small amount of liquid fuels coproduction.

The Subtask 2.2 Optimized Petroleum Coke Gasification Power Plant with Liquids Coproduction maximized the liquid fuels production at the expense of power production. As a result, the liquid fuels production increased to 10,450 bpd from 4,125 bpd, and the export power decreased from 617 MW to 367 MW.

IV.2 Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

The Subtask 1.3 plant design was optimized by applying nine Value Improving Practices (VIPs) to the Subtask 1.2 non-optimized petroleum coke IGCC coproduction plant.[1] As a result of this effort, plant performance was improved, the plant cost was reduced, and the return on investment was significantly improved. The results of this VIP and optimization study included:

- Simplified solids handling system
- Removal of the feed heaters and spare pumps
- Maximum use of slurry quench
- Maximum syngas moisturization
- Use of a cyclone and a dry particulate removal system to clean the syngas
- Removal of the T-120 post reactor residence vessel
- Simplified Claus plant, amine and sour water stripper
- Use of state-of-the-art GE 7FA+e gas turbines with 210 MW output and lower NOx
- Use of steam diluent in the gas turbines
- Development of a compact plant layout to minimize the use of large bore piping
- Used Bechtel's advanced construction techniques to reduce costs
- Added design features to reduce O&M costs and increase syngas availability

Table IV.1 shows the design input and output streams for the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. The plant processes 5,417 tpd of dry petroleum coke and produces 474 MW of export power. In addition, the plant exports 980,000 lb/hr of 750ºF/700 psia steam and 80 MMscfd of hydrogen to the adjacent petroleum refinery. It also produces 373.4 tpd of sulfur and 195.1 tpd of slag. No natural gas is consumed during

[1] "Task 1 Topical Report – IGCC Plant Cost Optimization", Gasification Plant Cost and Performance Optimization, Chapter II, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, May 30, 2002

Table IV.1

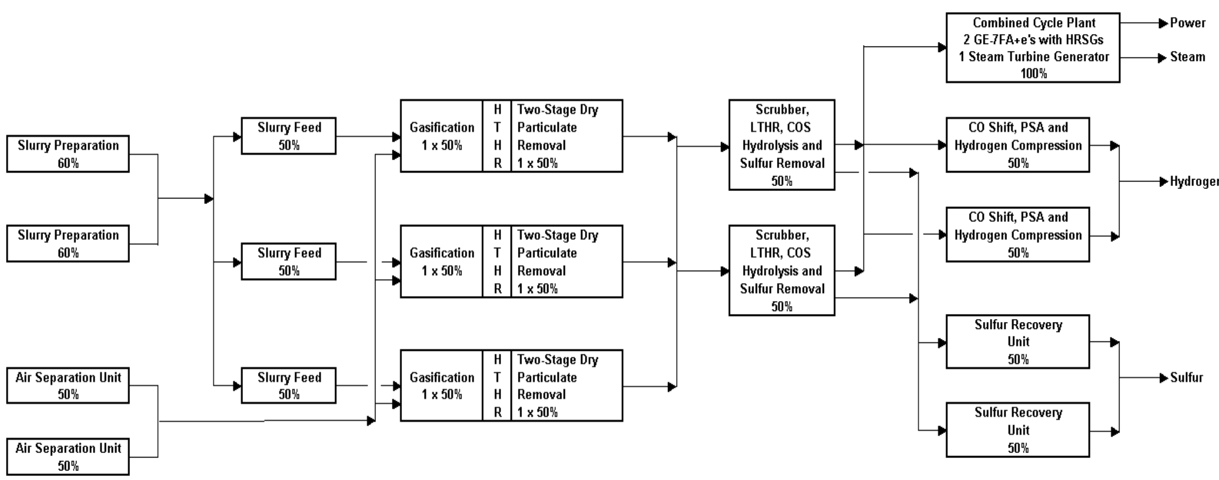
Design Input and Output Streams for the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

	Subtask 1.3 Next Plant
Plant Inputs	
Coke Feed, as received TPD	5,692
Dry Coke Feed to Gasifiers, TPD	5,417
Oxygen Production, TPD of 95% O_2	5,954
Total Fresh Water Consumption, gpm	5,223
Condensate Return from the Refinery, lb/hr	686,000
Flux, TPD	110.6
Natural Gas, MMBtu HHV/hr	0
Plant Outputs	
Net Power Output, MW	474.0
Sulfur, TPD	373.4
Slag, TPD (15% moisture)	195.1
Hydrogen, MMscfd	80
HP Steam, 750°F/700 psia	980,000
Fuel Gas Export, MMBtu/hr	0

Figure IV.1

Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

Simplified Block Train Diagram



Notes: Capacity percentages are based on total plant capacity.

design operations. However, the plant does use natural gas during startup and as a supplementary fuel to fire the combustion turbines when insufficient syngas is available.

On a lower heating value (LHV) basis, the plant has a thermal efficiency 40.5% when the heating value of the byproduct sulfur is included and 38.6% when the byproduct sulfur is not included. On a higher heating value (HHV) basis, the plant has a thermal efficiency 42.1% when the heating value of the byproduct sulfur is included and 40.3% when the byproduct sulfur is not included. These thermal efficiencies include the heating value of the hydrogen byproduct. However, they completely ignore any contribution from the 980,000 lb/hr of the 750ºF/700 psig export steam.

Figure IV.1 is a schematic block train diagram of the Subtask 1.3 Next Plant. The plant basically is a two train facility with a complete spare gasification train. No gasification trains contain a spare gasification vessel. There also are two sulfur recovery trains, two sulfur production trains, and two hydrogen production trains which are sized so that they only have sufficient capacity to process the output from two gasifiers simultaneously operating at design capacity. The combined cycle power block contains two General Electric 7FAe+ combustion turbines, each one with a dedicated HRSG, and a single steam generator.

IV.3 Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction

Starting from the Subtask 1.3 Next Plant, the Subtask 2.1 plant was developed by eliminating the export steam production and hydrogen production facilities and replacing them with a single-train, once-through Fischer-Tropsch hydrocarbon synthesis plant. A once-through system eliminates the cost of the expensive recycle system which requires recycle gas purification facilities in addition to the recycle compressor. The energy that was used to produce the export steam now is used to generate additional power. Even with almost the same coke feed rate to the gasifiers, the Subtask 2.1 process changes required adjustments to the steam and water flows both in and between the gasification block and the power generation block in order to effectively balance the systems.

Table IV.2 compares the design input and output stream flows for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. The Subtask 2.1 plant processes slightly less petroleum coke (5,376 vs. 5,417 dry tpd) than the Subtask 1.3 Next Plant. It also has a higher fresh water consumption of 6,472 gpm vs. 5,223 gpm. Furthermore, it consumes a small amount of natural gas, 23.2 MMBtu HHV/hr. Because the Subtask 2.1 plant does not export any hydrogen or steam, it produces more export power than the previous case (617 MW vs. 474 MW) in addition to 4,125 bpd of liquid fuel precursors from the F-T area.

On a lower heating value (LHV) basis, the plant has a thermal efficiency 47.8% when the heating value of the byproduct sulfur is included and 45.9% when the byproduct sulfur is not included. On a higher heating value (HHV) basis, the plant has a thermal efficiency 47.9% when the heating value of the byproduct sulfur is included and 46.0% when the byproduct sulfur is not included. These thermal efficiencies are higher than those that would be obtained from a coke IGCC power plant of a similar design because it includes the heating value of the liquid fuel that is produced. Since the second law of thermodynamics states this liquid fuel cannot be used at a 100% thermal efficiency, the thermal efficiency of the plant will be somewhat lower when the final disposition of the liquid fuel is considered.

Table IV.2

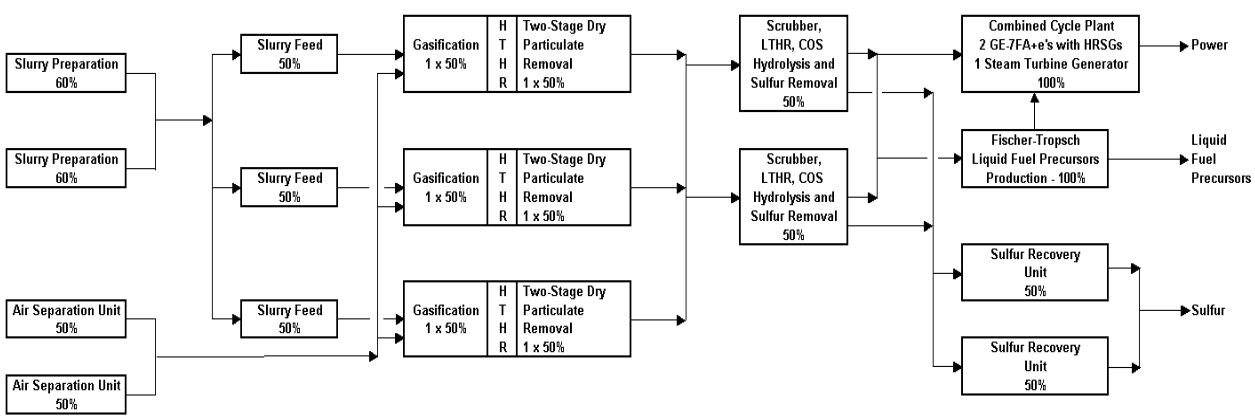
Design Input and Output Streams for the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant and the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction

	Subtask 1.3 Next Plant	Subtask 2.1 Power and Liquids Plant
Plant Inputs		
Coke Feed, as received TPD	5,692	5,649
Dry Coke Feed to Gasifiers, TPD	5,417	5,376
Oxygen Production, TPD of 95% O_2	5,954	5,919
Total Fresh Water Consumption, gpm	5,223	6,472
Condensate Return from the Refinery, lb/hr	686,000	0
Flux, TPD	110.6	109.7
Natural Gas, MMBtu HHV/hr	0	23.2
Plant Outputs		
Net Power Output, MW	474.0	617.0
Sulfur, TPD	373.4	370.6
Slag, TPD (15% moisture)	195.1	193.6
Hydrogen, MMscfd	80	0
HP Steam, 750^oF/700 psia	980,000	0
Liquid Fuel Precursors, bpd	0	4,125

Figure IV.2

Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction

Simplified Block Train Diagram

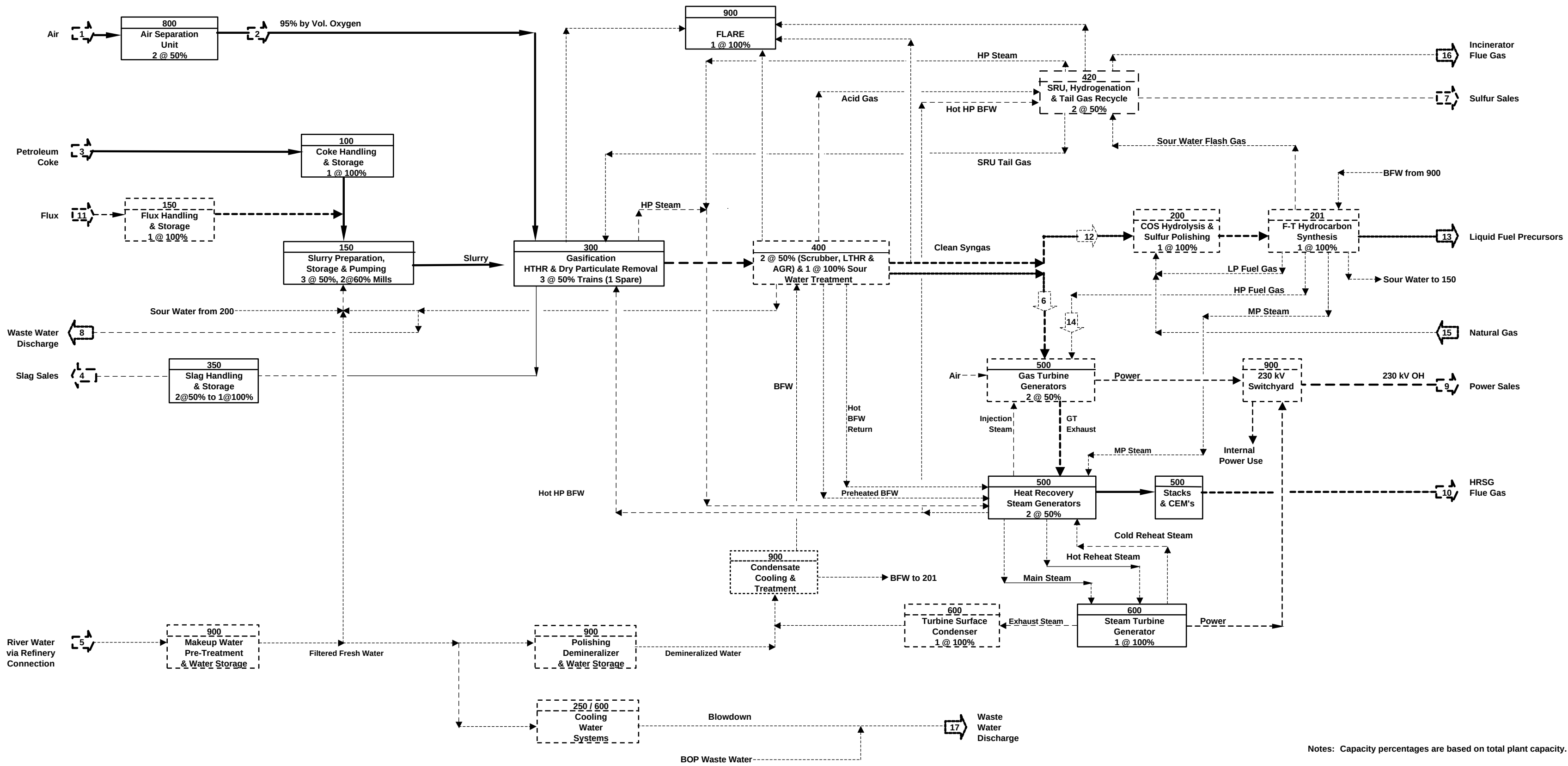


Notes: Capacity percentages are based on total plant capacity.

Figure IV.3

Block Flow Diagram of the Subtask 2.1 Coke

Gasification Power Plant with Liquids Coproduction



Notes: Capacity percentages are based on total plant capacity.

	1	2	3	4	5	6	7	8	9	10
Flow	Air 25,808 Tons/Day	Oxygen 5,919 Tons/Day	Coke 5,376 Tons/Day	Slag 193.6 Tons/Day	Water 3,236,000 Lb/Hr	Syngas 722,540 Lb/Hr	Sulfur 370.6 Tons/Day	Water 52,000 Lb/Hr	Power 617,000 kWe	Flue Gas 7,966,800 Lb/Hr
Nominal Pressure - psig	Atmos.	609	NA	NA	50	355	NA	62	NA	Atmos.
Temperature - F	70	240	Ambient	180	70	532	332	80	NA	265
HHV Btu/lb	NA	NA	14,848	NA	NA	4,325	3,983	NA	NA	NA
LHV Btu/lb	NA	NA	14,548	NA	NA	4,101	3,983	NA	NA	NA
Energy - MM HHV/hr	NA	NA	6,652	NA	NA	3,125	123	NA	NA	NA
Energy - MM LHV/hr	NA	NA	6,518	NA	NA	2,963	123	NA	NA	NA
Notes	Dry Basis	5,582 O2	Dry Basis	15%Wtr.	6,472 GPM	To GT	Sales	104 GPM	230 kV	

	11	12	13	14	15	16	17	18	19	20	21
Flow	Flux 109.7 Tons/Day	Syngas 403,502 Lb/Hr	Liq Fuel 48,897 Lb/hr	Fuel Gas 370,255 Lb/hr	Nat Gas 23.2 MMBtu/hr	Flue Gas 21,672 Lb/Hr	Water 711,000 Lb/Hr				
Nominal Pressure - psig	NA	360	50	365	50	Atmos.	Atmos.				
Temperature - F	NA	100	110	168	100	500	71				
HHV Btu/lb	NA	4,997	19,777	1,852	1,000	NA	NA				
LHV Btu/lb	NA	4,738	18,297	1,698	910	NA	NA				
Energy - MM HHV/hr	NA	2,016	967.0	685.6	23.2	NA	NA				
Energy - MM LHV/hr	NA	1,912	894.7	628.8	21.1	NA	NA				
Notes		For F-T	4,125 bpd				1,422 GPM				

DOE Gasification Plant Cost and Performance Optimization

Figure IV.3

Subtask 2.1

COKE GASIFICATION POWER PLANT

WITH LIQUID FUELS COPRODUCTION

BLOCK FLOW DIAGRAM

File: Fig IV.3.xls June 11, 2003

Figure IV.2 is a schematic block train diagram of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction. The syngas generation and clean up sections of the plant essentially are a two train facility, but it contains a complete spare gasification train. There also are two sulfur recovery trains and two sulfur production trains that are sized so that they only have sufficient capacity to process the output from two gasifiers simultaneously operating at design capacity. The combined cycle power block contains two General Electric 7FAe+ combustion turbines, each one with a dedicated HRSG, and a single steam generator. The F-T hydrocarbon synthesis section also is a single train facility.

Figure IV.3 is a simplified block flow diagram of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction. This plant can be considered to consist of three distinct main processing areas.

- The gasification island and air separation unit (Areas 100, 150, 250, 300, 350, 400, 420, and 800)
- The F-T hydrocarbon synthesis area (Areas 200 and 201)
- The power block (Areas 500 and 600)

In addition there is a balance of plant area (Area 900).

Appendix A contains a detailed description of the various processing blocks in the plant. All areas except the F-T hydrocarbon synthesis area (Areas 200 and 201) essentially are the same as the corresponding area in the Subtask 1.3 Next Plant.

The design for the Fischer-Tropsch Hydrocarbon Synthesis Area was developed based on the ASPEN Plus process flowsheet reactor model that was developed for the Baseline Design/Economics for Advanced Fischer-Tropsch Technology study.[2] That model was used to simulate the final syngas treating before the slurry-bed reactor, the F-T slurry-bed reactor system, and the cooling, separation, and recovery of the liquid product. About 35.8% of the syngas produced by the gasification block goes through the F-T area while the remaining 64.2% is sent directly to the power block.

The Fischer-Tropsch hydrocarbon synthesis area consists of two sub areas, Area 200 and Area 201. Area 200 is the Final Syngas Cleanup Area, which removes the final traces of sulfur from the syngas, before it is converted to hydrocarbons in Area 201, the Hydrocarbon Synthesis and Product Recovery Area.

The Final Syngas Cleanup Area, Area 200, reduces the sulfur concentration of the cleaned syngas from the acid gas removal area of the gasification block to less than 0.1 ppm of sulfur. This is accomplished by hydrolyzing the small amounts of carbonyl sulfide (COS) and trace amounts of other light organic sulfur compounds (such as CS_2) to hydrogen sulfide (H_2S), and removing the H_2S by reacting it with zinc oxide (ZnO) to produce solid zinc sulfide (ZnS) and water. The ZnO is permanently consumed, and the ZnS/ZnO mixture eventually is discarded.

[2] "Topical Report – Volume I, Process Design – Illinois No. 6 Coal Case with Conventional Refining", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.

Süd-Chemie G-41P RS hydrolysis catalyst is used to hydrolyze the COS to H_2S and H_2O. This is a potassium chromate on aluminum oxide catalyst and is provided in 1/8 inch extrudates. The expected catalyst life is greater than 60 months.

Süd-Chemie G-72E ZnO catalyst/sulfur adsorbent is used to capture the sulfur and reduce the residual syngas sulfur content to less than 0.1 ppm. In order to provide continuous H_2S removal, the process design uses a two bed reactor configuration with the two beds in series. Necessary piping is provided so that these two beds can be switched, and the spent adsorbent can be replaced without any interruption of service. When H_2S breakthrough occurs in the first bed (lead bed), it is taken out of service for adsorbent replacement, and the other bed (lag bed) is in service alone. After the adsorbent has been replaced, the bed with the freshly loaded adsorbent is put back in service as the lag bed. The two bed in series operation continues until H_2S breakthrough occurs in the other bed, and it is taken out of service for adsorbent replacement. The operating cycle repeats. Each catalyst bed is sized for a six month cycle length.

The hot syngas then enters the ZnO sulfur adsorption beds, 200R-2 and 200R-3. Although it is not shown in the drawing, these two beds are arranged in a lead-lag configuration so that one bed may be taken off line for ZnO replacement while the other remains in service.

The Fischer-Tropsch slurry-bed reactor converts the sulfur-free syngas primarily into olefinic hydrocarbons by the reaction

$$n\ CO + 2n\ H_2 \rightarrow C_nH_{2n} + n\ H_2O$$

The reaction is promoted by an iron-based catalyst which also promotes the water-gas shift reaction

$$CO + H_2O \rightarrow CO_2 + H_2$$

The slurry-bed F-T hydrocarbon synthesis reactor temperature is controlled by the generation of 440ºF/375 psia in tubes within the reactor. Most of this steam is sent to the combined cycle power block where it is superheated in the HRGS and used for power production.

In order to maintain a constant catalyst activity, there is a continual addition of fresh catalyst and a continual withdrawal of used catalyst from the slurry-bed reactor. The fresh catalyst must be pretreated in a reducing atmosphere at an elevated temperature to activate it. The catalyst pretreating system consists of a similar vessel to the slurry-bed reactor, but without the internal cooling facilities.

The lighter hydrocarbon products leave the slurry-bed reactor in the vapor phase, are cooled and the condensed liquid collected. The unconverted syngas (CO and H_2), carbon dioxide, methane, and the C2 and heavier material in the vapor is compressed and sent to the power block where it becomes fuel for the combustion turbine. The heavier hydrocarbons are removed as liquids from the reactor, separated from the suspended catalyst, cooled, and combined with the lighter products to form the liquid fuel precursors product.

IV.4 Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction

Starting from the Subtask 1.3 Next Plant and Subtask 2.1 designs, the Subtask 2.2 plant was developed by eliminating one gas turbine along with the export steam and hydrogen production facilities and replacing them with a large single-train, once-through Fischer-Tropsch hydrocarbon synthesis area.

Table IV.3 compares the design input and output streams for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction with those of the Subtask 2.1 non-optimized plant. The Subtask 2.2 plant consumes 5,417 tpd of dry petroleum coke, and produces 366.9 MW of export power and 10,450 bpd of liquid hydrocarbons. It also produces 373 tpd of sulfur and 195 tpd of slag. During periods when the plant produces insufficient power to satisfy its own internal demands, power is purchased to maintain the liquid hydrocarbon production. No natural gas is consumed during design operations. However, the plant does use natural gas during startup.

The Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction is shown schematically in Figure IV.4. The syngas generation and clean up sections of the plant essentially are a two train facility, but it contains a complete spare gasification train. There also are tow sulfur recovery trains and two sulfur production trains that are sized so that they only have sufficient capacity to process the output from two gasifiers simultaneously operating at design capacity. The combined cycle power block contains only one General Electric 7FAe+ combustion turbine, HRSG, and a single steam generator. The F-T hydrocarbon synthesis section also is a single train facility.

On a lower heating value (LHV) basis, the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction has a thermal efficiency 55.1% when the heating value of the byproduct sulfur is included and 53.2% when the byproduct sulfur is not included. On a higher heating value (HHV) basis, the plant has a thermal efficiency 56.7% when the heating value of the byproduct sulfur is included and 54.9% when the byproduct sulfur is not included. These efficiencies are significantly higher than those of the Subtask 2.1 non-optimized plant, which has a LHV efficiency of 47.8% and a HHV efficiency of 47.9%, both of which include the heating value of the byproduct sulfur. This is because the liquid hydrocarbon product is a larger portion of the useable energy output of the optimized Subtask 2.2 plant than it is in the non-optimized Subtask 2.1 plant. The thermal efficiencies of both of these plants are significantly higher than those of the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant which range between 38.6 and 42.1%.

Figure IV.5 is a simplified block flow diagram of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. As with Subtask 2.1, this plant consists of three distinct main processing areas.

- The gasification island and air separation unit (Areas 100, 150, 250, 300, 350, 400, 420, and 800)
- The F-T hydrocarbon synthesis area (Areas 200 and 201)
- The power block (Areas 500 and 600)

In addition there is a balance of plant area (Area 900).

Table IV.3

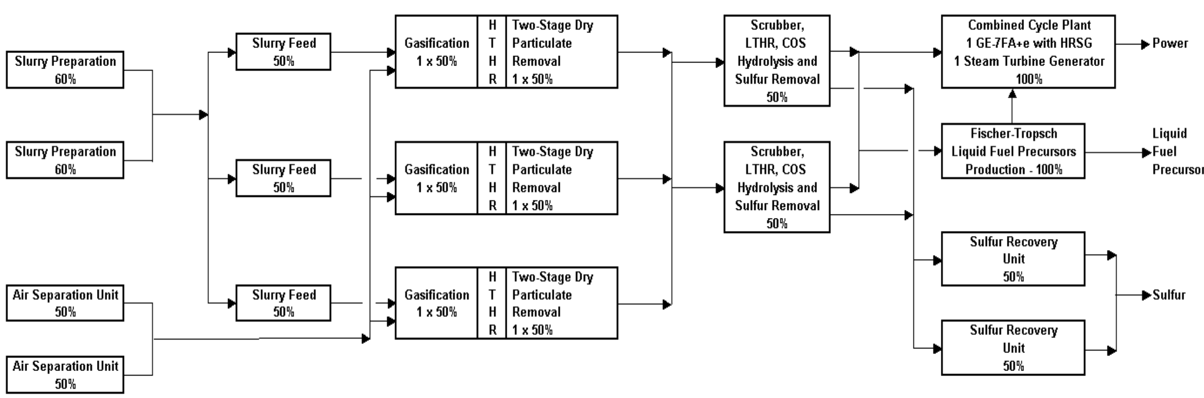
Design Input and Output Streams for the Subtask 2.1 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction and the Subtask 2.1 Optimized Coke Gasification Power Plant with Liquids Coproduction

	Subtask 2.1 [Non-optimized] Coke IGCC Coproduction Plant	Subtask 2.2 Optimized Coke IGCC Coproduction Plant
Plant Inputs		
Coke Feed, as received TPD	5,649	5,684
Coke Feed to Gasifiers, TPD	5,376	5,417
Oxygen Production, TPD of 95% O_2	5,919	5,877
Total Fresh Water Consumption, gpm	6,472	5,693
Flux, TPD	109.7	110.6
Plant Outputs		
Net Power Output, MW	617.0	366.9
Liquid Fuel Precursors, bpd	4,125	10,450
Sulfur, TPD	371	373
Slag, TPD (15% moisture)	194	195

Figure IV.4

Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction

Simplified Block Train Diagram

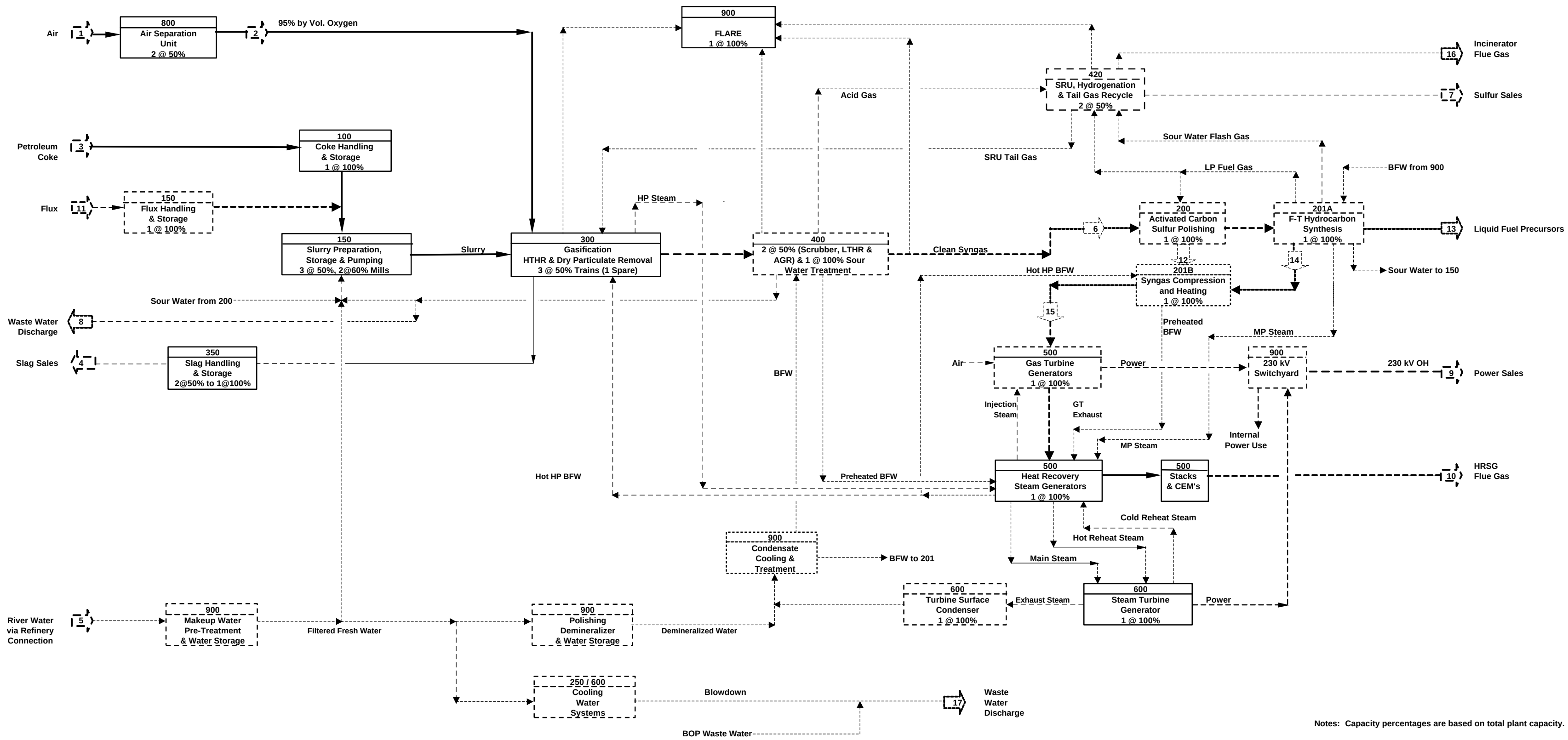


Notes: Capacity percentages are based on total plant capacity.

Figure IV.5

Block Flow Diagram of the Subtask 2.2 Optimized Coke

Gasification Power Plant with Liquids Coproduction



Notes: Capacity percentages are based on total plant capacity.

	1	2	3	4	5	6	7	8	9
Flow	Air 25,624 Tons/Day	Oxygen 5,877 Tons/Day	Coke 5,417 Tons/Day	Slag 195.1 Tons/Day	Water 2,846,500 Lb/Hr	Syngas 1,035,700 Lb/Hr	Sulfur 373.4 Tons/Day	Water 41,850 Lb/Hr	Power 366,900 kWe
Nominal Pressure - psig	Atmos.	609	NA	NA	50	365	NA	62	NA
Temperature - F	70	240	Ambient	180	70	100	332	80	NA
HHV Btu/lb	NA	NA	14,848	NA	NA	4,847	3,983	NA	NA
LHV Btu/lb	NA	NA	14,548	NA	NA	4,602	3,983	NA	NA
Energy - MM HHV/hr	NA	NA	6,703	NA	NA	5,020	124	NA	NA
Energy - MM LHV/hr	NA	NA	6,567	NA	NA	4,766	124	NA	NA
Notes	Dry Basis	5,583 O2	Dry Basis	15%Wtr.	5693 GPM		Sales	84 GPM	230 kV

	10	11	12	13	14	15	16	17	18	19	20	21
Flow	Flue Gas 3,983,400 Lb/Hr	Flux 110.6 Tons/Day	Syngas 80,210 Lb/Hr	Liq Fuel 123,280 Lb/hr	Fuel Gas 863,320 Lb/hr	GT Fuel 943,530 Lb/hr	Flue Gas 25,809 Lb/Hr	Water 585,500 Lb/Hr				
Nominal Pressure - psig	Atmos.	NA	360	50	314	445	Atmos.	Atmos.				
Temperature - F	227	NA	100	110	80	532	500	71				
HHV Btu/lb	NA	NA	4,847	19,689	1,744	2,008	NA	NA				
LHV Btu/lb	NA	NA	4,602	18,214	1,599	1,855	NA	NA				
Energy - MM HHV/hr	NA	NA	389	2,427	1,505	1,894	NA	NA				
Energy - MM LHV/hr	NA	NA	369	2,245	1,381	1,750	NA	NA				
Notes			No S	10450 bpd				1171 GPM				

DOE Gasification Plant Cost and Performance Optimization

Figure IV.5

Subtask 2.2

OPTIMIZED COKE GASIFICATION POWER PLANT WITH LIQUID FUELS COPRODUCTION

BLOCK FLOW DIAGRAM

File: Fig IV.5 March 28, 2003

Appendix B contains a detailed description of the various processing blocks in the plant. All areas except the F-T hydrocarbon synthesis area (Areas 200 and 201) essentially are the same as the corresponding areas in the Subtask 1.3 Next Plant.

The Fischer-Tropsch hydrocarbon synthesis area consists of two sub areas, Area 200 and Area 201. Area 200 is the Final Syngas Cleanup Area, which removes the final traces of sulfur from the syngas, before it is converted to hydrocarbons in Area 201, the Hydrocarbon Synthesis and Product Recovery Area.

The Final Syngas Cleanup Area, Area 200, has been redesigned for the Subtask 2.2 plant. It now uses impregnated activated carbon to reduce the sulfur concentration to less than 0.5 ppm. This is accomplished by absorbing the small amounts hydrogen sulfide, carbonyl sulfide (COS) and trace amounts of other light organic sulfur compounds (such as CS_2) on metal impregnated activated carbon. The active bed is regenerated weekly with medium-pressure steam and a small quantity of air, and the off gas is sent to the sour water stripper (SWS) overhead cooling system to condense the steam prior to going to Claus sulfur recovery. After its useful life, the deactivated carbon is sent to the gasifier for destruction and conversion to syngas and slag. The metal activator is entrained in the slag, which is a non-hazardous waste.

In order to provide continuous H_2S removal, the process design uses a three bed reactor configuration with two beds in series to remove sulfur (the second bed is a guard bed). The third bed is in regeneration. Necessary piping is provided so that these beds can be switched into any position, and when necessary, the spent adsorbent can be replaced without any interruption of service. When H_2S breakthrough occurs in the first bed (lead bed), it is taken out of service for regeneration (or adsorbent replacement, when necessary), and the other bed (lag bed) is placed in the first position. The freshly regenerated bed now becomes the second bed. This two bed in series operation continues until H_2S breakthrough occurs in the first bed, and it is removed from service for regeneration causing the operating cycle to repeat. Each carbon bed is sized for a one week cycle. Each activated carbon bed has an expected life of about three years so that, on average, one bed should be replaced each year.

The activated carbon beds are sized to process all the syngas from the gasification block. The syngas leaving the carbon beds is split in two streams with 92% going to the slurry bed F-T hydrocarbon synthesis reactor after being preheated. The remaining 8% of the syngas leaving the carbon beds is bypassed around the F-T reactor and sent to the combustion turbines after being mixed with the F-T offgas and preheated.

The F-T hydrocarbon synthesis section of the Subtask 2.2 plant is essentially the same as that of the Subtask 2.1 with a few minor changes for improved efficiency.

- Low-pressure steam from the combined cycle plant is used to preheat the syngas going to the F-T reactor.
- A second vapor/liquid separator was added to the F-T reactor vapor cooling loop to remove liquid water to prevent freezing in the downstream refrigerated cooler.
- The liquid product recovery from the vapor stream leaving the F-T reactor was improved by adding a refrigerated condenser at 40ºF following the cooling water condenser.
- High-pressure steam from the combined cycle plant is used to heat the catalyst pretreater instead of a fired furnace burning natural gas.

- A spare F-T offgas/syngas compressor was added to improve the reliability of the fuel supply to the combustion turbine.

As a result of the increased F-T liquids recovery from the reactor vapor, the combined gas turbine fuel gas has a lower heating value of about 164 Btu/scf unmoisturized. Based on previous information from General Electric, this fuel gas could be used in the GE7FA+e combustion turbine when moisturized to a lower heating value of 147 Btu/scf for NOx control but would require a higher inlet pressure. Thus, in the Subtask 2.2 design, the complete gas turbine fuel gas (F-T product gas and syngas bypassing the F-T area) is compressed to 475 psia. It is then sent to the gasification block where it is moisturized using low-level heat from syngas cooling and heated to 425ºF with intermediate pressure steam from the gasification block before going to the combustion turbine.

In the Subtask 1.3 Next Plant and Subtask 2.1 designs, the syngas going to the gas turbine has a heating value of over 275 Btu/scf. In order to minimize NOx production in the gas turbine, the turbine fuel gas must be diluted with an inert materialt to a lower heating value of about 147 Btu/scf. In these cases, the diluent is water which is vaporized in the moisturizer. The energy required to vaporize all this water essentially "goes up the stack." In the Subtask 2.2 design, the CO_2 produced in the F-T slurry-bed reactor acts as the diluent, and all the energy that went into vaporizing this water now can be used to generate steam which can be used to generate additional power. Furthermore, this reduces the raw water requirement which allows a smaller and cheaper makeup water treatment facility.

IV.5 Plant Costs

Table IV.4 shows the "overnight" EPC cost for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction and compares it with that of the Subtask 2.1 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Plant IGCC Coproduction Plant. These costs are on a mid-year 2000 basis; the same basis as those of the other Task 1 plant costs.[3]

The Subtask 2.2 EPC cost was developed from the Subtask 2.1 and Subtask 1.3 Next Plant EPC costs by subtracting the cost of the hydrogen production and compression facilities, and then adding the cost of the F-T hydrocarbon synthesis area and adjusting the cost of the power block. because it is now contains only one gas turbine and HRSG. No adjustments were made to the costs of the solids handling and ASU areas. The cost of the gasification block was adjusted to account for the cost of a smaller incinerator and the removal of one of the two syngas moisturizers. Adjustments also were made to the balance of plant area as appropriate.

The cost of the F-T area was estimated from the processing equipment sizes using an appropriate installation factor that was developed from previous cost estimates for similar facilities. The estimated cost of the large F-T slurry-bed hydrocarbon synthesis reactor is over 60% of the total equipment cost in the F-T area, and consequently, it dominates the cost of this area. Until wider experience is obtained with the construction of these large reactors, their estimated cost cannot have a high degree of accuracy.

[3] All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

Table IV.4

Capital Cost Summary of the Subtask 2.2 Optimized, Subtask 2.1 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction, and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

Plant Area	Subtask 2.2 Optimized Power and Liquids Plant	Subtask 2.1 [Non-optimized] Power and Liquids Plant	Subtask 1.3 Next Plant (Coproduction Plant)
Solids Handling	8,012,000	8,012,000	8,012,000
Air Separation Unit	107,246,000	107,246,000	107,246,000
Gasification	300,288,000	312,591,000	312,591,000
Hydrogen Production	0	0	42,931,000
F-T Liquids Area	72,368,000	34,270,000	0
Power Block	178,631,000	276,414,000	237,045,000
Balance of Plant	68,748,613	79,420,000	79,420,000
Total	735,294,000	817,953,000	787,245,000

Notes:
1 Because of rounding, the columns may not add to the total that is shown.
2. All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

The accuracy of the total installed cost for the Subtask 1.3 Next Optimized Coke IGCC Coproduction Plant was estimated to be on the order of ±11%. This level of accuracy reflects a high degree of confidence based on the large number of vendor quotes that were obtained and that the power block costs are based on a current similar Gulf Coast power project. This accuracy applies only to the total plant cost and does not apply to the individual areas or parts.

The accuracy of the total installed cost for the Subtask 2.1 Optimized Coke Gasification Power Plant with Liquids Coproduction is not as good. The estimated cost of the F-T area is only an order of magnitude cost estimate (nominally ±30%) because of the manner in which it was developed. Thus, the over estimate accuracy for the Subtask 2.1 plant probably is in the ±15% range. Because the cost of the F-T area of the Subtask 2.2 plant is a larger portion of the plant cost, the accuracy of the Subtask is less, and probably about ±20%.

IV.6 Availability Analysis

The common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period, all are dependent on the project cash flow. The net cash flow is the sum of all project revenues and expenses. Depending upon the detail of the financial analysis, the cash flow streams usually are computed on 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. Therefore, the annual production rate is a key parameter in determining the financial performance of a project.

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.[1] 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 three adjustments, this data was used to estimate the availability of the Subtask 2.1 and 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction under the several operating scenarios, with/without natural gas or backup power purchases. The first adjustment increased the availability of the air separation plant from the observed availability of 96.32% to the industry average availability of 98%. The second adjusted the availability of the first gasification stage to remove a slag tap plugging problem caused by an unexpected change in the coal blend to the gasifier. This adjustment is justified since a dedicated petroleum coke plant would be very unlikely to experience this problem. The third eliminated a short outage that occurred in the water treatment facility because this plant will have sufficient treated water storage to handle this type of outage.

Using the EPRI recommended procedure, availability estimates were calculated for the two subtasks under various operating scenarios of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction.[4]

For each plant, the potential syngas availability from two of the three gasification trains at full design capacity is 86.5% of the time, and from only one of the two trains, it is 99.63%. The equivalent syngas availability is 93.24% of the design capacity.

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.[5] However, the following availability and financial analyses are based on the data reported in the final repowering project report for consistency with the Task 1 results. This will cause the following results to be somewhat conservative, but they will be consistent with the previously reported Task 1 results.

Table IV.5 shows the design and annual average feed and product rates for the Subtask 2.2, Subtask 2.1 and Subtask 1.3 Next Plant under the “best” operating scenario; i.e., the operating scenario that maximized the return on investment under the basic economic parameters shown in Table II.1. The other operating scenarios that were considered are described in Appendices A and B. Although these other operating scenarios had lower ROIs than the cases shown under the basic economic parameters, they could have slightly higher ROIs under a different set of economic parameters.

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[4] 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.

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

Table IV.5

Design and Daily Average Feed and Product Rates for the Subtask 2.2, Subtask 2.1 and Subtask 1.3 Next Plant Cases

Subtask 2.2 Optimum Design

Case	Design	Daily Average w Power Purchase
Feed Rates		
Coke, TPD dry	5,417	4,984
Flux, TPD	110.6	101.8
Natural Gas, MMBtu/d	0	0
Product Rates		
Export Power, MW	366.9	316.4
Sulfur, TPD	373.4	343.6
Slag, TPD	195.1	179.5
F-T Liquids, bpd	10,450	9,702

Subtask 2.1 Non-Optimum Design

Case	Design	Daily Average w Gas Purchase
Feed Rates		
Coke, TPD dry	5,375	4,805
Flux, TPD	109.7	98.1
Natural Gas, MMBtu/d	553	369
Product Rates		
Export Power, MW	617.0	572.5
Sulfur, TPD	370.6	331.3
Slag, TPD	193.6	173.1
F-T Liquids, bpd	4,125	3,983

Subtask 1.3 Next Plant

Case	Design	Daily Average w Gas Purchase
Feed Rates		
Coke, TPD dry	5,417	4,842
Flux, TPD	110.6	98.9
Natural Gas, MMBtu/d	0	9,059
Product Rates		
Export Power, MW	474.0	448.4
Steam, Mlb/hr	980.0	974.6
Hydrogen, MMscfd	80.0	78.8
Sulfur, TPD	373.4	333.8
Slag, TPD	195.1	174.4
F-T Liquids, bpd	0	0

Because the Subtask 2.2 plant only has one combustion turbine, external power is purchased during turbine outages to keep the gasification island and F-T hydrocarbon synthesis area operating during turbine outages. The other two subtasks purchase natural gas during gasification island outages to produce power either for export or for internal use.

IV.7 Financial Model Results

Figure IV.6 shows the return on investment for the Subtask 1.3 Next Plant, Subtask 2.1 and Subtask 2.2 IGCC coproduction plants as a function of the power selling price using the basic economic parameters given in Table II.1 with a 10% loan interest rate. At a 27.0 \$/MW-hr export power price, the Subtask 2.2 Maximum F-T Liquids Case has an 18.24% ROI, the Subtask 2.1 Base Case has a 9.50% ROI, and the Subtask 1.3 Next Plant has a 9.05% ROI.

The ROI for the Subtask 2.1 plant has a greater slope versus power price than that of the Subtask 2.2 plant because the revenue generated from the power sales is a significantly larger portion of the total plant revenue. As such, any change in the power price will have a larger influence on the ROI. Thus, at power prices above 45 \$/MW-hr, the Subtask 2.1 non-optimized plant has a higher ROI than the Subtask 2.2 optimized plant because it produces more higher value export power and less liquid fuel precursors which are valued at 30 \$/bbl.

Table IV.6 shows the Return on Investments and Required Product Selling Prices for the Subtask 2.1 [Non-optimized] and Subtask 2.2 Optimized Coke Gasification Power Plants with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant with a natural gas price of 2.60 \$/MW-hr. At the basic economic conditions shown in Table II.1 (at a 10% loan interest rate), the Subtask 2.2 optimized plant with backup power purchase at 27 \$/MW-hr has a 18.2% ROI. This is over 8% greater than the ROI for either the Subtask 2.1 non-optimized plant or the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. The Subtask 2.2 optimized plant also requires a lower power selling price for a 12% ROI than either of the two other plants and a lower liquid fuels selling price than the Subtask 2.1 non-optimized plant.

With an 8% loan interest rate, all three cases have higher ROIs by about 3.5%. However, their relative ranking remains the same. The Subtask 2.2 optimized coproduction plant with backup power purchase still has the best ROI followed by the Subtask 2.1 and the Subtask 1.3 Next Plant. The relative ranking of the required selling prices of power and F-T liquids for the four cases also are the same.

It is difficult to predict the future value of either power, natural gas and/or the F-T liquid fuel precursors. The liquid fuels precursors price is related to the crude oil price which also can be highly variable both because of market forces and the influence of international politics. Various studies have been made which attempt to relate the value of the F-T liquids to that of crude oil by replacing crude oil in the refinery feed stream with the F-T liquids. The resulting values for the F-T liquids generally are above the crude oil values, but the specific amount can range from 2 \$/bbl up to 10 \$/bbl depending upon the refinery configuration, the specific crude oils being replaced, and the required refinery product mix.[6]

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[6] Marano, J. J., Rogers, S., Choi, G. N., and Kramer, S. J., "Product Valuation of Fischer-Tropsch Derived Fuels," ACS National Meeting, Washington, D. C., August 21-6, 1994.

Figure IV.6

Return on Investment vs. Power Price for the Subtask 2.1 [Non-optimized] and Subtask 2.2 Optimized Coke Gasification Power Plants with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

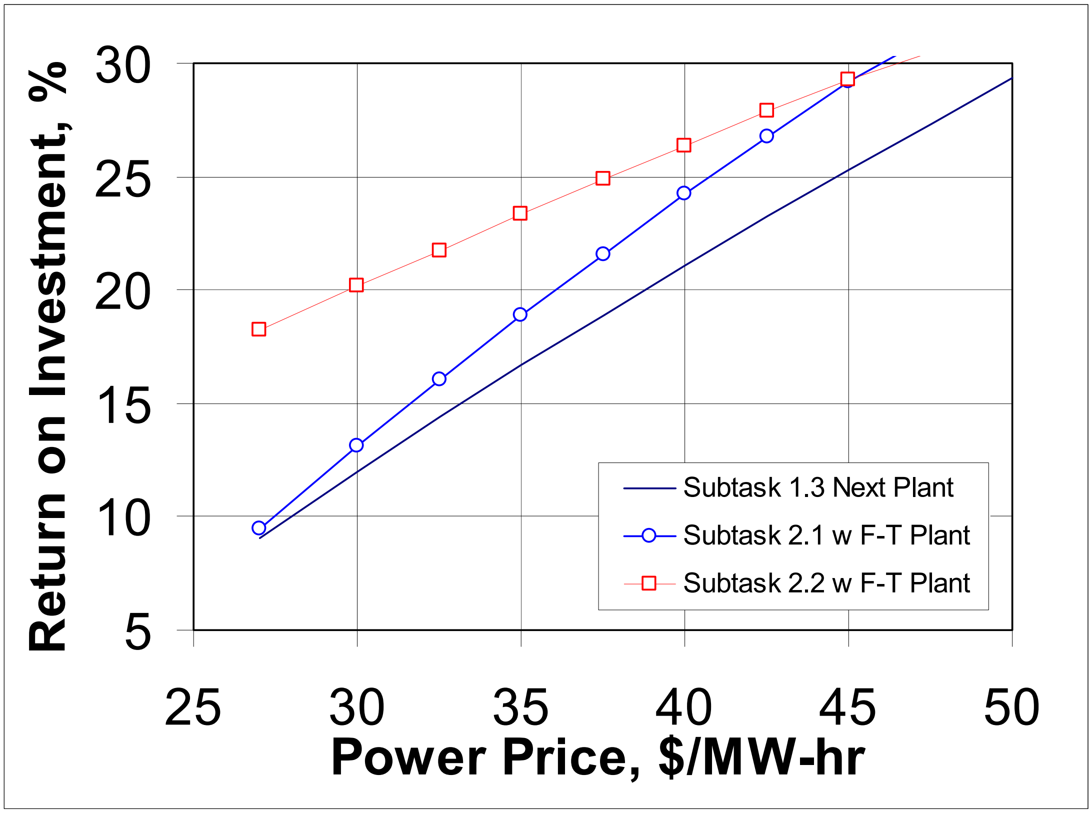


Figure IV.7 shows the effect of the liquid fuels precursors selling price on the return on investment versus the power selling price for the Subtask 2.1 Gasification Power Plant with Liquids Coproduction with a 10% loan interest rate. The solid 30 $/bbl line is the same line as shown on the previous figure for Subtask 2.1 plant.

Figure IV.8 shows the effect of the liquid fuels precursors selling price on the return on investment versus the power selling price for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. The solid 30 $/bbl line is the same line as shown on Figure IV.6 for the Subtask 2.2 plant. The dashed line is the corresponding 30 $/bbl line for the Subtask 2.1 non-optimized plant as shown in Figures IV.6 and IV.7.

Table IV.6

Return on Investments and Required Product Selling Prices for the Subtask 1.3 Next Plant, Subtask 2.1 and Subtask 2.2 Coproduction Plants (with a Natural Gas Price of 2.60 $/MMBtu)

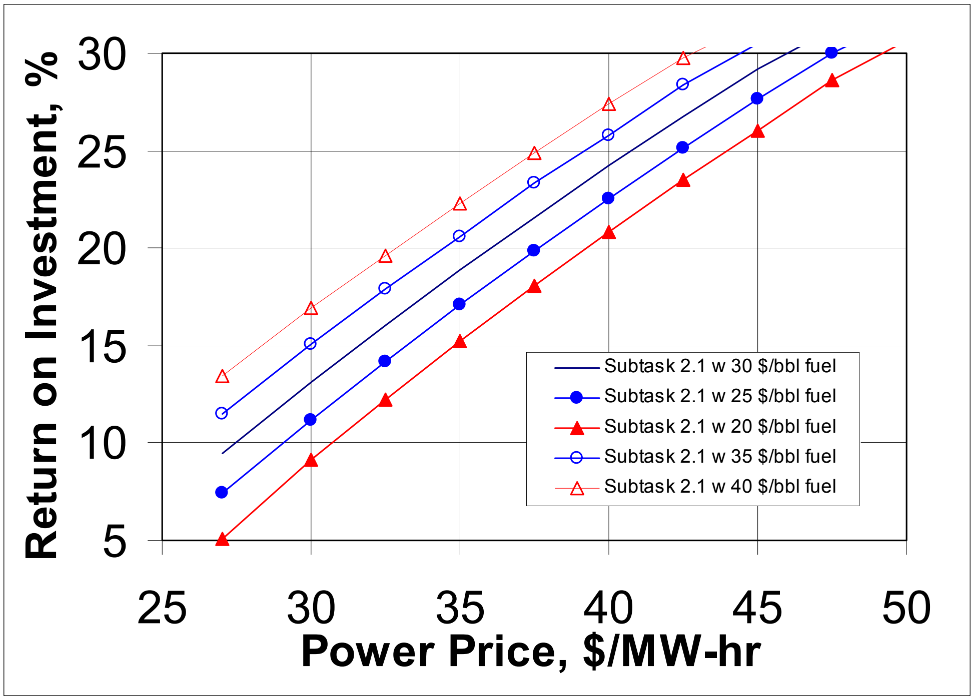
	Subtask 1.3 Next Plant w Gas Purchase	Subtask 2.1 Non-optimized w Gas Purchase	Subtask 2.2 Optimized w Power Purchase
With a 10% Loan Interest Rate			
Return on Investment with 27 $/MW-hr Power and 30 $/bbl Liquids	9.05%	9.50%	18.24%
Required Selling Price for a 12% ROI of			
Power with 30 $/bbl Liquids, $/MW-hr	30.02	29.04	17.71
Liquds with 27 $/MW-hr Power, $/bbl	NA	36.22	23.65
With a 8% Loan Interest Rate			
Return on Investment with 27 $/MW-hr Power and 30 $/bbl Liquids	12.70%	13.24%	21.81%
Required Selling Price for a 12% ROI of			
Power with 30 $/bbl Liquids, $/MW-hr	26.32	26.04	12.81
Liquds with 27 $/MW-hr Power, $/bbl	NA	27.05	20.30

Examination of this figure shows that at low power selling prices, it is better to maximize liquid fuels production as is done in Subtask 2.2 than maximize power production as is done in Subtask 2.1. Below a power price of about 35 $/MW-hr, the Subtask 2.2 plant has a higher ROI when the liquid fuels are 25 $/bbl or higher. As the power price increases, the Subtask 2.2 plant requires higher liquid selling prices to maintain higher ROIs than the Subtask 2.1 plant.

After commissioning all plants undergo a shakedown periods during which problem areas are corrected, inadequate equipment is repaired or replaced, and adjustments are made. Also as multiple plants start up and operate, the technology goes through a “learning curve” and improvements are incorporated into the next generation of plants. Consequently, performance is likely to improve as measured by increased capacity and/or improved on-stream factors. Figure IV.9 shows the effect of improved syngas availability on the return on investment for the Subtask 1.3 Next Plant, 2.1 and 2.2 IGCC coproduction plants. The abscissa is equivalent syngas availability; i.e., the total syngas availability from the three gasification trains expressed as a percentage of the time that two gasification trains will be producing syngas at the design rate. For the Subtask 2.2 optimized plant, as the syngas availability improves, the amount of backup power that has to be purchased is reduced until it completely disappears at the unattainable 100% syngas availability. For the Subtask 2.1 non-optimized plant, the amount of purchased natural gas decreases in a similar manner as the syngas availability improves. However, even at 100% syngas availability, a small amount of natural gas is required for furnace fuel in the F-T area. At the expected 86.85% syngas availability, the Subtask 2.2 plant has an ROI of 18.24%, the Subtask 2.1 plant has a

Figure IV.7

Return on Investment vs. Power Price Showing the Effect of the Liquids Price for the Subtask 2.1 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction



ROI of 9.50%, and the Subtask 1.3 Next Plant has a ROI of 9.05%. At 90% syngas availability, the ROI of the Subtask 2.2 plant increases to about 19.4%, that of the Subtask 2.1 plant increases to about 10.8%, and that of the Subtask 1.3 Next Plant increases to 10.1%. At the unattainable syngas availability of 100%, the Subtask 2.2 optimized plant will have an expected ROI of 23.1%, the Subtask 2.1 non-optimized plant will have an expected ROI of 14.9%, and the Subtask 1.3 Next Plant will have an expected 13.4% ROI.

The sensitivities of individual component prices and some financial parameters on the return on investment for Subtask 2.1 are given in Table V.2 of Appendix A and for Subtask 2.2 in Table V.3 of Appendix.

Figure IV.8

Return on Investment vs. Power Price Showing the Effect of the Liquids Price for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction

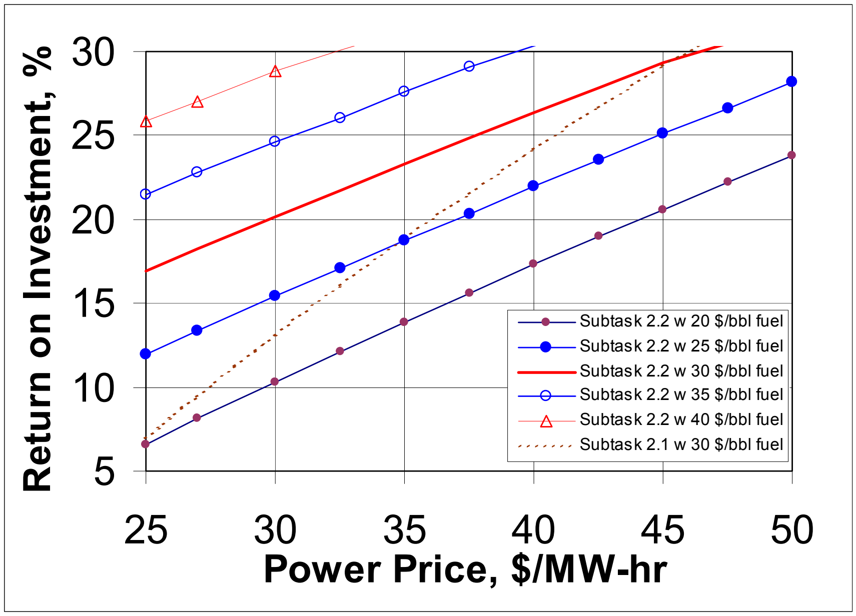
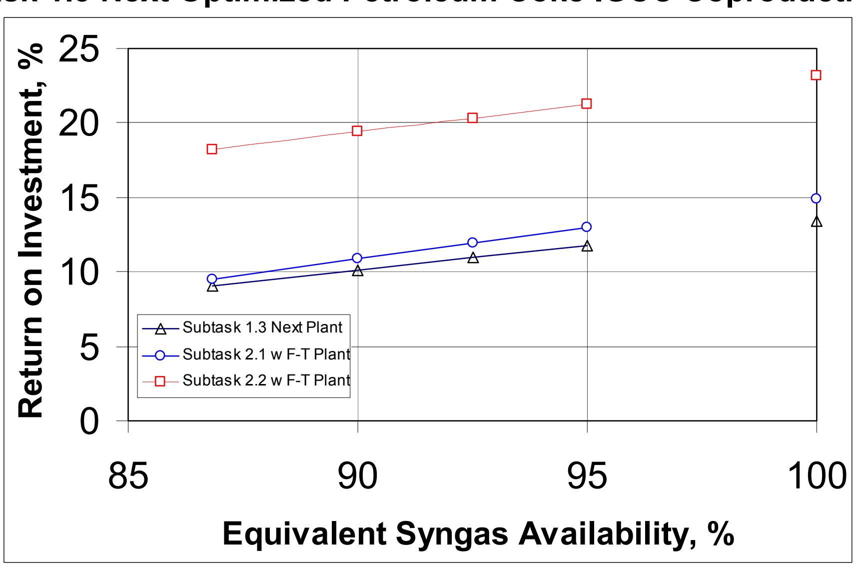


Figure IV.9

Return on Investment vs. Syngas Availability for the Subtask 2.1 [Non-optimized] and Subtask 2.2 Optimized Coke Gasification Power Plants with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant



Chapter V

Coal Cases

V.1 Introduction

The design for one coal IGCC gasification power plant with liquid fuels coproduction was developed in Task 2. The Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction was developed by combining the gasification area of the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant with the Fischer-Tropsch hydrocarbon synthesis section of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. The net result of this combination is a larger plant containing four parallel gasification trains.

V.2 Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant

The Subtask 1.6 plant design was optimized by applying nine Value Improving Practices (VIPs) to the Subtask 1.1 plant and by expanding the plant to four trains to reduce costs and improve operability.[1] As a result of this effort, plant performance was improved, the plant cost was reduced, and the return on investment was significantly improved. The results of this VIP and optimization study included:

- Simplified solids handling system
- Removal of the feed heaters and spare pumps
- Maximum use of slurry quench
- Maximum syngas moisturization
- Use of a cyclone and a dry particulate removal system to clean the syngas
- Smaller T-120 post reactor residence vessel
- Simplified Claus plant, amine and sour water stripper
- Use of state-of-the-art GE 7FA+e gas turbines with 210 MW output and lower NOx
- Use of steam diluent in the gas turbines
- Development of a compact plant layout to minimize the use of large bore piping
- Used Bechtel's advanced construction techniques to reduce costs
- Added design features to reduce O&M costs and increase syngas availability

Table II.1 shows the design input and output streams for the Subtask 1.6 IGCC Plant. The plant consumes 9,266 tpd of dry coal and produces 1,154.6 MW of export power. It also produces 237 tpd of sulfur and 1,423 tpd of slag. No natural gas is consumed during design operations. However, the plant does use natural gas during startup and as a supplementary fuel to fire the combustion turbines when insufficient syngas is available.

The resulting design configuration for the Subtask 1.6 IGCC Plant is shown in Figure V.1. The plant basically is a four-train gasification facility, without a spare gasification train or spare gasification capacity. There are also four gas turbines in two combined cycle trains.

Table V.1

[1] "Task 1 Topical Report – IGCC Plant Cost Optimization", Gasification Plant Cost and Performance Optimization, Chapter II, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, May 30, 2002

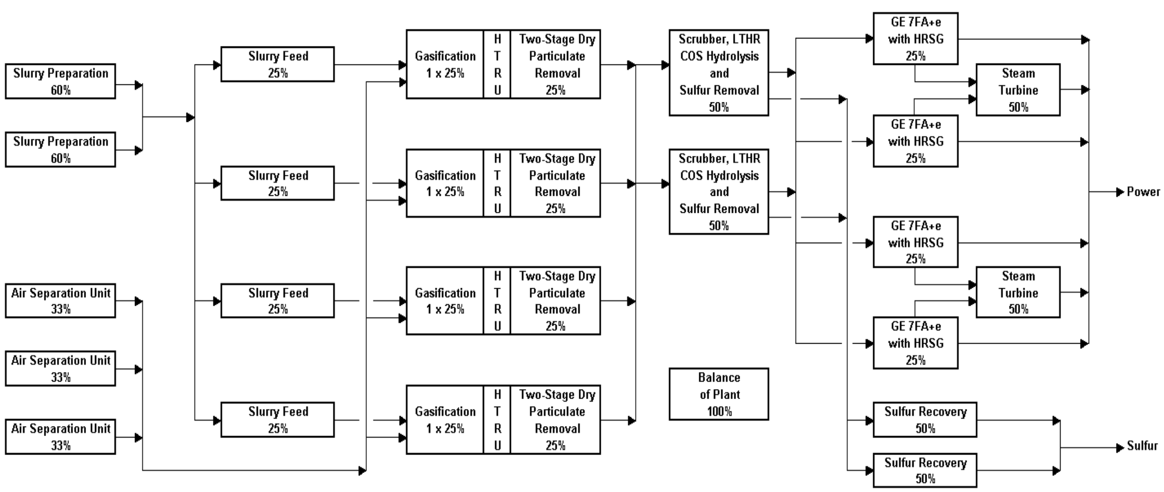
Design Input and Output Streams for the
Subtask 1.6 Nominal 1,000 MW Optimized Coal IGCC Plant

	Subtask 1.6 1,000 MW Plant
Plant Inputs	
Coal Feed, as received TPD	10,837
Coal Feed to Gasifiers, TPD	9,266
Oxygen Production, TPD of 95% O_2	8,009
Total Fresh Water Consumption, gpm	9,752
Plant Outputs	
Net Power Output, MW	1,154.6
Sulfur, TPD	237
Slag, TPD (15% moisture)	1,423

Figure V.1

Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant

Simplified Block Train Diagram



Note: Capacity percentages are based on total plant capacity.

The Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant has an overall LHV thermal efficiency of 42.4% and a HHV thermal efficiency of 40.8%, both of which include the heat content of the byproduct sulfur. When the byproduct sulfur is not considered, the plant has a LHV thermal efficiency of 41.6% and a HHV efficiency of 40.0%.

Without backup natural gas, at the expected 75.7% single train syngas availability, the required power selling price for a 12% ROI with 30 \$/bbl F-T liquids selling price is 48.1 \$/MW-hr. At an 80% syngas availability, the required power selling price drops by almost 4 \$MW-hr to 44.3 \$/MW-hr. At the unattainable 100% syngas availability, it is the same as the case with backup natural gas, 31.6 $/MW-hr, since at this point, no backup natural gas is required.

V.3 Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction

The Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction was developed from the Subtask 1.6 Nominal 1,000 MW IGCC plant using the design approach adopted for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. The coal gasification capacity of the plant was kept the same as Subtask 1.6; i.e., that amount which could be processed in four gasification trains, to allow direct comparison between these two cases. Power production was reduced to only one power block train consisting of two combustion turbines, two HRSGs, and a single steam turbine. In order to satisfy the fuel demand of the gas turbines, only about 82% of the available syngas could be processed in the F-T hydrocarbon synthesis reactors. The unconverted syngas and light hydrocarbons from the F-T synthesis section is compressed and combined with the remaining 18% of syngas bypassing the F-T reactors to provide fuel for the two combustion turbines.

Table V.2 compares the design input and output stream flows for the Subtask 2.3 and Subtask 1.6 Nominal 1,000 MW IGCC plant. From 9,266 tpd of dry coal, the plant produces 12,377 bpd of liquid fuel precursors, 675.9 MW of export power, and 237 tpd of sulfur. The LHV heating value of the liquid fuel product is 2,661 MMBtu/hr, or 780 MW_t. The export power production is reduced to 675.9 MW from 1154.7 MW for the Subtask 1.6 IGCC plant. Overall, the combined energy in the liquid fuel precursors and the electric power products (1,455 MW) is increased compared to Subtask 1.6 (1,154.7 MW) with the export power being only 46.5% of the plant output.

Figure 3.1 shows the train configuration of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. As shown, Subtask 2.3 includes three air separation units, four parallel gasification trains, two parallel 50% F-T hydrocarbon synthesis trains, two GE 7FA+e combustion turbines with HRSGs, and one steam turbine.

Appendix C contains a detailed description of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction

The F-T processing area of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction includes the following design improvements that were applied to Subtask 2.1:

- Use of regenerable activated carbon for final/trace sulfur removal
- Addition of refrigeration to increase the light oil recovery from the F-T area
- Replacement of the fired heater in the F-T catalyst preparation area with steam heating using high-pressure steam from the gasification block to eliminate the use of natural gas during normal operation
- Effective utilization of steam in the F-T area

Except for size, this design is the same as that used in Subtask 2.2.

On a lower heating value (LHV) basis, the plant has a thermal efficiency 53.2% when the heating value of the byproduct sulfur is included and 52.4% when the byproduct sulfur is not included. On a higher heating value (HHV) basis, the plant has a thermal efficiency 53.4% when the heating value of the byproduct sulfur is included and 52.6% when the byproduct sulfur is not included. These thermal efficiencies are higher than those that would be obtained from a coal IGCC power plant of a similar design because it includes the heating value of the liquid fuel that is produced. Since the second law of thermodynamics states this liquid fuel cannot be used at a 100% thermal efficiency, the thermal efficiency of the plant will be somewhat lower when the final disposition of the liquid fuel is considered.

Table V.2

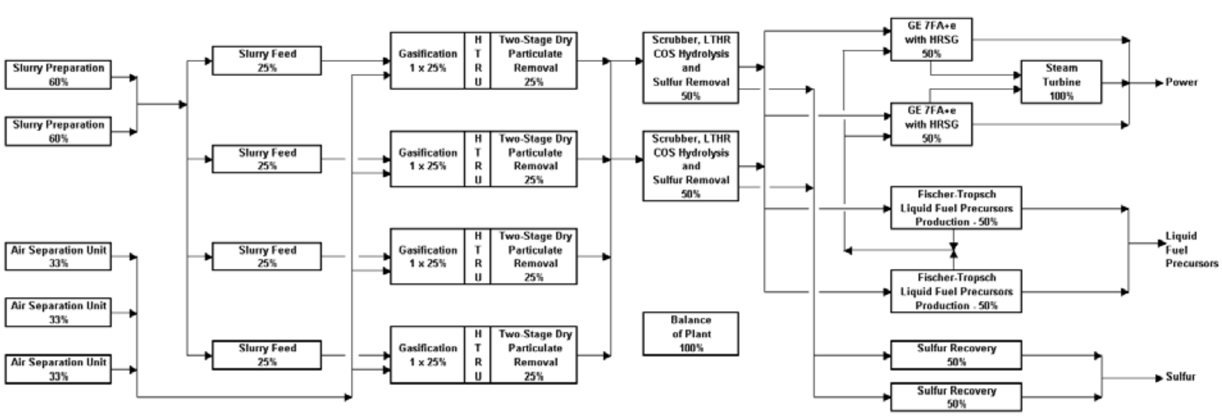
Design Input and Output Streams for the Subtask 2.3 Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 MW Coal IGCC plant

	Subtask 1.6 Nominal 1,000 MW IGCC Plant	Subtask 2.3 Optimized Power and Liquids Plant
Plant Inputs		
Coal Feed, as received TPD	10,837	10,837
Dry Coal Feed to Gasifiers, TPD	9,266	9,266
Oxygen Production, TPD of 95% O_2	8,009	7,919
Total Fresh Water Consumption, gpm	9,752	7,403
Plant Outputs		
Net Power Output, MW	1,154.6	675.9
Sulfur, TPD	273	237
Slag, TPD (15% moisture)	1,423	1,423
Liquid Fuel Precursors, bpd	0	12,377

Figure V.2

Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction

Simplified Block Train Diagram

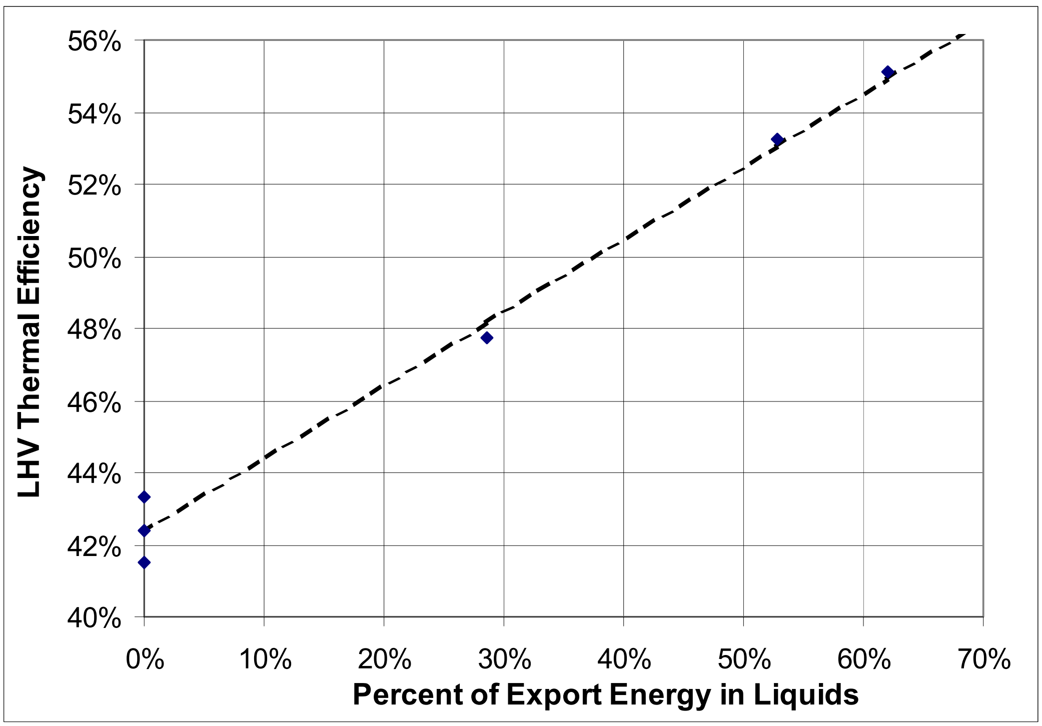


Note: Capacity percentages are based on total plant capacity.

The Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction has a LHV thermal efficiency of 53.2% including the byproduct sulfur. This is lower than that of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction (55.1% LHV thermal efficiency) because the Subtask 2.3 plant produces less F-T liquids (52.7% of the total energy in the products compared to 62.0% liquids for Subtask 2.2). Figure V.3 shows the relationship between the LHV thermal efficiency for the three Task 2 plants versus the total LHV energy in the export products (liquid fuels, power and sulfur). The three points with no liquid fuels production represent Subtasks 1.5A (coal), 1.5B (coke), and Subtask 1.6 in order of increasing thermal efficiency.

Figure V.3

Effect of Liquids Production on the LHV Thermal Efficiency



Liquids production for Subtask 2.3 was limited for two reasons:

- Gasification of coal in the E-GAS™ system produces significant by-product methane compared to coke gasification. This limits the quantity of CO and H_2 in the syngas available for conversion to F-T liquids and leads to increased F-T offgas which must be used in a gas turbine.
- Increased offgas requires two gas turbines, which in turn requires that more syngas be bypassed around the F-T plant to fully load the gas turbine.

However, in Subtask 2.3, the lower efficiency from the reduced liquids production is somewhat offset by the use of a more efficient reheat steam cycle.

Several options were considered to optimize the plant configuration and to improve the project economics. These options focused on maximizing the production of F-T liquids at the expense of reduced power output to take advantage of the high value of the F-T liquids and the higher efficiency of F-T liquids production compared to the conversion to electricity. If all the available syngas were sent to F-T hydrocarbon synthesis (in a once-through configuration), the plant would produce 15,016 bpd (compared to 12,377 bpd for the current design). In this configuration the F-T offgas available for gas turbine fuel would be 2,729 MMBtu/hr LHV at 185 Btu/scf (compared to about 1,750 MMBtu/hr LHV at a minimum energy content of 200 Btu/scf as specified by General Electric for the 7FA+e combustion turbine). The current design, which bypasses some syngas around the F-T reactors, sends 3,532 MMBtu/hr at 210 Btu/scf to fire two GE 7FA+e turbines (1,766 MMBtu/hr per turbine). A better approach would be to use a General Electric 9F or 7G/H class gas turbine, but the 9F model turbine is a 50-Hertz machine, and General Electric is not currently offering the advanced machines for syngas service.

Reducing the amount of fuel gas was also considered. One approach is to increase F-T reactor conversion to offset the high methane production during gasification with the coal feedstock. However this approach does not appear to be realistic because of the requirements for adequate reactor sizing (residence time and height), mixing, and heat transfer. Hydrogen recovery from the F-T offgas also was considered, but was ruled out because of the low concentration of hydrogen. The last option considered was to provide fuel to one GE 7FAe+ gas turbine and use the excess offgas for supplemental duct firing of the HRSG. However, this option was dropped from consideration because of the large quantity of energy (970 MMBtu/hr), which would cause a high HRSG inlet temperature. Also, the offgas would only be converted to electricity at the moderate steam cycle efficiency. In any case, the overriding considerations are to minimize capital cost, and to maximize availability and use of all process equipment to produce high value products and maximize revenue.

Figure V.4 is a simplified block flow diagram of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. This figure also contains the flow rates of the major plant input and output streams.

V.4 Plant Costs

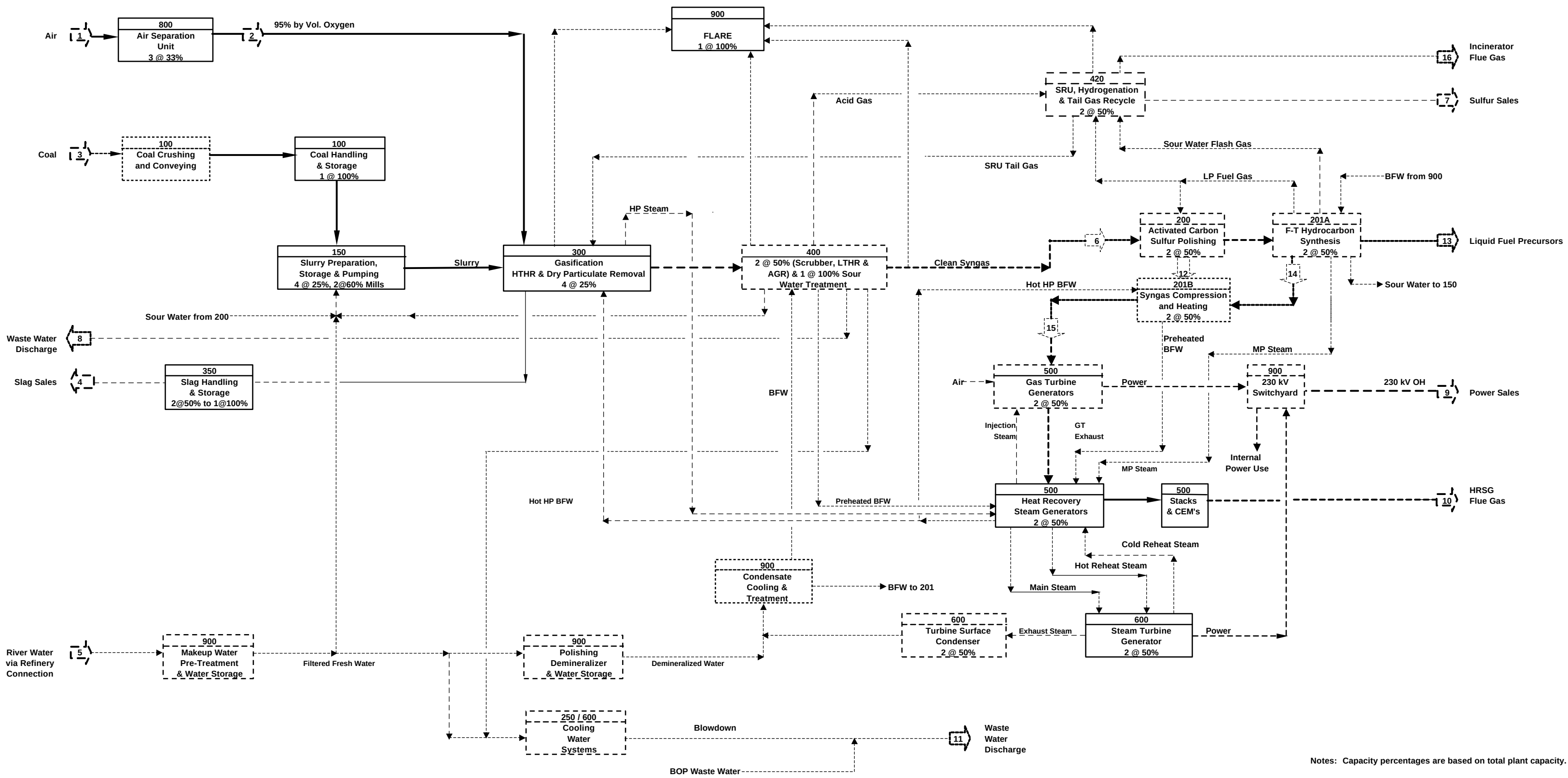
Table V.3 shows the “overnight” EPC cost for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and compares it with the Subtask 1.6 Nominal 1,000 MW IGCC. These costs are on a mid-year 2000 basis; the same basis as those of the other Task 1 plant costs.[2]

The Subtask 2.3 EPC cost was developed from the Subtask 1.6 IGCC plant and Subtask 2.2 liquids coproduction plant EPC costs by adding the cost of the F-T hydrocarbon synthesis area to the Subtask 1.6 cost estimate and by adjusting the cost of the power block

[2] All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

Figure V.4

Block Flow Diagram of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction



Notes: Capacity percentages are based on total plant capacity.

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16					
	Air	Oxygen	Coal	Slag	Water	Syngas	Sulfur	Water	Power	Flue Gas	Water	Syngas	Liq Fuel	Fuel Gas	GT Fuel	Flue Gas					
Flow	34,528	7,919	9,266	1,423	3,702,000	1,477,400	236.5	55,563	675,900	7,970,000	695,500	263,425	146,018	1,041,200	1,304,625	24,758					
	Tons/Day	Tons/Day	Tons/Day	Tons/Day	Lb/Hr	Lb/Hr	Tons/Day	Lb/Hr	kWe	Lb/Hr	Lb/Hr	Lb/Hr	Lb/hr	Lb/hr	Lb/hr	Lb/Hr					
Nominal Pressure - psig	Atmos.	610	NA	NA	50	365	NA	62	NA	Atmos.	Atmos.	360	50	314	350	Atmos.					
Temperature - F	59	240	Ambient	180	70	100	332	80	NA	237	71	100	100	80	400	500					
HHV Btu/lb	NA	NA	12,749	NA	NA	5,255	3,983	NA	NA	NA	NA	5,255	19,701	2,376	2,957	NA					
LHV Btu/lb	NA	NA	12,275	NA	NA	4,894	3,983	NA	NA	NA	NA	4,894	18,225	2,154	2,707	NA					
Energy - MM HHV/hr	NA	NA	9,844	NA	NA	7,764	78	NA	NA	NA	NA	1,384	2,877	2,474	3,858	NA					
Energy - MM LHV/hr	NA	NA	9,478	NA	NA	7,231	78	NA	NA	NA	NA	1,289	2,661	2,243	3,532	NA					
Notes	Dry Basis	7,468 O2	Dry Basis	15%Wtr.	7404 GPM		Sales	Check 111 GPM	230 kV		1390 GPM	No S	12377 bpd								

DOE Gasification Plant Cost and Performance Optimization

Figure V.4

Subtask 2.3

OPTIMIZED COAL GASIFICATION POWER

PLANT WITH LIQUID FUELS COPRODUCTION

BLOCK FLOW DIAGRAM

File: Fig V.4.xls

May 7, 2003

to reflect only one 2x1x1 combined cycle train. No adjustments were made to the costs of the solids handling area. The ASU cost was reduced to reflect a slight reduction in oxygen usage. The cost of the gasification block was adjusted to account for the removal of two syngas moisturizers. Adjustments were made to the balance of plant area, as appropriate.

The cost of the F-T area was estimated from the processing equipment sizes using an appropriate installation factor that was developed from previous cost estimates for similar facilities. The estimated cost of the large F-T slurry-bed hydrocarbon synthesis reactor is over 60% of the total equipment cost in the F-T area, and consequently, it dominates the cost of this area. Until wider experience is obtained with the construction of these large reactors, their estimated cost cannot have a high degree of accuracy.

The accuracy of the total installed cost for the Subtask 1.6 Nominal 1,000 MW IGCC Plant was estimated to be on the order of ±15%. This level of accuracy reflects a high degree of confidence based on the large number of vendor quotes that were obtained and that the power block costs are based on a current similar Gulf Coast power project. This accuracy applies only to the total plant cost and does not apply to the individual areas or parts.

The accuracy of the total installed cost for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction is not as good. The estimated cost of the F-T area is only an order of magnitude cost estimate (nominally ±30%) because of the manner in which it was developed. Thus, the over estimate accuracy for the Subtask 2.2 plant probably is in the ±20% range. Because the cost of the F-T area of the Subtask 2.3 plant also is a large portion of the plant cost, the accuracy of the Subtask 2.3 is approximately the same, ±20%.

Table V.3

Capital Cost Summary of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant

Plant Area	Subtask 2.3 Optimized Power and Liquids Plant	Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant
Solids Handling	28,317,000	28,317,000
Air Separation Unit	149,791,000	151,496,000
Gasification	434,094,000	443,301,000
F-T Liquids Area	94,283,000	0
Power Block	348,788,000	493,795,000
Balance of Plant	103,785,000	114,419,000
Total	1,159,058,000	1,231,328,000

Notes:
1 Because of rounding, the columns may not add to the total that is shown.
2. All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

V.5 Availability Analysis

As described previously in Section IV.6, a similar availability analysis also was made for the Subtask 1.6 coal IGCC power plant and the Subtask 2.3 coal power plant wilt liquid fuels coproduction.

The syngas production and cleanup area of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction are configured identical to that of the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant in that they both contain the same number of units used to generate the syngas. These areas contain two 60% slurry preparation areas, three 33.3% Air Separation Units, and four 25% gasification blocks, each with associated slurry feed, syngas cooling and cleanup sections. The final syngas cleanup and conditioning section consisting of a wet scrubber, low temperature heat recovery (LTHR), COS hydrolysis, sulfur removal, and sulfur recovery is in two 50% trains; the same as the Subtask 1.6 design. In the Subtask 1.6 design, the syngas is sent to four 25% GE7FA+e combustion turbines and HRSGs with one 50% steam turbine associated with two combustion turbines. In the Subtask 2.3 design, most of the syngas is sent to two Fischer-Tropsch hydrocarbon synthesis trains with the remainder along with the unconverted syngas from the F-T area going to two GE7FA+e combustion turbines and HRSGs. One steam turbine converts the steam from the other areas of the plant to power.

An availability analysis of both these facilities showed that the syngas availability from one gasification train including scheduled downtime is about 76%. Therefore, two modes of operation were considered; one without the use of backup natural gas and one that uses backup natural gas when sufficient syngas is not available to fire the combustion turbines. In the Subtask 2.3 case, backup natural gas only is used when insufficient syngas and F-T off gas are available to fully load a turbine. No turbine is operated only on backup natural gas.

Table V.4 shows the design and annual average feed and product rates for both operating scenarios, with and without backup natural gas for the Subtask 1.6 and Subtask 2.3 plants. At design conditions, both plants process the same amount of dry coal, 9,266 tpd. However, because of some efficiency improvements developed since the Subtask 1.6 design was completed, the Subtask 2.3 plant uses slightly less oxygen. The Subtask 2.3 average rates were developed based on the premise that it was best to maximize power production from all available gas turbines by using backup natural gas even if it were necessary to fully fire the turbine on natural gas.

On a daily average basis without backup natural gas, the Subtask 2.3 plant processes slightly less coal that the Subtask 1.6 plants because there are some situations where it was necessary to slightly reduce the coal rate in order not to overload a combustion turbine. It processes 6,899 tpd of dry coal (74.4% of design) to produce 474.4 MW of power (70.2% of design) and 9,889 bpd of liquid hydrocarbons (80.7% of design).

On a daily average basis with backup natural gas, the Subtask 2.3 plant processes 6,929 tpd of dry coal (74.8% of design) and consumes 26.5 MMscfd of natural gas to produce 613.7 MW of power (90.8% of design) and 10,397 bpd of liquid hydrocarbons (84.0% of design).

Table V.4

Design and Daily Average Feed and Product Rates for the Subtask 1.6 Coal IGCC Power Plant and the Subtask 2.3 Coal IGCC Coproduction Plant

	Subtask 1.6 1,000 MW Coal IGCC Power Plant			Subtask 2.3 Coal IGCC Coproduction Plant		
		Daily Average			Daily Average	
	Design	Without Backup Gas	With Backup Gas	Design	Without Backup Gas	With Backup Gas
Feeds						
Coal, TPD dry	9,266	7,018	7,018	9,266	6,899	6,929
Natural Gas, Mscfd	0	0	34,961	0	0	26,466
River Water, gpm	9,752	7,386	NC	7,404	5,513	NC
Products						
Export Power, MW	1,154.6	874.5	1,081.0	675.9	474.4	613.7
Liquid Hydrocarbons, bpd	---	---	---	12,377	9,989	10,397
Sulfur, TPD	236.6	179.2	179.2	236.6	176.1	176.9
Slag, TPD	1,423	1,078	1,078	1,423	1,059	1,064
Performance						
Oxygen Consumption,						
TPD of 95% O_2	8,009	6,066	6,066	7,919	5,896	5,922
TPD O_2/TPD dry coal	0.86	0.86	0.86	0.85	0.85	0.85
Water Discharge, gpm						
Process Water	59	45	45	111	83	83
Clear Water	1,248	945	NC	1,390	1,035	NC
Total Discharge	1,307	990	NC	1,501	1,118	NC
Heat Rate, HHV Btu/kW	8,526	8,526	8,245	NC	NC	NC
Thermal Efficiency, %HHV*	40.0%	40.0%	41.4%	52.6%	52.6%	56.7%

V.6 Financial Analysis Results

Figure V.5 shows the return on investment (ROI) as a function of the export power price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction under both operating scenarios and compares them with the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant with 2.60 $/MMBtu natural gas at a 10% loan interest rate. This figure shows that generally the operating scenarios that use backup natural gas have higher Return on Investments (ROIs) than the cases without backup natural gas. Below a power selling price of about 38 $/MW-hr, the Subtask 2.3 plant with the F-T liquids at 30 $/bbl has a higher ROI than the Subtask 1.6 power plant. Above this power selling price, the Subtask 1.6 plant has a higher ROI. The same situation is true for the two operating scenarios without backup natural gas except that with these cases, the breakeven power selling price is slightly higher, about 40 $/MW-hr.

Figure V.5

Return on Investment vs. Power Price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant (10% Loan Interest Rate)

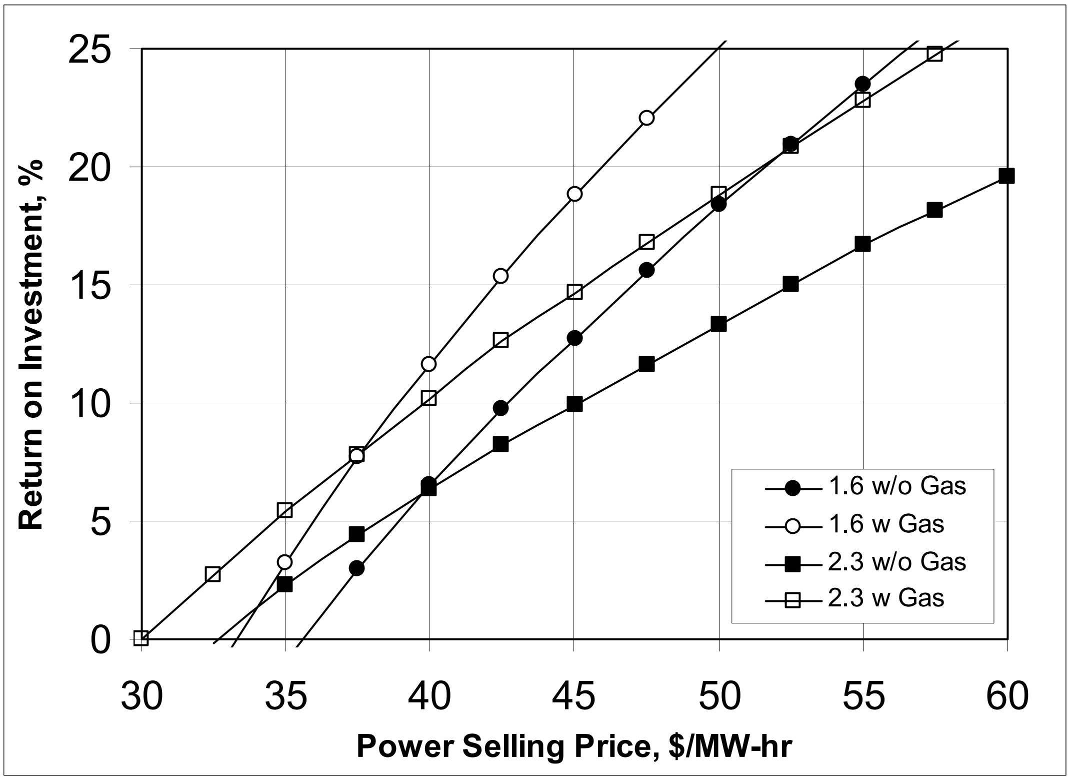


Table V.5 compares the power and F-T liquids selling prices required by the Subtask 1.6 and Subtask 2.3 plants to generate a 12% ROI for the two operating scenarios. At the basic economic conditions shown in Table II.1 (at a 10% loan interest rate), the Subtask 2.3 Coproduction Plant with backup power purchase requires a 42.02 $/MW-hr power selling price for a 12% ROI, and without backup power purchase, the required power selling price is 48.06 $/MW-hr. These required power selling prices are higher than those for the corresponding Subtask 1.6 cases. With a fixed 27 $/MW-hr power selling price, the required selling prices of the F-T liquids to produce a 12 ROI are 48.59 and 50.97 $/bbl for the cases with and without backup natural gas cases, respectively.

With an 8% loan interest rate the relative ranking of the cases remains almost the same except that the required selling prices are lower. However, the Subtask 2.3 case with backup natural gas now has a slightly lower power selling price than the Subtask 1.6 case. This is a result of the Subtask 2.3 case having a lower EPC cost than the Subtask 1.6 Case.

Table V.5

Required Power Selling Prices for the for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Power Plant With and Without Backup Natural Gas

	Subtask 1.6		Subtask 2.3	
	Without Backup Gas	With Backup Gas	Without Backup Gas	With Backup Gas
With a 10% Loan Interest Rate				
Required Selling Price for a 12% ROI of Power with 30 $/bbl Liquids, $/MW-hr	44.37	40.23	48.06	42.02
Liquds with 27 $/MW-hr Power, $/bbl	---	---	50.97	48.59
With a 8% Loan Interest Rate				
Required Selling Price for a 12% ROI of Power with 30 $/bbl Liquids, $/MW-hr	41.34	37.77	42.93	38.06
Liquds with 27 $/MW-hr Power, $/bbl	---	---	45.87	43.69

Figure V.6 shows the effect of the liquid fuels precursors selling price on the return on investment versus the power selling price for the Subtask 2.2 Maximum F-T Liquids Case with a 10% loan interest rate and 2.60 $/MMBtu natural gas. The solid 30 $/bbl line is the same line as shown on the previous figure for the Subtask 2.3 coproduction plant with backup natural gas. The dashed line represents the Subtask 1.6 power plant with backup natural gas. The ROI for the Subtask 1.6 plant has a greater slope versus the power price than that of the Subtask 2.3 plant because the revenue generated from the power sales is a significantly larger portion of the total plant revenue. As such, any change in the power price will have a larger influence on the ROI.

This figure shows that the Subtask 2.3 coproduction plant requires F-T liquids selling prices above 30 $/bbl to generate ROIs greater than 10% with power prices below 40 $/MW-hr. With a 38 $/MW-hr power selling price, the Subtask 2.3 coproduction plant will have higher ROIs that the Subtask 1.6 power plant only when the F-T liquids are selling for 30 $/bbl or greater. As the power selling price increases, the Subtask 2.3 coproduction plant requires higher F-T liquids prices to be competitive with the Subtask 1.6 plant. At a 50 $/MW-hr power price, the F-T liquids should be about 40 $/bbl or greater for the Subtask 2.3 plant to have a higher ROI.

Figure V.6

Return on Investment vs. Power Price Showing the Effect of the Liquids Price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Power Plant

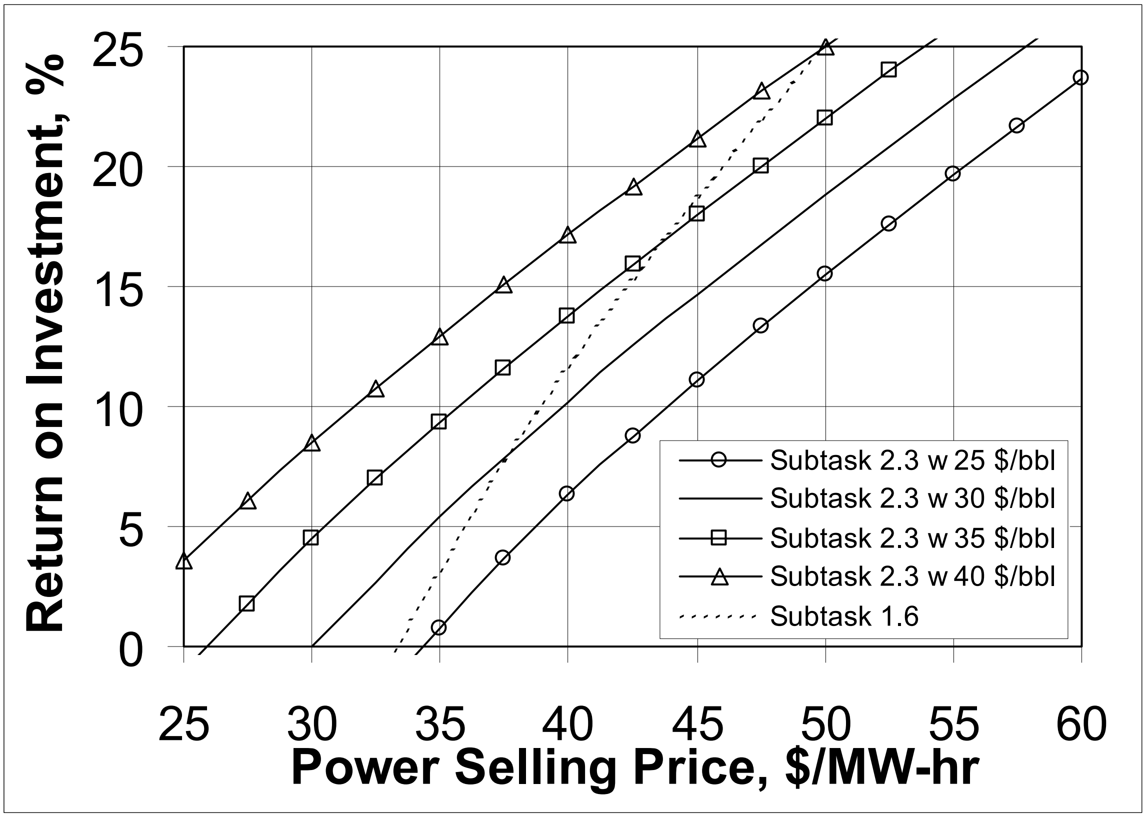
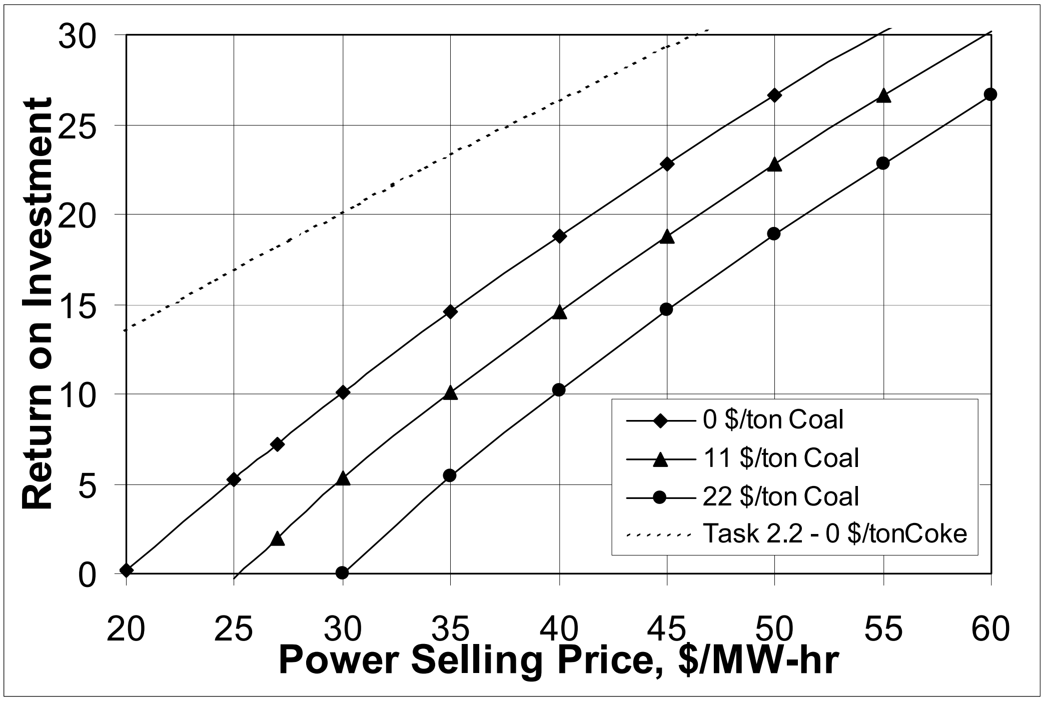


Figure V.7 shows the effect of the coal price on the return on investment on the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. The base coal price is 22.0 $/ton. Also shown on the figure are two other coal prices, 11.0 $/ton (50% of the base price) and 0 $/ton. As expected, as the coal price decreases, the ROI increases. For comparison, the return on investment of the Subtask 2.2 optimized coke is shown as the dotted line on the figure. This return is based on zero net coke price. The higher returns of the Subtask 2.2 plant shows that the cost of the coal alone does not account for the entire difference in returns between the two plants. Part of this difference is attributable to the higher fraction of high value liquid yields and higher thermodynamic efficiency of the Subtask 2.2 plant as discussed at the beginning of this chapter. The other part appears to be the higher availability of the gasification area of the Subtask 2.2 coke plant, which contains a spare gasification train (two operating and one spare), compared to the Subtask 2.3, which does not contain a spare train. Thus, on a daily average basis, the Subtask 2.3 plant uses a significant amount of higher priced natural gas (compared to coal) to increase export power production. Finally, Subtask 2.2 uses CO_2 instead of steam as diluent for NOx control in the combustion turbine which further increases the export power production from Subtask 2.2.

Figure V.7

Return on Investment vs. Power Price Showing the Effect of the Coal Price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction



The sensitivities of individual component prices and some financial parameters on the return on investment for the Subtask 2.3 coal power plant with liquids coproduction are given in Table V.3 of Appendix C

After commissioning all plants undergo a shakedown periods during which problem areas are corrected, inadequate equipment is repaired or replaced, and adjustments are made. Also as multiple plants start up and operate, the technology goes through a “learning curve” and improvements are incorporated into the next generation of plants. Consequently, performance is likely to improve as measured by increased capacity and/or improved on-stream factors. Figure V.8 shows the effect of improved syngas availability on the return on investment for the Subtask 1.6 and Subtask 2.3 plants. The abscissa is the single train syngas availability; i.e., that percentage of the time that one syngas train will be delivering syngas at the deign rate. This improved availability can be the result of “learning curve” improvements or design changes that are yet to be developed. For the Subtask 2.3 plant, as the syngas availability improves, the amount of backup natural gas is reduced until it disappears at the unattainable 100% syngas availability. At the expected 75.7% single train syngas availability, the Subtask 2.3 plant with backup natural gas requires power selling price of 42.02 $/MW-hr with 30 $/bb F-T liquids selling price to generate a 12% ROI. At an 80% syngas availability, the required power selling price drops by almost 2 $/MW-hr to 40.1 $/MW-hr. At the unattainable 100% syngas availability, no backup natural gas is required, and the required power selling price for a 12% ROI is 31.6 $/MW-hr.

Without backup natural gas, at the expected 75.7% single train syngas availability, the Subtask 2.3 plant requires a power selling price of 48.1 $/MW-hr for a 12% ROI with 30 $/bbl F-T liquids. At an 80% syngas availability, the required power selling price drops by almost 4 $MW-hr to 44.3 $/MW-hr. At the unattainable 100% syngas availability, it is the same as the case with backup natural gas, 31.6 $/MW-hr, since at this point, no backup natural gas is required.

Figure V.8

Effect of Syngas Availability on the Required Power Selling Price for a 12% Return on Investment for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Power Plant

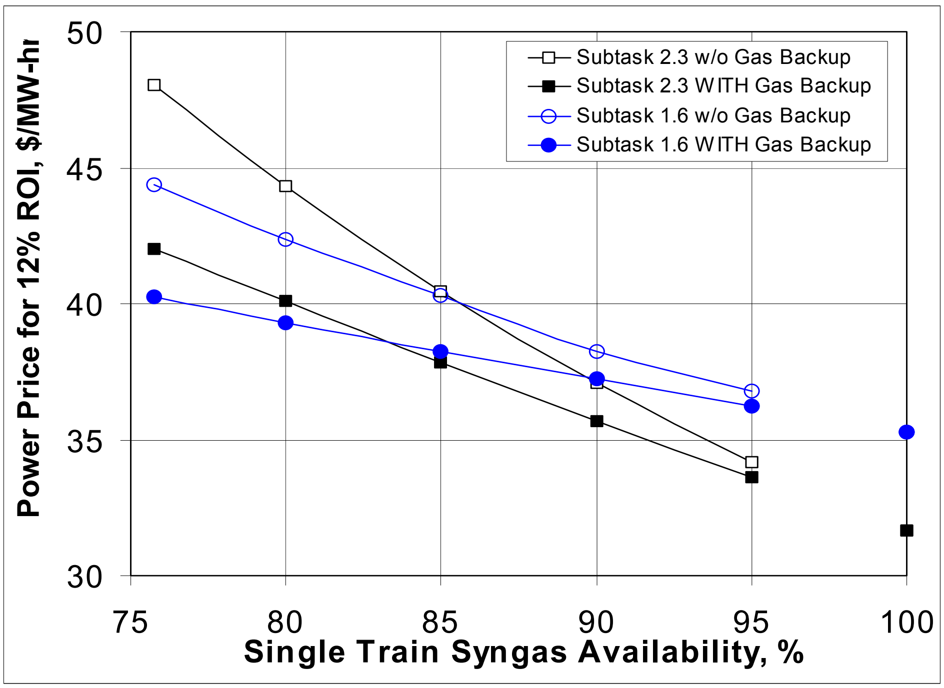


Figure V.8 also shows similar curves for the Subtask 1.6 coal IGCC power only plant. At high syngas availabilities (above 85%) without backup natural gas, the Subtask 2.3 coproduction plant design requires a lower power selling price for a 12% ROI than the Subtask 1.6 plant. With backup natural gas, this crossover point (at about 84% syngas availability) is at a required power selling price of about 38 $/MW-hr. This is the same power price where the Subtask 1.6 curve intersects the Subtask 2.3 curve with a 30 $/bbl liquids price in Figure V.6.

V.7 Effect of a Spare Gasification Train on Plant Performance

One way to increase availability and to improve the daily average production from the plant at minimal extra cost is to enlarge the gasification capacity of each train by one third so that each train now is 33.3% of the total design capacity of the plant. If this is done, then the gasification section of the plant becomes a three train facility with a spare train. With this redesign, now there are three air separation units supplying oxygen to three operating gasification trains. Table V.6 shows the effect of this redesign on the daily average feed and product rates for the coal gasification plant with liquids coproduction with backup natural gas.

Table V.6

Design and Daily Average Feed and Product Rates for Two Train Configurations of the Subtask 2.3 Optimized Coal Gasification Plant with Liquids Coproduction

	Design	4 x 25% Gasification Trains		3 x 33.3% Gasification Trains	
		Daily Avg. Rate	% of Design	Daily Avg. Rate	% of Design
Feeds					
Coal, dry tpd	9,266	6,929	74.8%	8,097	87.4%
Natural Gas, Mscf/hr	0	1,103	---	345	---
Products					
Export Power, MW	675.9	613.7	90.8%	618.9	91.6%
F-T Liquids, bpd	12,377	10,397	84.0%	11,260	91.0%
Sulfur, tpd	236.5	176.9	74.8%	206.7	87.4%
Slag, tpd (15% water)	1,423	1,064	74.8%	1,244	87.4%

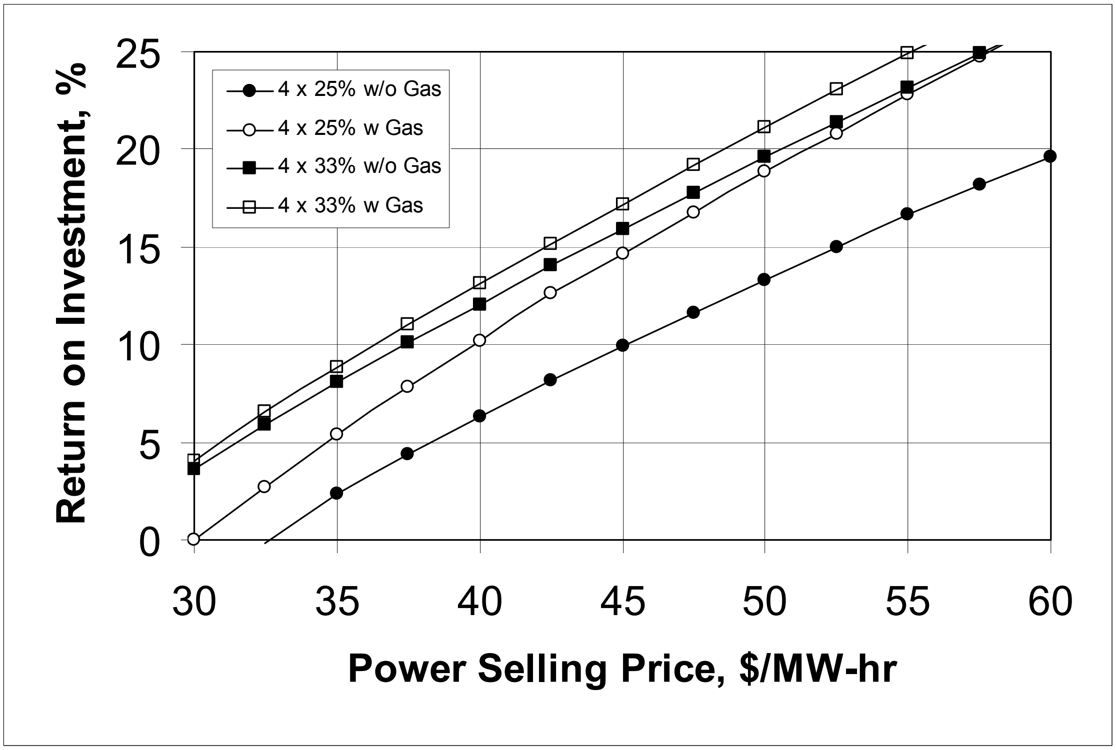
Increasing the capacity of each gasification train by 33.3% to create a spare train configuration increases the daily average coal consumption rate by over 1,000 tpd of dry coal to 8,097 tpd (87.4% of design capacity) from 6,929 tpd (74.8% of design capacity). The increased coal consumption results in increased product rates and lower backup natural gas usage rates. The daily average F-T liquids production increases to 11,260 bpd (91.6% of design) from 10,395 bpd (84% of design). The daily average power production does not increase as much, only to 618.9 MW (91.0% of design) because syngas is more available, and it, rather than backup natural gas, is used to generate power. The increase in the byproduct sulfur and slag production rates are directly proportional to the increase in the coal consumption rate.

Increasing the size of the four gasification trains from 25% to 33.3% of design capacity was estimated to increase the plant cost by about 43 MM$ to 1,202.06 MM$.

Figure V.9 shows the effect of the size and number size of gasification trains on the return on Investment versus power price for the 4 x 25% and 4 x 33.3% cases, both with and without backup natural gas. The two 4 x 33.3% cases each have higher returns than their corresponding 4 x 25% case. Without backup natural gas at a power selling price of 40 $/MW-hr, the 4 x 33.3% case has an ROI of 12.1%, which is about 5.7 ROI percent higher than the 4 x 25% case. With backup natural gas, the ROI increase for the 4 x 33.3% case is not as great, only about 2.9 ROI percent, from 10.2% to 13.1% ROI.

Figure V.9

Return on Investment vs. Power Price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction Showing the Effect of the Number of Gasification Trains



Chapter VI

Market Potential and Future Applications

VI.1 Market Potential

Integrated Gasification Combined Cycle (IGCC) and IGCC Coproduction plants are technically and economically ready for market expansion into a domestic power and energy market whose growth has been dominated by natural gas in recent years. Over the past few years, natural gas prices have undergone some wide swings and have increased significantly. Whereas the coal and petroleum coke prices have remained stable at their previous levels. As illustrated by the results of this study, gasification technologies have achieved economic parity with conventional coal fired plants in many power generation scenarios with less environmental impact. In refinery implementation scenarios, they provide a wide range of additional benefits including beneficial utilization of petroleum coke products and facilitating independence from the natural gas market.

VI.1.1 Coal

The domestic coal-to-power market is rebounding due to market and governmental concerns with natural gas price volatility, fuel diversity, and energy independence. Events in the past few years have shown the fragile nature of the natural gas market, where regional supply restrictions led to incredible price swings and cascaded into power pricing surges and shortages. The projected price of natural gas is expected to be high over the next several years.[1] While this is causing an increase in exploration and production of domestic natural gas resources, data suggests that the accessible reserves are becoming more costly to produce, and the long-term production potential is decreasing. The environmental impacts of obtaining this production also are evoking greater debate at both the local and national levels. Furthermore, LNG imports will be increaseing

Gasification is viewed as the environmentally superior process for power generation from coal. The superior environmental performance of the Wabash River facility, which was permitted in 1993 and demonstrated from 1995 to present, remains as the benchmark for the coal industry in terms of SOx, particulate emissions, and solid waste generation. Conventional combustion technology coal fired plants that have been announced in 2001 for Kentucky and Illinois are barely equivalent to the demonstrated performance of the Wabash River facility, even with state-of-the-art clean up systems on the flue gas exhaust. NOx emission performance of gasification is tied to the combustion turbine technology, which also has made great progress since the Wabash River installation. Continued advances in turbine technology will improve the penetration of gasification plants into ozone non-attainment areas.

Furthermore, coal powered generation is seen as a key to U.S. energy independence and reducing dependence on foreign energy sources. It is, as often mentioned in the literature, our most plentiful domestic energy resource with centuries of reserves in the ground. It is in the best interests of the United Stated to effectively use this resource in an efficient manner.

[1] 'Greenspan predicts high natural gas prices into next year," Associated Press story, June 10, 2003, http://www.chron.com/cs/CDA/story.hts/business/1944409.

Both independent power producers and utilities are evaluating the gasification option for greenfield coal baseload power plants they seek to develop. While the implementation of many of these planned plants has been affected by the current economic slowdown, it seems certain that their will be a new generation of coal fired power generation after nearly two decades of minimal activity, and that gasification will be a contender for a share of this new market expansion.

Further in the future, markets will develop for the repowering of aging, environmentally pressured coal fired plants and for the refueling of recent natural gas powered combined cycle plants.

VI.1.2 Petroleum Coke

Both refineries and chemical plants in the United States also are being significantly impacted by the volatility of natural gas prices. In most refineries, the costs of their steam, hydrogen and power usage are tied to natural gas pricing. The refineries with cokers have the additional burden of having to sell or dispose of petroleum coke, the "bottom of the barrel" in the refining process. Much of this petroleum coke is sold for shipment overseas, but these markets are softening because of the additional refining and coking plants being brought onstream in the early part of this decade. World coker capacity is expected to grow from 6 million metric tons per year in 2000 to over 16 million metric tons per year by 2004 because of facilities under construction or in final planning.[2] (These totals exclude China and the former Soviet Union). Much of this capacity is in the U.S. Gulf Coast, Mexico and South America, all of which will have the tendency to depress domestic prices. Delivered petroleum coke prices fluctuate, but generally they are minimal or negative at the refinery gates.

Gasification of petroleum coke not only generates the steam, power and hydrogen that the refineries and chemical plants use, but it has the synergistic ability to eliminate the need to sell or dispose of the low value petroleum coke. A typical refinery has twice the volume of petroleum coke needed for gasification to supply its hydrogen needs, providing an excess for potential power generation, liquid fuels coproduction, and/or supplying steam for the process plant. The Task 2 portion of this study has shown the economics of coke gasification may be improved by the coproduction of liquid fuel precursors in addition to power and/or hydrogen and steam.

It appears that the first domestic commercial applications for petroleum coke gasification will be for plants that are associated with petroleum refineries and/or chemical facilities where they can co-produce hydrogen and steam in addition to electric power. Experience gained in the design and operation of either coal or coke plants will lead to additional cost reductions which will make either coal and coke IGCC power plants more competitive with current base-load power plants, especially with the current high natural gas prices.

[2] Ziesmer, Ben, "World Petroleum Coke Market Trends", presented at the Gasification Technologies Conference, San Francisco, CA, Oct. 9, 2000.

VI.2 Environmental Drivers

Gasification technology is expected to have a significant share of the future power market because

1. It is a “clean”, environmentally friendly process,
2. It can accept various low-cost feedstocks, such as petroleum coke, biomass and wastes,
3. Syngas, the intermediate product, is a versatile feedstock for the production of various chemicals, such as hydrogen, methanol, acetic acid, etc.,
4. It can capture most of the pollutants, such as sulfur, carbon dioxide, hydrocarbons, and particulates, and
5. It has the potential to achieve 60% or higher thermal efficiency for power production by integration with fuel cells, advanced turbines, and hydrogen-fed turbines.

Global Energy has demonstrated the flexibility of their gasifier to handle both coal and petroleum coke. The Subtask 1.3 Next Plant, Subtask 1.5, Subtask 2.1 and Subtask 2.2 results confirm that the use of low cost petroleum coke can improve the overall economics of a gasification project. Furthermore, these results also demonstrate that coproduction of hydrogen and/or liquid fuel precursors and power may enhance the overall economic of the project.

Recently, there have been changes in the power market that are favorable to the use of IGCC plants for power generation. The deregulation of the utility industry brings a different set of power plant owners who may be eager to ally with non-utility plant owners in developing cogeneration projects. These owners are more comfortable with the complexity of IGCC plants. The use of low value feedstocks and the synergistic effects of coproduction have improved the overall economics of IGCC projects. As more of these cogeneration plants are built and operated, the capital and operating costs of the IGCC plant component will drop. Project financing also will be more readily available as confidence on the overall plant performance increases due to the more positive operating experience.

Another major driver for the gasification technology is its ability to reduce the emission of pollutants. In the near-term, gasification plants can capture more than 99% of the sulfur in the feed. Particulate and NOx emissions are equal to or less than those from conventional pulverized coal plants and natural gas combined cycle plants.

Table VI.1 summarizes the emissions produced by the Subtask 1.3 Next Plant, Subtask 1.6, and Task 2 gasification plants. They are very low for plants consuming such large amounts of coal or petroleum coke. Because of the diverse nature of the products from these plants, the fairest comparison probably should be based on the amount of emissions per unit of fuel input. On this basis, the liquid fuels coproduction plants are best. The sulfur emissions are the lowest because of the extra clean up step required before the F-T hydrocarbon synthesis reactor. The CO and NOx emissions are lower because the liquid fuels leaving the plant are not yet consumed to do their final useful work, such as power a diesel engine.

In a carbon constrained environment, gasification will be the preferred power generating technology because it can produce a concentrated carbon dioxide stream that can facilitate more efficient CO_2 capture. If the syngas is treated in a water-gas shift unit where carbon monoxide is converted by reaction with water into carbon dioxide and hydrogen. If a coal-water slurry fed gasifier is employed (similar to that of Global Energy’s E-GAS™ gasifier),

Table VI.1

Enviromental Emissions Summary* of the Subtask 1.3 Next Plant, Subtask 1.6 Subtask 2.1, Subtask 2.2, and Subtask 2.3 Coal and Coke Gasification Plants

	Subtask 1.3 Next Optimixed Pet Coke IGCC Coproduction Plant	**Subtask 2.1** Coke Gasification Power Plant with Liquid Fuels	**Subtask 2.2** Optimized Coke Gasification Plant with Liquid Fuels	**Subtask 1.6** Nominal 1,000 MW Coal IGCC Power Plant	**Subtask 2.3** Optimized Coal Gasification Plant with Liquid Fuels
Total Plant Emissions					
Exhaust Flow Rate, lb/hr	8,625,800[+]	7,988,470	4,254,450	15,950,100	7,995,000
Emissions:					
SOx, ppmvd	22	21	34	15	22
SOx as SO_2, lb/hr	350	321	276	438	329
SOx as SO_2, lb/MMBtu (HHV)	0.052	0.048	0.041	0.044	0.033
NOx, ppmvd (at 15% oxygen, dry basis)	14	13	16	13	16
NOx as NO_2, lb/hr	166	136	94	275	166
NOx as NO_2, lbMM Btu (HHV)	0.025	0.020	0.014	0.028	0.017
CO, ppmvd	13	10	10	10	10
CO, lb/hr	89	66	37	131	65
CO, lb/MMBtu (HHV)	0.013	0.010	0.006	0.013	0.007
VOC and Particulates, lb/hr	NIL	NIL	NIL	NIL	NIL
Opacity	0	0	0	0	0
Sulfur Removal, %	99.4	99.5	99.6	98.9	99.5

* Expected emissions performance

the effluent stream from the water-gas shift unit will contain about 49 mole percent hydrogen and 42 mole percent carbon dioxide.

Thus, gasification technology is an attractive choice for utilizing today's low cost feedstocks, such as coal, coke or possibly biomass, to produce clean power with or without the coproduction of selected byproducts, such as liquid fuel precursors, hydrogen or other chemicals. It has a high thermal efficiency. Furthermore, it can easily be modified to meet the challenges from future regulations related to greenhouse gas emissions.

VII.3 Future Applications

The results of this study by Bechtel, Global Energy and Nexant provide a firm basis for defining IGCC power and coproduction plant performance and cost basis needed for development of additional expansion into these markets. The results of this cooperative effort, that started with examination of the Wabash River construction cost database, has produced profiles of competitive gasification based facilities for several markets and multiple timeframes.

The strongest drivers for the implementation of gasification are its favorable environmental performance and efficiency for utilization of solid fuels. These factors make it a viable alternative to the conventional coal combustion technologies that historically have been more widely utilized.

Achievement of the installed cost goals through application of the optimization techniques that were used in the study will be realized in the first plants built, and they will provide a demonstrated basis for additional projects and an impetus for further improvements. Expanding the confidence of installed cost numbers gained in the study will bring additional commitments from other prospective customers. Operating costs already have been demonstrated to a great extent at Wabash River, and as more experience is gained, further reductions are to be expected.

The importance of the petroleum coke gasification applications to this generation of projects cannot be underemphasized. These projects, utilizing low cost petroleum coke as a feedstock and producing higher value coproducts will be the first to enter the marketplace since several of these have already started development. Wabash River already has demonstrated petroleum coke gasification at a commercial scale over a sustained period of time. New plants will help demonstrate the integration with petroleum refineries and the attainment of the necessary operating levels required to support refinery operations. New standards for capital costs and operating costs will be set as well. These petroleum coke plants, which will be the leaders of the next generation of gasification applications, will support the technology and confirm the economics for coal fueled IGCC power plants with/without coproduction that will follow them.

As crude oil supplies dwindle and both petroleum and natural gas become more expensive, a petroleum coke gasification plant with liquid fuels coproduction could become a reality. Such as plant would provide an economical way for an oil refinery to dispose of the byproduct coke and simultaneously convert it into one of its major products. Furthermore, coke gasification would allow the refinery to stabilize its power, and possibly hydrogen costs, from fluctuations in the natural gas and power markets. This plant also would provide operating experience for development of improved gasification plant designs for both the power industry and could also lead to byproduct liquids production.

Future federal and state incentive programs that are aimed at increasing the fuel diversity of our power generation resources may enhance the economics of the coal-to-power IGCC facilities. As such, these programs also tend to have a stabilizing effect on power prices because it would reduce their dependence on natural gas.

Summary, Conclusions and Recommendations

VII.1 Summary

Gasification systems are inherently clean, relatively efficient, and commercially available for converting inexpensive fuels such as coal and petroleum coke into electric power, steam, hydrogen, and chemicals. However, the gasification system also is relatively complex and costly.

This study concerned the optimization of coal and petroleum coke gasification systems to reduce the cost of power and associated co products primarily by reducing the plant cost. It shows the potential of IGCC based systems to be competitive with, if not superior to, conventional combustion based power systems because of their higher efficiency, superior environmental performance, and competitive cost.

Task 1 was divided into nine basic subtasks. Subtasks 1.1 and 1.2 developed non-optimized designs for coal and coke IGCC power and coproduction plants. Subtasks 1.3 through 1.7 and 1.3 Next Plant developed optimized designs for coal and coke IGCC power and coproduction plants. Subtask 8 performed a review of warm gas cleanup systems. Subtask 1.9 documented the availability analysis study (and results) that was performed as part of the Value Improving Practices portion of the optimization efforts.

For each case, detailed process simulation models were developed providing elementally balanced mass and heat balances. From these balances, P&IDs, equipment sizes, line sizes, and plant layouts were developed for each case. Coupled with the actual Wabash River cost data, this information allowed detailed cost estimates to be developed with a low degree of uncertainty. This detailed information is confidential.

Task 2 was divided into three subtasks. These subtasks dealt with converting two of the optimized plants developed during Task 1 into IGCC power plants with liquid fuels coproduction.

Subtask 1.3 Next Plant developed an optimized design, cost estimate and economics for a Petroleum Coke IGCC Coproduction Plant processing about 5,417 tpd of dry petroleum coke and producing about 80 MMscfd of hydrogen and 980,000 lb/hr of industrial-grade steam (750^{o}F/700 psig) in addition to electric power. The Subtask 1.3 Next plant produced 474 MW of export power and 373 tpd of sulfur. It has an EPC cost of 787 MM mid-year 2000 dollars.[1]

Starting from the Subtask 1.3 Next Plant, Subtask 2.1 developed a petroleum coke gasification power plant with liquids coproduction by eliminating the export steam and hydrogen production facilities and replacing them with a single-train, once-through Fischer-Tropsch hydrocarbon synthesis plant. A once-through system eliminates the cost of the expensive recycle system which requires recycle gas purification facilities in addition to the recycle compressor. The energy that was used to produce the export steam now is used to generate additional power. This plant produces 617 MW of export power and 4,125 bpd of liquid fuel precursors from slightly less petroleum coke (5,376 vs. 5,417 dry tpd) than the

[1] All reported costs are mid-year 2000 costs. They are presented here to show the relative differences between the cases. Current cost estimates should be developed for any proposed application.

Subtask 1.3 Next Plant. On a higher heating value (HHV) basis, this plant has a thermal efficiency 47.9% when the heating value of the byproduct sulfur is included. It cost 818 MM mid-year 2000 dollars.

Subtask 2.2 developed an optimized design for a petroleum coke gasification power plant with liquids coproduction by maximizing the liquid fuel precursors production at the expense of power production. In this design, about 92% of the syngas goes through the once-through slurry-bed F-T hydrocarbon synthesis reactor. The unconverted syngas and light hydrocarbons from the F-T area are mixed with the remaining 8% of the syngas, compressed, and sent to the single gas turbine for power generation. This plant produces 10,450 bpd of liquid fuel precursors and 367 MW of export power from 5,417 tpd of dry petroleum coke. It has an EPC cost of 735 MM mid-year 2000 dollars. On a higher heating value basis, this plant has a thermal efficiency 56.7% when the heating value of the byproduct sulfur is included and 54.9% when the byproduct sulfur is not included. With 27 $/MW-hr and 30$/bbl liquids, this plant has a 18.2% ROI, and the Subtask 2.1 plant only has a 9.50% ROI. (Both cases assume an 80% loan rate at 10% annual interest.

Subtask 1.6 developed a current day optimized design, cost estimate and financial analysis for a nominal 1,000 MW coal fed IGCC power plant using four gasifiers and four GE 7FA+e combustion turbines. The plant consumes 9,266 tpd of dry Illinois No. 6 coal and generates 1,155 MW of export power. It cost 1,231 MM mid-year 2000 dollars (1,066 $/kW) and can export power at 44.4 $/MW-hr without natural gas backup while producing a 12% ROI. With 2.60 $/MMBtu backup natural gas, the required power selling price for a 12% ROI drops to 40.2 $/MW-hr. On a higher heating value (HHV) basis, this plant has a thermal efficiency 42.4% when the heating value of the byproduct sulfur is included.

The Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction was developed from the Subtask 1.6 plant using the design approach adopted for the optimized Subtask 2.2 coke plant. The coal gasification capacity of the plant was kept the same as Subtask 1.6. F-T liquids production was maximized, and power production was reduced to only one power block train consisting of two combustion turbines, two HRSGs, and a single steam turbine. The unconverted syngas and light hydrocarbons from the F-T synthesis section is compressed and combined with the 18% of syngas bypassing the F-T reactors to provide fuel for the two combustion turbines.

The plant produces 12,377 bpd of liquid fuel precursors, 675.9 MW of export power, and 237 tpd of sulfur from 9,266 tpd of dry Illinois No. 6 coal. This plant has an EPC cost of 1,159 MM mid-year 2000 dollars. On a higher heating value (HHV) basis, the plant has a thermal efficiency of 53.4% when the heating value of the byproduct sulfur is included. This thermal efficiency is lower than that of the Subtask 2.2 optimized petroleum coke coproduction plant because this plant produces less liquid fuel and more power on a relative basis that the coke plant. With 30 $/bbl liquids and 2.60 $/MMBtu natural gas, this plant requires a power selling price of 42 $/MW-hr to produce a 12% ROI; whereas the Subtask 1.6 plant requires a power selling price of only 40.2 $/MW-hr.

Enlarging the gasification train capacity of the coal plant by 33% so that the plant would have three operating trains and a spare gasification train to make it similar to that of the petroleum coke case, would improve the ROI by about 6 to 8%. With 30 $/bbl liquids, the plant still would require power selling prices of 40 plus $/MW-hr to justify building the facility.

As more IGCC plants, either with or without coproduction facilities, are built and operated, availability should improve which will increase the plant ROI at given power price, or lower the required product selling prices for a given ROI. At low power prices relative to oil prices, IGCC power plants with liquid fuels coproduction will be favored, and conversely when power prices are high relative to oil prices, IGCC power only power plants will be preferred.

Based on the above results, in order for a gasification power plant with liquids coproduction to have a better ROI than a conventional IGCC power plant, the plant design must be balanced. Some features that contribute to this balanced design include

- The use of large, cost efficient gasification trains to minimize cost
- Inclusion of a spare gasification train for maximum availability
- The syngas should have high CO and H_2 contents and a low methane content to allow the F-T area to produce an offgas with a minimal Btu content.
- High conversion in the F-T section so that it can produce an offgas with a high CO_2 content for NOx control
- The ability to process all, or almost all, of the syngas in the F-T reactors
- A large, efficient combustion turbine that is correctly sized to process all the fuel gas with minimum additional steam dilution for NOx control

The Subtask 2.2 Optimized Petroleum Coke IGCC Power Plant with Liquids Coproduction does a good job of satisfying most of the above criteria. However, the Subtask 2.3 coal plant produces a syngas with a methane content that is about 2.6 times greater than the syngas produced by the gasification of coke because of the higher volatiles content of the coal. As a result, the F-T offgas has a higher Btu content and requires more steam dilution for NOx control. Furthermore, the total amount of F-T offgas contains too much energy for one GE 7FA+e turbine, and not enough for two turbines. Consequently, about 18% of the syngas has to be bypassed around the F-T reactors to fully load the two GE 7FA+e turbines. This significantly reduces the liquids production. Ideally, a single larger turbine [or two smaller turbines] that would require bypassing only very little, if any, syngas around the F-T reactors would result in a better balanced plant that could have a better return on investment.

The balanced approach in which the gas turbine fuel gas is diluted with CO_2 to a level where only minimal or no additional steam dilution for NOx control also could be applied to an ICGG power plant that co-produces hydrogen (instead of liquid fuels) for power generation with fuel cells. In such a plant, CO_2 production by the shift reaction in excess of that needed for NOx control would be captured for possible sequestration.

Gasification is viewed as the environmentally superior process for power generation from coal. The Wabash River facility demonstrated the superior environmental performance of gasification in terms of SOx, NOx, and particulate emissions. In a carbon-constrained environment, the CO_2 easily can be captured for sequestration or other uses. Even without CO_2 capture, CO_2 emissions are reduced because gasification plants are more efficient than conventional coal power plants.

With low coal and coke prices and high oil prices, the return of a gasification power plant can be improved by adding hydrocarbon liquids coproduction. This is especially true for a coke plant associated with a petroleum refinery because besides providing a means of disposing of the byproduct coke, the plant can convert it into liquid hydrocarbons, which when upgraded in the refinery become the main refinery products, liquid transportation fuels.

As natural gas and power prices increase and environmental constraints for coal fired generation plants tighten, coal IGCC will further penetrate the power market. As more coal and coke IGCC plants are built, further improvements can be expected which should lead to additional cost reductions and improved availability that will make IGCC the preferred option for new base-load power plants.

VII.2 Conclusions

Tasks 1 and 2 of this study have shown that:

- Optimization of IGCC plants has resulted in significant capital and operating cost savings.
- Additional cost savings appear likely as some of the concepts developed in this study are researched, developed and implemented.
- The Value Improving Practices used in this study provided a structured method for reducing both the plant cost as well as the operating and maintenance costs.
- Substantial capital cost reductions can be obtained by optimization of the plant layout to reduce the plant size.
- Petroleum coke-fueled IGCC coproduction plants are economically competitive in today's economic environment.
- Power generation by gasification of coal is not yet competitive with coal combustion plants, but the gap has narrowed substantially. Further developments will make IGCC competitive.
- Petroleum coke- and coal-fueled IGCC power plants are very similar. There are differences, but the costs of the two plants are similar.
- Information from the design, construction and operation of petroleum coke gasification plants will further the development and commercialization of coal-fueled plants.
- The ROI of an IGCC power plant can be lowered by the use of backup natural gas to fire the gas turbine when syngas is unavailable.
- As natural gas prices increase, coal-fueled IGCC power plants will be favored over gas-fired combined cycle plants.
- Coproduction of liquid fuels can enhance the economics of IGCC power plants when oil prices are high and power prices are low.
- The balance between power and liquid fuels coproduction to produce an optimum plant design depends upon syngas composition and F-T offgas composition to produce a gas turbine fuel and turbine design in order to minimize steam dilution for NOx control.

VII.3 Recommendations

Technology development will be the key to the long-term commercialization of gasification technologies and integration of this environmentally superior solid fuels technology into the existing mix of power plants. Task 1 of this study recommended that further development of the following area would be beneficial:

- Development of the "G/H-class" combustion turbine for syngas applications
- Gasifier advancements including slurry feed vaporization in the second stage

- Demonstration of warm gas clean-up technologies (e.g., SCOHS)
- Testing of advanced wet and dry filtration systems
- Development and implementation of large capacity fuel cells; optimization of the integration of gasification with advanced fuel cell processes
- Further advances in Fischer-Tropsch technology or other gas-to-liquids technologies for the production of liquid transportation fuels from coal
- Develop a lower cost means of producing oxygen such as the ITM ceramic membrane system

As a result of this Task 2 study, additional research and development efforts in the following areas also would be beneficial:

- Equipment modifications and revised operating procedures should be developed to improve the overall plant availability. Such items could include the development of longer lasting refractory, improved more durable burner designs, better heat recovery equipment, and better filtering systems for solids separation.
- Improved F-T catalysts that produce a product distribution with less methane and light ends and can operate at high once-through conversions
- Verification that activated carbon can be used to reduce the residual sulfur in the syngas going to the F-T reactor to a level where it will not be detrimental to catalyst performance
- Investigate the use of a slurry-bed F-T reactor with iron catalyst followed by another F-T reactor (either slurry or fixed bed) to maximize carbon monoxide and hydrogen conversion to produce a gas turbine fuel gas that does not require dilution for NOx control
- Develop a design for a balanced IGCC Coproduction power plant that co-produces hydrogen for power production by fuel cells based on the balanced approach developed in this study. Excess CO_2 produced by the shift reaction above that required for NOx control in the gas turbine would be captured for possible sequestration.

Chapter VIII

List of Acronyms and Abbreviations

ºC	degrees Celsius
ºF	degrees Fahrenheit
$	United Stated dollars
$/bbl	United States dollars per barrel
$/kW	United States dollars per kilowatt
$/MMBtu	United States dollars per million British thermal units
$/Mscf	United States dollars per thousand standard cubic feet
$/MW-hr	United States dollars per megawatt hour
$/ton	United States dollars per ton
%	percent
%/yr	percent per year
AGR	acid gas removal
ASU	air separation unit
Atm	atmosphere(s)
bbl	barrels
bpd	barrels per day
bpsd	barrels per stream day
BFW	boiler feed water
Btu	British thermal unit(s)
Btu/scf	British thermal units per standard cubic foot
CEM	continuous emission monitoring
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
DOE	Department of Energy
EIA	Energy Information Agency
E-GAS™	name of the gasification technology at Wabash River when this project started
EPA	Environmental Protection Agency
EPC	engineering, procurement and construction
EPRI	Electric Power Research Institute
F-T	Fischer-Tropsch

ft, ft^2, ft^3	foot (feet), square feet, cubic feet
gal	gallon(s)
GE	General Electric
GT	gas turbine
GTG	gas turbine generator
GTL	gas to liquids
H_2O	water
H_2S	hydrogen sulfide
HHV	higher heating value
HP	high pressure
hr	hour(s)
HRSG	heat recovery steam generator
HTRU	high temperature heat recovery unit
HV	high voltage
IGCC	integrated gasification combined-cycle
in, in^2, in^3	inches, square inches, cubic inches
IP	intermediate pressure
IRR	internal rate of return
kg	kilogram
KO	knock out
kW	kilowatt
kW-hr	kilowatt-hour
kV	kilovolt
lb	pound(s)
lb/hr	pounds per hour
lb/MMBtu	pounds per million British thermal units
lb/MW-hr	pounds per megawatt hour
LP	low pressure
L/V	liquid/vapor
LHV	lower heating value
LPG	liquefied petroleum gas
LTHR	low temperature heat removal
M$	thousands of United States dollars
MCC	motor control center
MDEA	methyldiethylamine, a chemical
min	minute(s)
Mlb	thousands of pounds
Mlb/hr	thousands of pounds per hour
MMlb	millions of pounds
MM	million(s)
MM$	millions of United States dollars

MMBtu	millions of British thermal units
MMBtu/bbl	millions of British thermal units per barrel
MP	medium pressure
Mscf	thousands of standard cubic feet
Mscf/hr	thousands of standard cubic feet per hour
MW	megawatts
MW-hr	megawatt-hours
NOx	nitrogen oxides
NPV	piping and instrument drawing
PH	a measure of acidity
PLC	programmable logic controller
ppm	parts per million
ppmv	parts per million by volume
ppmvd	parts per million by volume dry
PSA	pressure swing adsorption
psia	pounds per square inch absolute
psig	pounds per square inch gauge
ROI	return on investment
ROM	run of mine
S/C	subcontract
scf	standard cubic foot (feet) at 60ºF and 1 atmosphere
scfm	standard cubic feet per minute
SCOHS	selective catalytic oxidation of hydrogen sulfide
SO_2	sulfur dioxide
SOx	sulfur oxides
ST	steam turbine
STG	steam turbine generator
SRU	sulfur recovery unit
tpd	tons per day
VIPs	value improving practices
VOC	volatile organic compounds
vol	volume
wt	weight
yr	year
ZLD	zero liquid discharge
ZnO	zinc oxide, a chemical
ZnS	zinc sulfide, a chemical

Chapter IX

Acknowledgements

The help, assistance, guidance, and support of the following individuals and organizations is hereby acknowledged.

The United States Department of Energy's National Energy Technology Laboratory (DOE/NETL) for their financial support

John Rockey of DOE/NETL for his inspiration, guidance and promotion of this study

Gary Stiegel of DOE/NETL for his ongoing support

The managements of Bechtel, Global Energy and Nexant for their backing and financial support

General Electric for their assistance with the performance of the combustion turbine

Praxair for their assistance with the performance of the air separation unit

UOP for their assistance with the performance of the PSA unit

Aspen Technology, Borsig, Callidus Technologies, Emtrol, Nooter, Süd-Chemie, Svedala, USFilter, and other equipment manufactures and suppliers for their assistance with designs, performance specifications and cost quotations

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Prepared for:
United States Department of Energy National Energy Technology Laboratory
Contract No. DE-AC26-99FT40342

Gasification Plant Cost and Performance Optimization

Task 2 Topical Report Coke/Coal Gasification With Liquids Coproduction

Volume 2 Appendices A through C

Submitted By:

BECHTEL

GLOBAL ENERGY

Nexant

SEPTEMBER 2003

U. S. Department of Energy
National Energy Technology Laboratory (NETL)

Gasification Plant Cost and Performance Optimization
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Task 2 Topical Report
Coke/Coal Gasification
With Liquids Coproduction

Volume 2 of 2
Appendices A through C

September 2003

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Appendix A

Subtask 2.1

[Non-Optimum] Petroleum Coke Gasification

Power Plant with Liquids Coproduction

Subtask 2.1

Executive Summary

A design for a [non-optimum] coke gasification power plant with liquid fuel precursors coproduction using Fischer-Tropsch technology has been developed. Subtask 2.2 will optimize the design of the plant. The plant consumes 5,375 tpd of coke (dry basis), 110 tpd of flux and 23.2 MMBtu/hr (HHV) of natural gas to produce 617 MW-hr of export power and 4,125 bpd of liquid fuel precursors. It also produces 371 tpd of elemental sulfur and 194 tpd of slag. The plant is located on the U.S. Gulf Coast adjacent to a petroleum refinery.

The design of the plant was developed from the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant that processes about the same amount of coke and produces 80 MMscfd of 99% hydrogen and 980,000 lb/hr of 750°F/700 psia steam for the adjacent refinery in addition to 474 Mw-hr of export power. It also produces about the same amounts of byproduct sulfur and slag. The Subtask 1.3 design was modified by replacing the hydrogen facilities with the F-T hydrocarbon synthesis area. The unconverted syngas and light hydrocarbon products from the F-T area are compressed and sent to the combustion turbines to generate power. A larger steam turbine is used to generate additional power from the steam that previously was sent to the refinery.

The Fischer-Tropsch hydrocarbon synthesis area basically consists of three sections; final sulfur removal, slurry-bed F-T reactor, and product recovery sections. The final sulfur removal section which only treats the syngas going to the F-T reactor contains a second COS hydrolysis reactor which converts the residual COS in the syngas to H_2S. The H_2S is removed from the syngas by ZnO adsorption beds. The sulfur-free syngas is fed to the slurry-bed F-T reactor which converts syngas to hydrocarbons over an iron-based catalyst. The heat of reaction is removed by generation of 440°F/375 psia steam inside tubes that are placed within the slurry-bed. The lighter hydrocarbon products and unconverted syngas leave the reactor as vapors and are cooled to recover the condensed hydrocarbons as liquids. The unconverted syngas and non-condensable light hydrocarbons (primarily C1 through C3s) are compressed and sent to the combustion turbines for power generation. The heavier products are removed from the reactor as liquids, separated from the entrained catalyst by filtration, cooled, mixed with the lighter hydrocarbons, and sent to the adjacent petroleum refinery for separation and incorporation into liquid transportation fuels.

The F-T liquid fuel precursors essentially are a bottomless, sulfur-free crude oil. Basically they are straight-chain 1-olefins and paraffins without any aromatics. The diesel fraction has a very high cetane number (>70) and is a premium diesel fuel blending component. The naphtha fraction is a low octane material that requires further upgrading for use as a gasoline blending component. However, it is an excellent feedstock for an ethylene cracker. Linear programming studies have shown that the F-T liquid fuel precursors may be worth up to 10 $/bbl more than crude oil depending upon the specific refinery configuration and product demands.

This coke plant with liquid fuel precursors coproduction has a slightly better return on investment than the Subtask 1.3 Next Plant. At export power prices below 30 $/MW-hr, this plant can have a return on investment greater than 12% when the liquid fuel precursors are worth 30 $/bpd.

Under normal operations, only a minimal amount of natural gas is used for furnace fuel. When insufficient syngas is available, natural gas is used to supplement the available syngas to maintain full power production. However, it may be more profitable to operate at reduced export power capacity or to suspend liquids production depending upon the relative prices for the liquid fuel precursors, export power and natural gas.

During development of this design, several ideas were generated for optimizing the process, such as using activated carbon to remove the sulfur from the syngas going to the F-T reactor, using a refrigeration system to increase the liquids recovery, and using a larger F-T reactor to produce more liquids at the expense of power production. These ideas will be incorporated in the Subtask 2.2 plant design.

Subtask 2.1
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Attachment

Section 1

Introduction

The objective of this Gasification Plant Cost and Performance Optimization Project is to develop optimized engineering designs and costs for several Integrated Gasification Combined Cycle (IGCC) plant configurations. These optimized IGCC plant systems build on the commercial demonstration cost data and operational experience from the Wabash River Coal Gasification Repowering Project.[1] The Wabash River Repowering Project contains an E-GAS™ gasifier that processes a nominal 2,500 TPD of as-received coal to produce clean syngas for a GE 7A combustion turbine and steam for repowering an existing steam turbine.

Task 1 of this IGCC Plant Cost and Performance Optimization study consists of the following nine subtasks:

- Subtask 1.1 – Expand the Wabash River Project facility design to a greenfield unit
- Subtask 1.2 – Petroleum Coke based IGCC plant with the coproduction of steam and hydrogen
- Subtask 1.3 – Optimized petroleum coke based IGCC plant with the coproduction of steam and hydrogen
- Subtask 1.4 – Optimized coal to power IGCC plant
- Subtask 1.5 – Comparison between single-train coal and coke fueled IGCC power plants
- Subtask 1.6 – Optimized coal fueled 1,000 MW IGCC power plant
- Subtask 1.7 – Optimized single-train coal to hydrogen plant
- Subtask 1.8 – Review the status of warm gas clean-up technology applicable to IGCC plants
- Subtask 1.9 – Discuss the Value Improving Practices availability and reliability optimization program

Task 1 has been completed. The Task 1 Topical Report was issued to the Department of Energy on May 30, 2002.[2]

Task 2 has the objectives of developing optimum plant configurations for IGCC power plants with the coproduction of liquid fuel precursors. Task 2 is divided into the three subtasks.

Subtask 2.1 – [Non-Optimum] Petroleum Coke Gasification Power Plant with Liquids Coproduction Starting with the same petroleum coke and gasification plant designs generated in the Subtask 1.3 Next Optimized Coke IGCC Coproduction Plant, a design shall be developed for a coke gasification power plant co-producing liquid transportation fuel precursors containing a single-train, once-through Fischer-Tropsch (F-T) gas-to-liquids (GTL) plant. The liquid hydrocarbons from the F-T hydrocarbon synthesis section will be recovered and sent to the adjacent petroleum refinery for upgrading and blending into premium liquid transportation fuels. The unconverted syngas and non-condensable hydrocarbons from the F-T hydrocarbon synthesis section will be used for power production in the combined-cycle power block.

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[1] "Wabash River Coal Gasification Repowering Project, Final Technical Report", U. S. Department of Energy, Contract Agreement DE-FC21-92MC29310, August 2000.
[2] "Task 1 Topical Report – IGCC Plant Cost Optimization", Gasification Plant Cost and Performance Optimization, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, May 30, 2002.

Subtask 2.2 – Optimum Petroleum Coke Gasification Power Plant with Liquids Coproduction The Subtask 2.1 plant shall be optimized to develop an optimized coke gasification power plant co-producing liquid transportation fuel precursors. Optimization activities primarily will be concerned with the F-T area and overall plant integration. Since the Subtask 2.1 gasification area was developed from an optimized IGCC petroleum coke gasification coproduction plant, a review of the plant is appropriate at this time to ensure that the previous modifications are still applicable to this case.

Subtask 2.3 - Optimum Coal Gasification Power Plant with Liquids Coproduction The Subtask 2.2 plant shall be converted to a coal-fueled gasification unit using Illinois No. 6 coal, retaining the optimized portions and incorporating those optimizations developed in Subtask 1.6. This will involve combining the optimized coal gasification plant developed in Subtask 1.6 with the Subtask 2.2 plant, to develop an optimized coal gasification power plant co-producing liquid transportation fuel precursors. Because of differences in the syngas generation area and resulting syngas composition, the F-T hydrocarbon synthesis area, and overall plant integration probably will require modification.

This report describes the design, performance and economics of the Subtask 2.1 non-optimum petroleum coke gasification power plant co-producing liquid fuel precursors. Section 2 provides background information on two previous studies that are the basis for this current study. Section 2.1 describes the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. Section 2.2 briefly describes the F-T hydrocarbon synthesis reaction and provides an overview of the indirect coal liquefaction plant that was developed during the previous Department of Energy study.

Section 3 describes the Subtask 2.1 Petroleum Coke Gasification Power Plant with Liquids Coproduction that was developed in this study. Section 4 provides an availability analysis of the plant, and Section 5 give a financial analysis of plant performance. Section 6 contains a summary of this subtask and some ideas that were generated during the development of this plant design for optimizing the Subtask 2.2 plant.

An attachment summarizes the revenue calculations for the plant under different feed and product price scenarios for three potential operating scenarios.

Section 2

Background

During Task 1, several designs were developed for petroleum coke IGCC coproduction plants that supplied an adjacent petroleum refinery with 750ºF/750 psig steam and hydrogen.[2] Subtask 1.2 developed a first pass design and cost estimate for a petroleum coke IGCC coproduction plant located on the Gulf Coast based on the Subtask 1.1 Wabash River Greenfield Plant. Subtask 1.3 developed designs for three optimized petroleum coke IGCC coproduction plants, each with the same design capacity. The only difference between these plants was the amount of spare equipment inside the gasification block. The Base Case design contains two gasification trains with each gasification train having a spare gasification reactor vessel that can be placed in service when the other reactor requires refractory replacement. The minimum cost case eliminated the spare gasification vessel in each train. The spare gasification train case has three complete gasification trains beginning with the slurry feed pumps and continuing through the syngas particulate removal systems. The remainder of the facility is sized so that only two gasification trains can operate simultaneously at design capacity. The Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant is based on the Subtask 1.3 spare train case because that case has the highest return on investment. Section 2.1 describes the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant.

In 1991, Bechtel with Amoco as the main subcontractor was awarded DOE contract DE-AC22-91PC90027 to develop designs and computer process simulation models for indirect coal liquefaction plants using advanced Fischer-Tropsch Technology.[3] Subsequently, the simulation model was improved by adding additional components.[4] The Fischer-Tropsch (F-T) hydrocarbon synthesis section of this ASPEN process simulation model was used to develop the design of the F-T hydrocarbon synthesis section of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction. Section 2.2 briefly describes the F-T hydrocarbon synthesis reaction and presents an overview of the entire facility.

2.1 Subtask 2.1 Next Optimized Petroleum Coke IGCC Coproduction Plant

The Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant was developed by applying nine Value Improving Practices (VIPs) to the Subtask 2.1 plant to reduce costs and improve operability.[5] As a result of this effort, plant performance was improved, the plant cost was reduced, and the return on investment was significantly improved. The results of this VIP and optimization study included:

[3] "Topical Report – Volume I, Process Design – Illinois No. 6 Coal Case with Conventional Refining", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.
"Topical Report – Volume IV, Process Flowsheet (PFS) Models", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.
[4] "Topical Report VI – Natural Gas Fischer-Tropsch Case, Volume II, Plant Design and Aspen Process Simulation Model", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, August, 1996.
[5] "Task 1 Topical Report – IGCC Plant Cost Optimization", Gasification Plant Cost and Performance Optimization, Chapter II, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, May 30, 2002

- Simplified solids handling system
- Removal of the feed heaters and spare pumps
- Maximum use of slurry quench
- Maximum syngas moisturization
- Use of a cyclone and a dry particulate removal system to clean the syngas
- Removal of the T-120 post reactor residence vessel
- Simplified Claus plant, amine and sour water stripper
- Use of state-of-the-art GE 7FA+e gas turbines with 210 MW output and lower NOx
- Use of steam diluent in the gas turbines
- Development of a compact plant layout to minimize the use of large bore piping
- Used Bechtel's advanced construction techniques to reduce costs
- Added design features to reduce O&M costs and increase syngas availability

Table II.1 shows the design input and output streams for the Subtask 1.3 Next Plant. The plant processes 5,417 tpd of dry petroleum coke and produces 474 MW of export power. In addition, the plant exports 980,000 lb/hr of 750ºF/700 psia steam and 80 MMscfd of hydrogen to the adjacent petroleum refinery. It also produces 373.4 tpd of sulfur and 195.1 tpd of slag. No natural gas is consumed during design operations. However, the plant does use natural gas during startup and as a supplementary fuel to fire the combustion turbines when insufficient syngas is available.

The resulting design configuration for the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant is shown in Figure 2.1. The plant basically is a two train facility, but with three gasification trains, two operating and one spare. The two-train sections of the plant are sized so that they only have sufficient capacity to process the output from two gasifiers simultaneously operating at design capacity.

Figure 2.2 is a Block Flow Diagram of the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant showing the flow rates of the input/output streams and the major interconnecting streams between the process blocks.

Table II.1

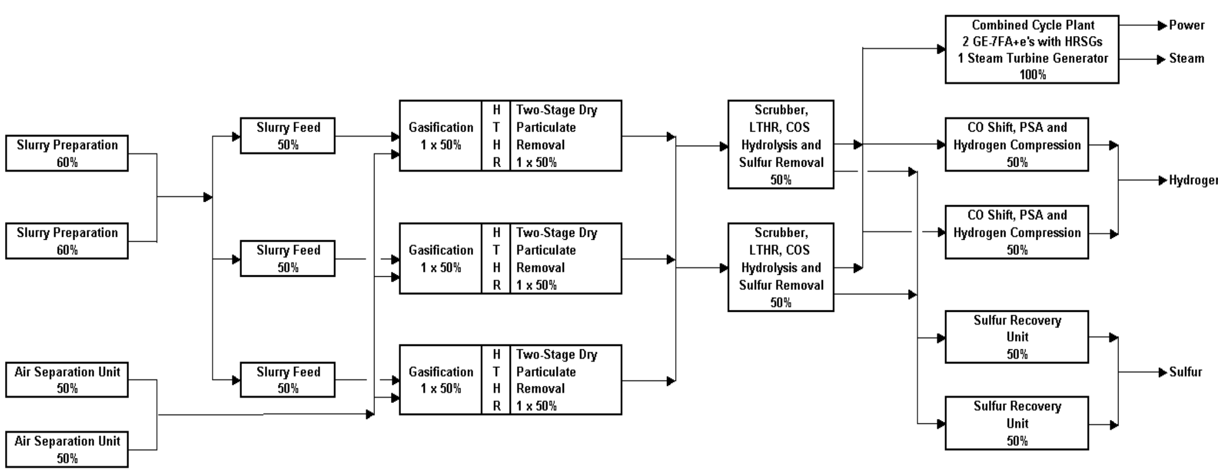
Design Input and Output Streams for the
Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

	Subtask 1.3 Next Plant
Plant Inputs	
Coke Feed, as received TPD	5,692
Dry Coke Feed to Gasifiers, TPD	5,417
Oxygen Production, TPD of 95% O_2	5,954
Total Fresh Water Consumption, gpm	5,223
Condensate Return from the Refinery, lb/hr	686,000
Flux, TPD	110.6
Natural Gas, MMBtu HHV/hr	0
Plant Outputs	
Net Power Output, MW	474.0
Sulfur, TPD	373.4
Slag, TPD (15% moisture)	195.1
Hydrogen, MMscfd	80
HP Steam, 750°F/700 psia	980,000
Fuel Gas Export, MMBtu/hr	0

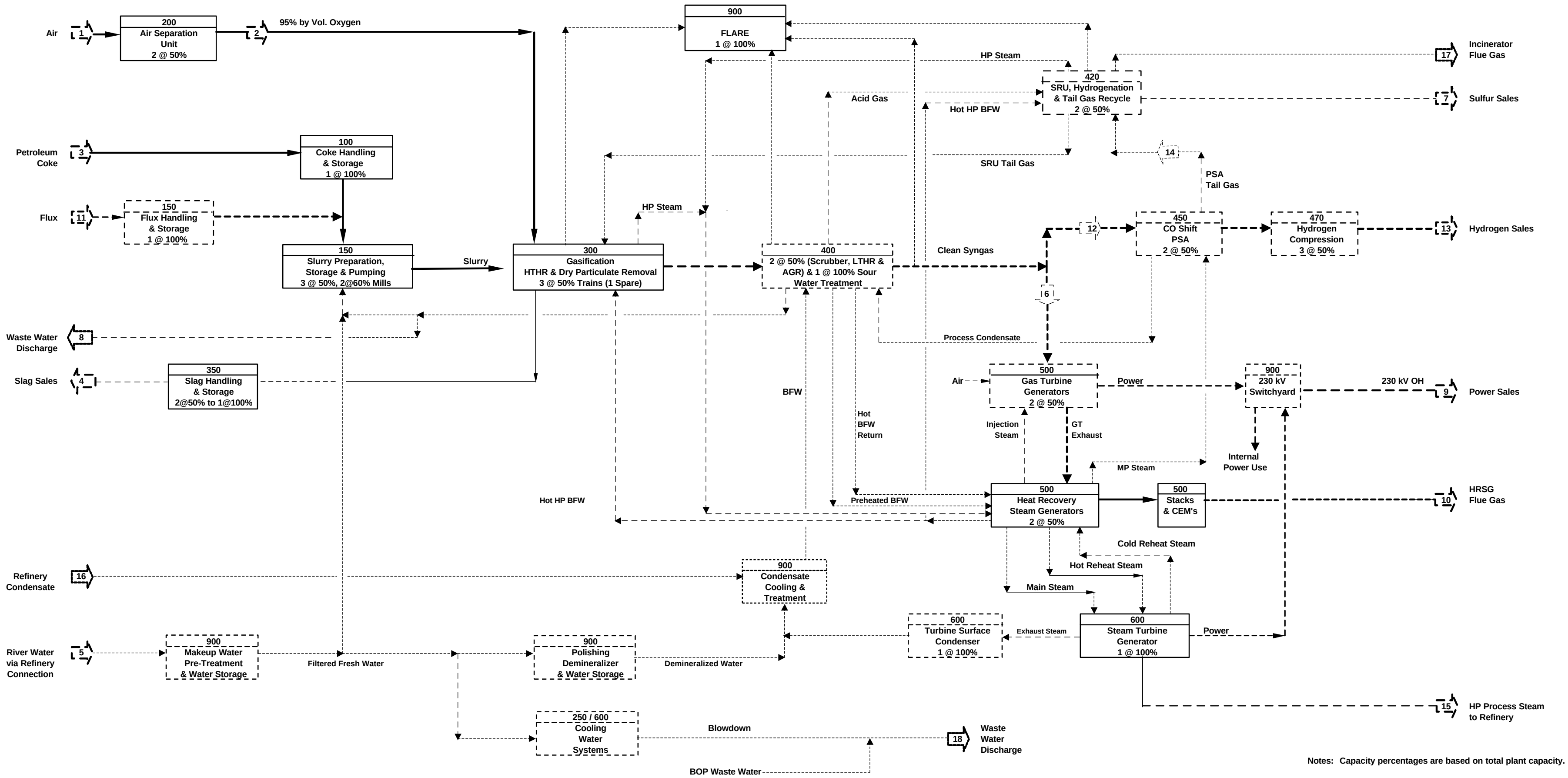
Figure 2.1

Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

Simplified Block Train Diagram



Notes: Capacity percentages are based on total plant capacity.



Notes: Capacity percentages are based on total plant capacity.

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21
	Air	Oxygen	Coke	Slag	Water	Syngas	Sulfur	Water	Power	Flue Gas	Flux	Syngas	Hydrogen	Tail Gas	HP Steam	Condensate	Flue Gas	Water			
Flow	25,961	5,954	5,417	195.1	2,611,500	1,016,830	373.4	49,177	474,000	7,966,800	110.6	363,028	80	93.4	980,000	686,000	658,750	504,000			
	Tons/Day	Tons/Day	Tons/Day	Tons/Day	Lb/Hr	Lb/Hr	Tons/Day	Lb/Hr	kWe	Lb/Hr	Tons/Day	Lb/Hr	MMSCFD	MMSCFD	Lb/Hr	Lb/Hr	Lb/Hr	Lb/Hr			
Nominal Pressure - psig	Atmos.	609	NA	NA	50	350	NA	62	NA	Atmos.	NA	350	1,000	5	700	200	Atmos.	Atmos.			
Temperature - F	70	240	Ambient	180	70	530	332	80	NA	253	NA	530	120	113	750	190	500	71			
HHV Btu/lb	NA	NA	14,848	NA	NA	3,725	NA	NA	NA	NA	NA	3,725	NA	753	NA	NA	NA	NA			
LHV Btu/lb	NA	NA	14,548	NA	NA	3,533	NA	NA	NA	NA	NA	3,533	NA	659	NA	NA	NA	NA			
Energy - MM HHV/hr	NA	NA	6,703	NA	NA	3,788	124	NA	NA	NA	NA	1,352	1,083	281	NA	NA	NA	NA			
Energy - MM LHV/hr	NA	NA	6,567	NA	NA	3,592	124	NA	NA	NA	NA	1,282	917	246	NA	NA	NA	NA			
Notes	Dry Basis	5,615 O2	Dry Basis	15%Wtr.	5,223 GPM	To GT	Sales	98 GPM	230 kV			For H2	Sales	373 MLb/hr	Sales	Return		1,008 GPM			

DOE Gasification Plant Cost and Performance Optimization

Figure 2.2

Subtask 1.3

NEXT OPTIMIZED PETROLEUM COKE IGCC

COPRODUCTION PLANT

BLOCK FLOW DIAGRAM

File: Fig 2.2.xls February 21, 2002

2.2 Fischer-Tropsch Hydrocarbon Synthesis Process

Fischer-Tropsch hydrocarbon synthesis is an old process in which synthesis gas or syngas (carbon monoxide and hydrogen) react over a catalyst to produce aliphatic hydrocarbons (principally normal paraffins and straight chain 1-olefins). It was used by Germany during the Second World War to make liquid fuels for military use. Subsequent cost reductions may have made F-T processes competitive in certain situations. Currently, there is a lot of interest in using the F-T process to monetize remote natural gas by converting it into an easily transportable synthetic crude oil that can be upgraded to liquid transportation fuels.

In general, the F-T hydrocarbon synthesis reactions for olefins and normal paraffins can be written as

$$n\ CO + 2n\ H_2 \rightarrow C_nH_{2n} + n\ H_2O$$

$$n\ CO + (2n+2)\ H_2 \rightarrow H\text{-}(CH_2)_n\text{-}H + n\ H_2O$$

As seen from the above reaction stoichiometry, the ideal syngas composition is just over 2 moles of hydrogen for each mole of carbon monoxide.

The reaction is very exothermic. Traditionally, at a large scale the reaction has been performed over solid catalyst that is placed in small diameter tubes immersed in a cooling medium (such as boiling water) to remove the heat of reaction. The hydrocarbon product yield distribution can be characterized by a Schultz-Flory distribution in which the molar ratio of a component containing n carbon atoms to one with n+1 carbon atoms is a constant called alpha (α). As the reaction temperature increases, the yield distribution shifts to lighter hydrocarbons; i.e., the α parameter gets smaller. As time has progressed, more sophisticated mathematical yield models using multiple α parameters have been developed to represent the F-T reaction yields.

In the 1950s, the slurry-bed reactor was developed in which fine catalyst particles are suspended in a liquid, and the reactant syngas is bubbled up through the catalyst/liquid mixture. Steam is generated within cooling coils immersed in the slurry-bed to remove the heat of reaction. This system has a high heat transfer rate resulting in a cheaper reactor with a higher productivity rate than catalyst particles packed in tubes. The lighter hydrocarbon products and unconverted syngas are withdrawn as vapor from the top of the reactor. Slurry is withdrawn from the reactor and pumped through a hydroclone and filter system which separates the clarified liquid products from the catalyst. The concentrated catalyst/slurry stream is returned to the reactor. A constant (steady-state) catalyst activity is maintained by continually withdrawing a small portion of catalyst from the reactor and replacing it with fresh catalyst.

Iron-based and promoted cobalt-based catalysts are the two primary catalysts currently used for F-T synthesis. Iron-based catalysts promote the water gas shift reaction which produces hydrogen from carbon monoxide and water; whereas cobalt catalysts generally do not. Therefore, for a syngas with a low hydrogen to carbon molar ratio, an iron based catalyst is preferred because it will produce hydrogen within the slurry-bed F-T synthesis reactor; whereas with a cobalt based catalyst, additional hydrogen has to be produced externally to the F-T synthesis reactor.

In the early 1990s, Bechtel developed several designs for indirect coal liquefaction plants using Fischer-Tropsch technology (references 3 and 4). Table II.2 shows the major input and output streams for the Baseline plant. The plant consumes 20,323 tpd of ROM Illinois No. 6 coal (8.6

wt% water) and 3,119 bpsd of normal butane to produce at total of 50,491 bpsd of petroleum products (1,921 bpsd of C3 LPG, 23,915 bpsd of gasoline, and 24,655 bpsd of distillate fuels). The plant is divided into the following three processing areas as shown in Figures 2.3 and 2.4.

Area 100	Clean Syngas Production Area
Area 200	Fischer-Tropsch Synthesis Loop
Area 300	Product Upgrading and Refining Area

The Area 100 Clean Syngas Production Area grinds and dries the coal, gasifies the coal in six Shell gasifiers (five operating and one spare), scrubs and cleans the syngas, and recovers 46.69 Mlb/hr of sulfur for sale. Plant 104, the COS Hydrolysis and Low Temperature Gas Cooling Plant, catalytically hydrolyzes the carbonyl sulfide (COS) and other trace sulfur compounds to H_2S, most of which is adsorbed in the downstream Acid Gas Removal Plant. The hydrolysis reaction is promoted by United Catalysts' C-53-2-01 catalyst. Plant 106, the Acid Gas Removal Plant, is a UPO Amine Guard SF unit which reduces the total sulfur content (H_2S plus COS) of the syngas to 5 ppm. Plant 108, the Sulfur Polishing Plant, reduces the sulfur content of the syngas to <0.1 ppm by passing it over solid zinc oxide (ZnO) at 650ºF to form solid zinc sulfide (ZnS). The G-72D ZnO catalyst from United Catalysts is used to adsorb the sulfur. The ZnS is discarded.

The Area 200 Fischer-Tropsch Synthesis Loop is designed to obtain a high CO conversion by recycling unconverted syngas. Therefore, in addition to the F-T hydrocarbon synthesis plant, this area contains a CO_2 removal plant, dehydration and compression facilities, a hydrocarbon recovery area, hydrogen recovery facilities, and an autothermal reforming unit.

The F-T hydrocarbon synthesis area contains 25 slurry-bed reactors (24 operating and one spare) arranged in eight parallel trains with each train having three reactors in parallel. The reactors have an inside diameter of 15.7 feet. The heat of reaction is removed by steam generation in bayonet tubes which are suspended within the reactor by a double tubesheet. This essentially is the maximum diameter reactor that can be built with this design because the tubesheet thickness will become excessive.[6] The design wax yield of 50 wt% at a 487.6ºF reactor temperature and 304 psia pressure was selected based on an economic analysis. The total fresh feed to the 24 slurry-bed reactors is 1,191.5 MMscfd. Unconverted syngas recycle and steam addition increase the total reactor feed to 1,537.4 MMscfd. This is 64.0 MMscfd (2.669 MMscf/hr) for each individual reactor.

Fischer-Tropsch catalyst-wax separation is performed using a 35,000 bpd Kerr-McGee ROSE-SR unit, the design of which was based on a DOE sponsored pilot plant test. The ROSE-SR unit design is based on an estimate of the characteristics of the catalyst system. Experimentation is needed to confirm this design.

Catalyst replacement was specified at 0.5 wt% per day of total catalyst inventory. A portion of the ROSE-SR unit underflow stream is split off for catalyst removal to counteract the fresh catalyst addition. The used catalyst is recovered by filtration, washed with naphtha, and dried in a Holo-Flite heated screw conveyor and solvent recovery system.

The Plant 204 Hydrocarbon Recovery and Separation Plant produces four F-T product streams that are sent to Area 300, the Product Upgrading and Refining Area, for upgrading into finished petroleum fuels, namely gasoline and diesel. The Hydrogen Recovery Plant recovers a high

[6] "Final Report – Slurry Vs. Fixed-bed Reactors for Fischer-Tropsch and Methanol", Slurry Reactor Design Studies, U. S. Department of Energy, Contract Number DE-AC22-89PC89867, June 1990.

purity hydrogen stream which is used in Area 300 for product upgrading. The Autothermal reformer 1.) increases the H2/CO ratio in the recycle gas, 2.) minimizes the build up of light ends in the Area 200 recycle loop, 3.) provides operating flexibility in case of emergencies, and 4.) converts excess fuel gas to F-T reactor feed.

The Area 300 Product Upgrading and Refining Area, shown in Figure 2.4, essentially is a small refinery. It contains a saturated gas plant, C3/C4/C5 alkylation unit, C4 isomerization unit, C5/C6 isomerization unit, catalytic reformer, naphtha hydrotreater, distillate hydrotreater, and a wax hydrocracker. To increase the gasoline yield, normal butanes are purchased, isomerized to isobutene in the C4 isomerization unit, and sent to the alkylation plant where they are reacted with the C3, C4 and C5 olefins to make alkylate, a high-octane gasoline blending component. The F-T naphtha goes to a naphtha hydrotreater which saturates the olefins and separates the low octane pentanes and hexanes from the C7+ naphtha. These pentanes and hexanes are isomerized to improve their octane ratings. The C7+ naphtha is sent to a catalytic reformer where it isomerized, dehydrogenated and cyclized to reformate, a high-octane gasoline blending component. The byproduct hydrogen is consumed in both isomerization units, both hydrotreaters, and the wax hydrocracker. The F-T distillate is hydrotreated in the distillate hydrotreater to saturate the olefins and sent to the diesel pool. The F-T wax goes to the wax hydrocracker where it is cracked to distillate and lighter components which undergo further processing in the previously described units. The distillate is sent to the diesel pool. Except for a small amount of aromatic compounds produced during hydrocracking, the diesel pool is primarily straight chain paraffins which are excellent diesel fuel blending components. The light gases that were produced during the previously described processing steps go to the saturated gas plant, from which the C4s are sent to the isomerization unit, the C3s are sold either as LPG or as a petrochemical feedstock, and the C2 and lighter components become fuel gas.

Table II.2

Design Input and Output Streams for the Baseline Indirect Coal Liquefaction Plant

Plant Inputs				
Illinois No. 6 ROM Coal*	1,693.6	Mlb/hr	20,323	TPD
Electric Power	54.36	MW		
Normal Butane	26.50	Mlb/hr	3,119	bpsd
Raw Water	10,042	gpm		
Plant Outputs				
C3 LPG	14.22	Mlb/hr	1,921	bpsd
Gasoline	251.44	Mlb/hr	23,915	bpsd
Diesel	278.21	Mlb/hr	24,655	bpsd
Sulfur	46.69	Mlb/hr		
Slag	187.03	Mlb/hr		

* As received coal containing 8.6 wt% water

Figure 2.3

Block Flow Diagram of Areas 100 and 200 of the Baseline Indirect Coal Liquefaction Plant

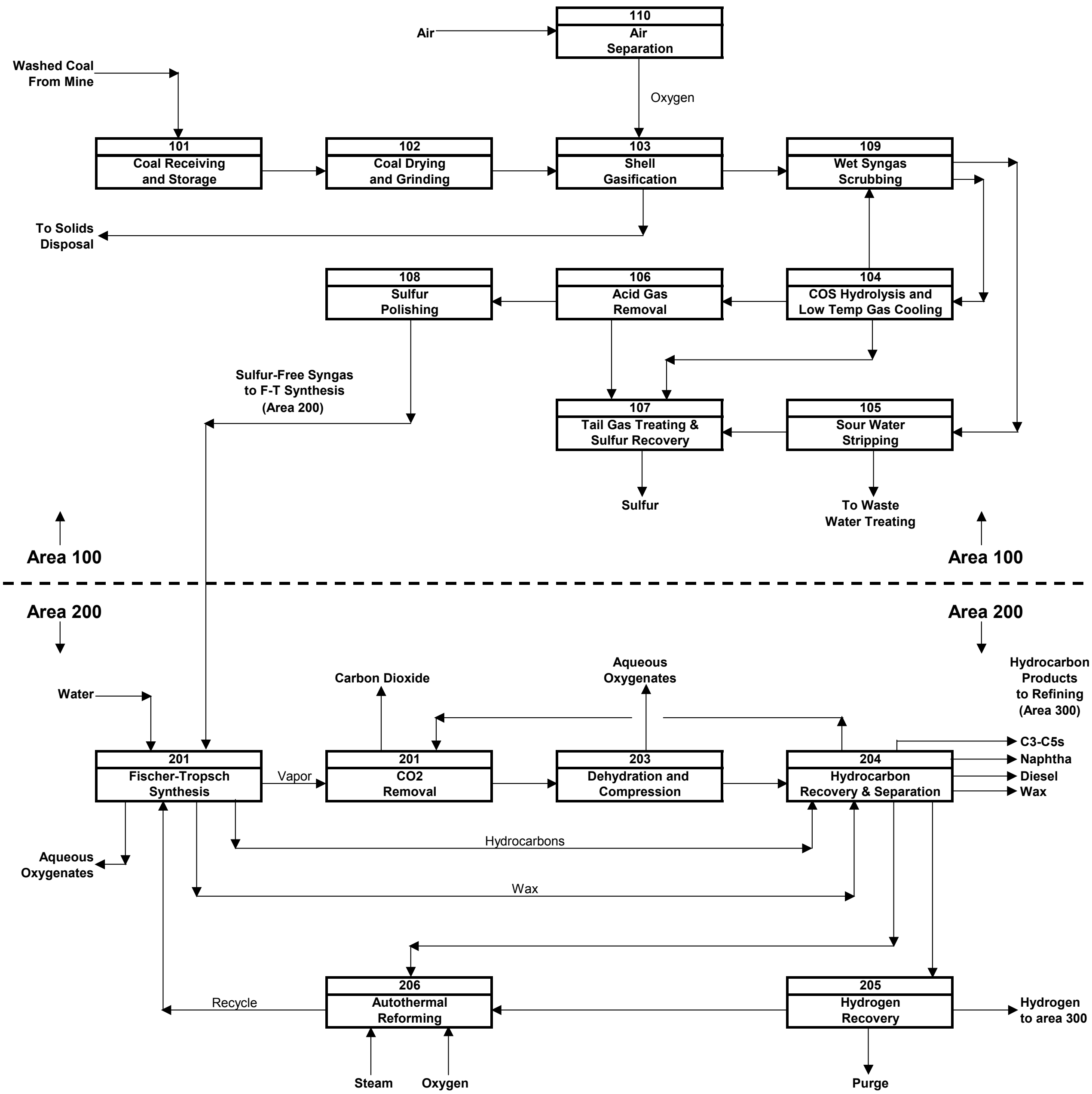
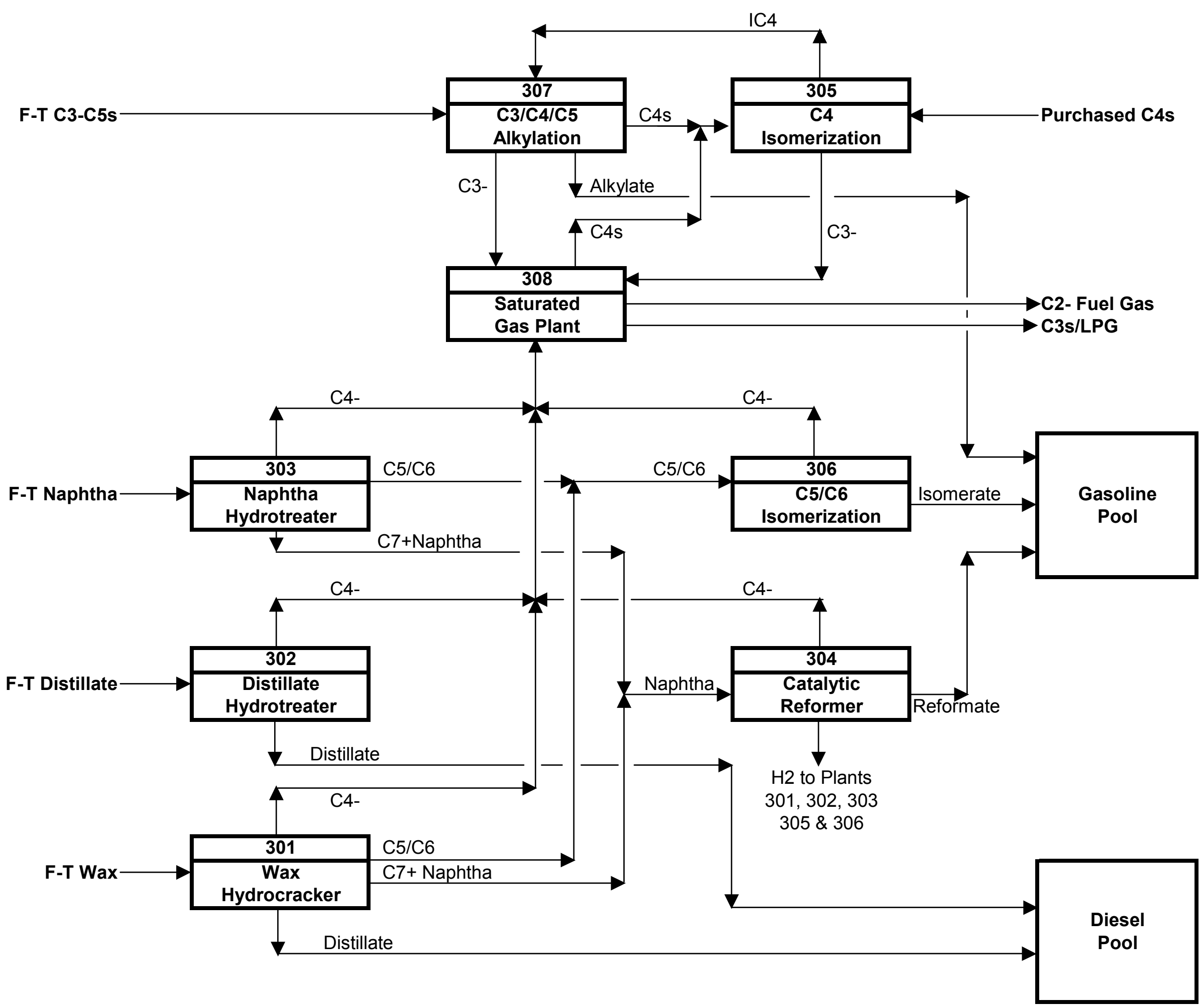


Figure 2.4

Block Flow Diagram of Area 300
of the Baseline Indirect Coal Liquefaction Plant



Section 3

Description of the Subtask 2.1 Coke Gasification Power Plant With Liquids Coproduction

The Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction was developed from the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant which produces 80 MMscfd of hydrogen and 980,000 lb/hr of 750ºF/700 psig steam for the adjacent petroleum refinery. Figure 3.1 shows the train configuration of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction.

Starting from the Subtask 1.3 Next Plant, the Subtask 2.1 plant was developed by eliminating the export steam production and replacing the hydrogen production facilities with a single-train, once-through Fischer-Tropsch hydrocarbon synthesis plant. A once-through system eliminates the cost of the expensive recycle system which includes recycle gas purification facilities in addition to the recycle compressor. The energy that was used to produce the export steam now is used to generate additional power. Even with almost the same coke feed rate to the gasifiers, the process changes required adjustments to the steam and water flows both in and between the gasification block and the power generation block in order to effectively balance the systems.

Table III.1 compares the design input and output stream flows for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. The Subtask 2.1 plant processes slightly less petroleum coke (5,376 vs. 5,417 dry tpd) than the Subtask 1.3 Next Plant. It also has a higher fresh water consumption of 6,472 gpm vs. 5,223 gpm. Furthermore, it consumes a small amount of natural gas, 23.2 MMBtu HHV/hr. Because the Subtask 2.1 plant does not export any hydrogen or steam, it produces more export power than the previous case (617 MW vs. 474 MW) in addition to 4,125 bpd of liquid fuel precursors from the F-T area.

On a lower heating value (LHV) basis, the plant has a thermal efficiency 47.8% when the heating value of the byproduct sulfur is included and 45.9% when the byproduct sulfur is not included. On a higher heating value (HHV) basis, the plant has a thermal efficiency 47.9% when the heating value of the byproduct sulfur is included and 46.0% when the byproduct sulfur is not included. These thermal efficiencies are higher than those that would be obtained from a coke IGCC power plant of a similar design because it includes the heating value of the liquid fuel that is produced. Since the second law of thermodynamics states this liquid fuel cannot be used at a 100% thermal efficiency, the thermal efficiency of the plant will be somewhat lower when the final disposition of the liquid fuel is considered.

Figure 3.2 is a simplified block flow diagram of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction. This plant can be considered to consist of three distinct main processing areas.

- The gasification island and air separation unit (Areas 100, 150, 250, 300, 350, 400, 420, and 800)
- The F-T hydrocarbon synthesis area (Areas 200 and 201)
- The power block (Areas 500 and 600)

Table III.1

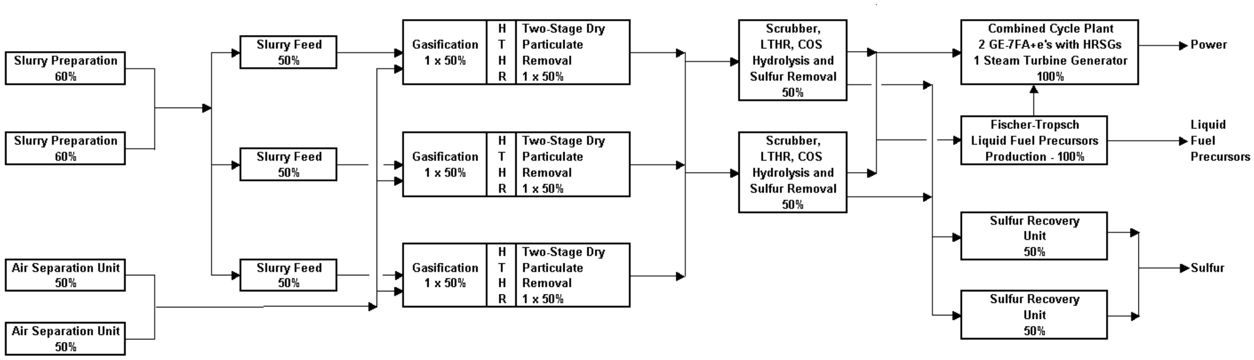
Design Input and Output Streams for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

	Subtask 1.3 Next Plant	Subtask 2.1 Power and Liquids Plant
Plant Inputs		
Coke Feed, as received TPD	5,692	5,649
Dry Coke Feed to Gasifiers, TPD	5,417	5,376
Oxygen Production, TPD of 95% O_2	5,954	5,919
Total Fresh Water Consumption, gpm	5,223	6,472
Condensate Return from the Refinery, lb/hr	686,000	0
Flux, TPD	110.6	109.7
Natural Gas, MMBtu HHV/hr	0	23.2
Plant Outputs		
Net Power Output, MW	474.0	617.0
Sulfur, TPD	373.4	370.6
Slag, TPD (15% moisture)	195.1	193.6
Hydrogen, MMscfd	80	0
HP Steam, 750°F/700 psia	980,000	0
Liquid Fuel Precursors, bpd	0	4,125

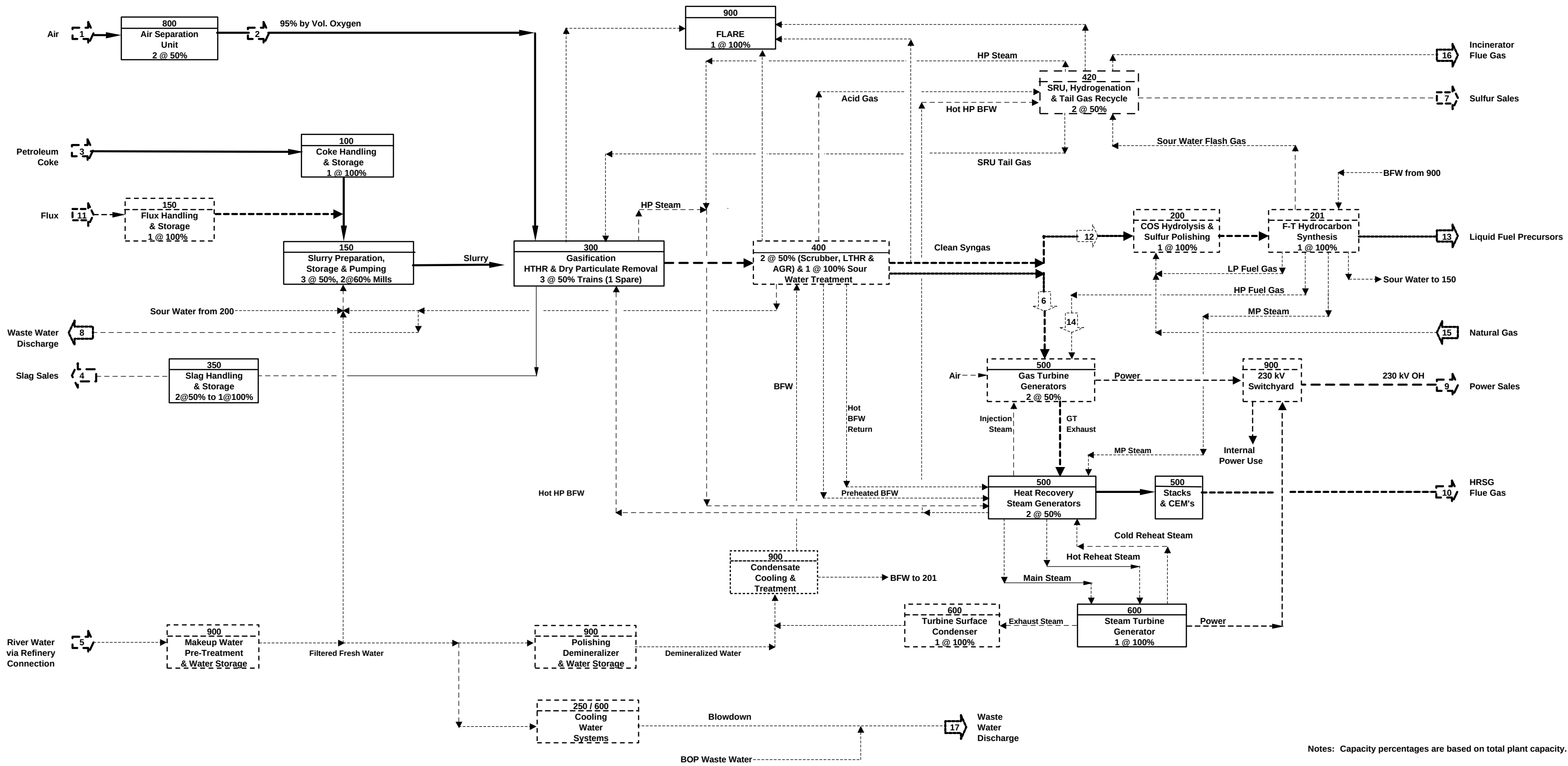
Figure 3.1

Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction

Simplified Block Train Diagram



Notes: Capacity percentages are based on total plant capacity.



Notes: Capacity percentages are based on total plant capacity.

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21
Flow	Air 25,808 Tons/Day	Oxygen 5,919 Tons/Day	Coke 5,376 Tons/Day	Slag 193.6 Tons/Day	Water 3,236,000 Lb/Hr	Syngas 722,540 Lb/Hr	Sulfur 370.6 Tons/Day	Water 52,000 Lb/Hr	Power 617,000 kWe	Flue Gas 7,966,800 Lb/Hr	Flux 109.7 Tons/Day	Syngas 403,502 Lb/Hr	Liq Fuel 48,897 Lb/hr	Fuel Gas 370,255 Lb/hr	Nat Gas 23.2 MMBtu/hr	Flue Gas 21,672 Lb/Hr	Water 711,000 Lb/Hr				
Nominal Pressure - psig	Atmos.	609	NA	NA	50	355	NA	62	NA	Atmos.	NA	360	50	365	50	Atmos.	Atmos.				
Temperature - F	70	240	Ambient	180	70	532	332	80	NA	265	NA	100	110	168	100	500	71				
HHV Btu/lb	NA	NA	14,848	NA	NA	4,325	3,983	NA	NA	NA	NA	4,997	19,777	1,852	1,000	NA	NA				
LHV Btu/lb	NA	NA	14,548	NA	NA	4,101	3,983	NA	NA	NA	NA	4,738	18,297	1,698	910	NA	NA				
Energy - MM HHV/hr	NA	NA	6,652	NA	NA	3,125	123	NA	NA	NA	NA	2,016	967.0	685.6	23.2	NA	NA				
Energy - MM LHV/hr	NA	NA	6,518	NA	NA	2,963	123	NA	NA	NA	NA	1,912	894.7	628.8	21.1	NA	NA				
Notes	Dry Basis	5,582 O2	Dry Basis	15%Wtr.	6,472 GPM	To GT	Sales	104 GPM	230 kV			For F-T	4,125 bpd				1,422 GPM				

DOE Gasification Plant Cost and Performance Optimization

Figure 3.2

Subtask 2.1

COKE GASIFICATION POWER PLANT

WITH LIQUIDS COPRODUCTION

BLOCK FLOW DIAGRAM

File: Fig 3.1.xls July 24, 2003

In addition there is a balance of plant area (Area 900). The remainder of this section describes the three main processing areas, the balance of plant area, and discusses the plant EPC cost.

3.1 Air Separation Unit and Gasification Island

The gasification island and the air separation unit basically are the same as those in the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant, as described in Appendix D of reference 2. Therefore, detailed descriptions of these areas will not be repeated here. The Gasification Island deals with the coke handling, gasification, syngas cooling and cleanup, sulfur production, and slag handling. These are Areas 100, 150, 250, 300, 350, 400, and 420. The Air Separation Unit has been renumbered to Area 800 for this case in order to allow the F-T hydrocarbon synthesis area to be Areas 200 and 201, which are more consistent with the nomenclature used in the indirect baseline study.

The fuel handling system (Area 100) provides the means to receive, unload, store, and convey the delayed petroleum coke to the storage facility. The coke and flux are mixed by the weigh belt feeders and transferred by coke feed conveyors to the day storage bins above the rod mills in the slurry preparation area.

The slurry preparation area also contains the flux receiving and storage facilities as well as the rod mills for grinding the coke. In order to produce the desired slurry solids concentration, coke is fed to each rod mill with water that is recycled from other areas of the gasification plant. Prepared slurry is stored in agitated tanks before being introduced into the first-stage of the gasifier.

The gasification, high temperature heat recovery, and particulate removal system (Area 300) is the heart of the Gasification Island. Global Energy's E-GAS™ gasification process consists of two stages, a slagging first-stage and an entrained flow non-slagging second-stage. The slagging section, or first-stage, is a horizontal refractory lined vessel into which the coke and flux slurry, recycle solids, and oxygen are atomized via opposing mixer nozzles. The coke and flux slurry, recycle solids, and oxygen are fed sub-stoichiometrically to the gasifier vessel at an elevated temperature and pressure to produce a high temperature syngas. The oxygen feed rate to the mixers is carefully controlled to maintain the gasification temperature above the ash fusion point; thereby ensuring good slag removal while producing high quality syngas.

The raw synthesis gas generated in the first-stage flows up from the horizontal section into the second-stage of the gasifier. The non-slagging second-stage is a vertical refractory-lined vessel into which additional coke slurry is injected via an atomizing nozzle to mix with the hot syngas stream exiting the first-stage. No oxygen is introduced into the second-stage. This additional slurry lowers the temperature of the gas exiting the first-stage by vaporizing the water in the slurry feed and by the endothermic nature of the steam and CO_2 reactions with carbon, thereby generating syngas and increasing cold gas efficiency.

The coke is almost totally gasified to form a synthetic fuel gas consisting primarily of hydrogen, carbon monoxide, carbon dioxide, and water. Sulfur in the coke is primarily converted to hydrogen sulfide (H_2S) with a small portion converted to carbonyl sulfide (COS); both of which are easily removed by downstream processing.

The gas and entrained particulate matter exiting the gasifier is further cooled in a firetube heat recovery boiler system to produce saturated steam at 1,650 psia which is superheated in the HRSG and used for power generation. The syngas leaving the high temperature heat recovery unit passes through a two-step cyclone/dry char filter particulate removal system to remove solids from the syngas. The recovered particulates are recycled to the gasifier. Water-soluble impurities are removed from the syngas in a wet scrubbing column following the dry char filters.

Mineral matter in the coke and flux form a molten slag which flows continuously through the tap hole into a water quench bath located below the first-stage. The slag then is crushed and removed through a continuous pressure let-down system as a slag/water slurry. This continuous slag removal technique eliminates high-maintenance, problem-prone lock hoppers and completely prevents the escape of raw gasification products to the atmosphere during slag removal.

The Area 350 slag handling and storage system processes and stores the slag. The slag slurry leaving the slag crushers at the outlet of the quench section of the gasifier flows continuously through the pressure let down system and into a dewatering bin. After passing through a settling tank to remove fine particles, the clear water is cooled in heat exchangers before it is returned to the gasifier quench section. The dewatered slag is loaded into trucks or rail cars for transport to market or to storage. The fines from the bottom of the settling tank are recycled to the slurry preparation area.

Area 400 contains the COS hydrolysis unit, low temperature heat recovery system, sour water treatment system, and the acid gas removal system.

Since COS is not removed efficiently by the downstream Acid Gas Removal (AGR) system, the COS must be converted to H_2S in order to obtain the desired high sulfur removal level. This is accomplished by the catalytic reaction of the COS with water vapor in the COS hydrolysis unit to create H_2S and CO_2. The H_2S is removed in the downstream AGR section and the CO_2 remains in the syngas.

Upon exiting the COS hydrolysis unit, the syngas is cooled in a series of shell and tube exchangers which condense water, ammonia, some carbon dioxide, and hydrogen sulfide in an aqueous solution. This water goes to the sour water treatment unit. Some of the cooled syngas goes to the syngas recycle compressor for use in various areas of the plant; such as for quench gas in the second-stage of the gasifier and for back pulsing the dry char filters.

The heat removed prior to the AGR unit provides moisturizing heat for the product syngas, steam for the AGR stripper, and condensate heat. Cooling water provides trim cooling to ensure that the syngas enters the AGR at a sufficiently low temperature.

The sour water treatment system removes the small amounts of dissolved gases (i.e., carbon dioxide, hydrogen sulfide, ammonia, and other trace contaminants) from the condensed water and any other process water. The gases are stripped out of the sour water in a two-step process. First, the acid gases are removed in the acid gas stripper column by steam stripping. The stripped gases go to the Sulfur Recovery Unit (SRU). The water from the acid gas stripper column, is cooled, and a major portion is recycled to slurry preparation. The remainder is treated in the ammonia stripper column to remove the ammonia, filtered to remove trace organics and solids, and then sent to the waste water management system. The stripped ammonia is combined with water that will be recycled back to the slurry mix tank after being cooled with cooling tower water.

The acid gas removal (AGR) system removes the H_2S from the syngas to produce a sweet syngas. The H_2S is removed from the sour syngas in an absorber column at high pressure and low temperature using a solvent, methydiethanolamine (MDEA). After the hydrogen sulfide removal, the syngas going to the gas turbine is heated and moisturized. Non-moisturized syngas is sent to Area 200 for sulfur polishing before F-T synthesis.

The H_2S rich MDEA solution leaving the absorber goes to a stripper column where the H_2S is removed by steam stripping at a lower pressure. The concentrated H_2S exits the top of the stripper column and goes to the Sulfur Recovery Unit. The lean amine exits the bottom of the stripper, is cooled, and then recycled to the absorber. An online MDEA reclaim unit continuously removes impurities from the lean amine to improve system efficiency.

The Area 420 sulfur recovery unit (SRU) processes the concentrated H_2S from the AGR unit and the CO_2 and H_2S stripped from the sour water in a reaction furnace, a waste heat recovery boiler, and then in a series of Claus catalytic reaction stages where the H_2S is converted to elemental sulfur. The sulfur is recovered as a molten liquid and sold as a by-product.

The tail gas stream, composed of mostly carbon dioxide and nitrogen with trace amounts of sulfur dioxide, exits the last catalytic stage and goes to tail gas recycling. It is hydrogenated to convert all the remaining sulfur species to H_2S, cooled to condense the bulk of the water, compressed, and then injected into the gasifier. This allows for a very high sulfur removal efficiency with low recycle rates.

Area 800 contains two 50% capacity Air Separation Units (ASUs) to deliver the required oxygen for the coke gasification process. Each ASU consists of several subsystems and major pieces of equipment, including an air compressor, air cooling system, air purification system, cold box, and product handling and backup systems.

Gaseous oxygen leaves the cold boxes at moderate pressure and is compressed in centrifugal compressors and delivered to the gasifiers. Nitrogen tanks with steam vaporizers provide gaseous nitrogen for various in-plant uses such as purging vessels.

The Area 250 cooling water system provides cooling water to the gasification island and ASU. A second system provides the cooling duty for the power block.

The major components of the cooling water system consist of a cooling tower, circulating water pumps, and appropriate piping for distribution of the cooling water around the facility. Both cooling towers are multi-cell mechanically induced draft towers, sized to provide the design heat rejection at the ambient conditions corresponding to the maximum summer temperature. Chemical treatment systems, including metering pumps, storage tanks and unloading facilities provide the necessary biocide, pH treatment and corrosion inhibiting chemicals for the circulating water system. Cooling tower blowdown discharges to the wastewater management system.

3.2 Fischer-Tropsch Hydrocarbon Synthesis Area

The design for the Fischer-Tropsch Hydrocarbon Synthesis Area was developed based on the ASPEN Plus process flowsheet reactor model that was developed for the Baseline Design/Economics for Advanced Fischer-Tropsch Technology study.[4] The ASPEN Plus process flowsheet model of the Fischer-Tropsch (F-T) hydrocarbon synthesis area that was developed for this study does not consider the following systems:

- Filter system (and associated hydrocarbon circulation loop) which removes the catalyst from the liquid product leaving the slurry-bed F-T reactor
- Used catalyst removal and disposal system
- Fresh catalyst handling and pretreatment systems

The designs for these systems were developed based on the previous study.

The Fischer-Tropsch hydrocarbon synthesis area consists of two sub areas, Area 200 and Area 201. Area 200 is the Final Syngas Cleanup Area, which removes the final traces of sulfur from the syngas, before it is converted to hydrocarbons in Area 201, the Hydrocarbon Synthesis and Product Recovery Area.

3.2.1 Final Syngas Cleanup Area

The Final Syngas Cleanup Area, Area 200, reduces the sulfur concentration of the cleaned syngas from the acid gas removal area of the gasification block to less than 0.1 ppm of sulfur. This is accomplished by hydrolyzing the small amounts of carbonyl sulfide (COS) and trace amounts of other light organic sulfur compounds (such as CS_2) to hydrogen sulfide (H_2S), and removing the H_2S by reacting it with zinc oxide (ZnO) to produce solid zinc sulfide (ZnS) and water. The ZnO is permanently consumed, and the ZnS/ZnO mixture eventually is discarded.

Süd-Chemie's G-41P RS hydrolysis catalyst is used at a 300ºF operating temperature to hydrolyze the COS to H_2S and H_2O. This is a potassium chromate on aluminum oxide catalyst and is provided in 1/8 inch extrudates. At the design volumetric hourly space velocity of 3,000 vol/vol-hr, the expected catalyst life is greater than 60 months.

Süd-Chemie's G-72E ZnO catalyst/sulfur adsorbent is used to capture the sulfur and reduce the residual syngas sulfur content at 650ºF. In order to provide continuous H_2S removal, the process design uses a two bed reactor configuration with the two beds in series. Necessary piping is provided so that these two beds can be switched, and the spent adsorbent can be replaced without any interruption of service. When H_2S breakthrough occurs in the first bed (lead bed), it is taken out of service for adsorbent replacement, and the other bed (lag bed) is in service alone. After the adsorbent has been replaced, the bed with the freshly loaded adsorbent is put back in service as the lag bed. The two bed in series operation continues until H_2S breakthrough occurs in the other bed, and it is taken out of service for adsorbent replacement. The operating cycle repeats. Each catalyst bed is sized for a six month cycle length.

Figure 3.3 contains a schematic flow diagram of the Final Syngas Cleanup Area, Area 200, and the F-T slurry-bed reactor and product recovery area, Area 201.

The cleaned syngas from the gasification block is preheated to 292ºF in heat exchanger 201E-1 with hotter sulfur-free syngas from exchanger 200E-2. The preheated syngas leaving the 200E-1 heat exchanger is mixed with 440ºF/375 psia stream that was generated in the slurry-bed F-T reactor and fed to the 200R-1 COS Hydrolysis Reactor where the following chemical reaction converts the COS to H_2S.

$$COS + H_2O \rightarrow H_2S + CO_2$$

The syngas leaving the 200R-1 reactor is heated to 520ºF in exchanger 200E-2 with the hot sulfur-free syngas leaving the 200R-2 and/or 200R-3 ZnO sulfur adsorbers. The syngas then is heated to 650ºF in the 200F-1 furnace. The furnace fuel is a mixture of natural gas and the low-pressure fuel gas recovered from the F-T reaction products. This furnace is oversized for startup.

The hot syngas then enters the ZnO sulfur adsorption beds, 200R-2 and 200R-3. Although it is not shown in the drawing, these two beds are arranged in a lead-lag configuration so that one bed may be taken off line for ZnO replacement while the other remains in service.

The sulfur-free syngas leaving the ZnO beds is cooled to 257ºF by preheating the entering cleaned syngas in the 200E-1 and 200E-2 heat exchangers. This sulfur-free syngas stream is the feed stream to the 201R-1 slurry-bed F-T hydrocarbon synthesis reactor.

3.2.2 Fischer-Tropsch Slurry-bed Reactor Area

The Fischer-Tropsch slurry-bed reactor converts the sulfur-free syngas primarily into olefinic hydrocarbons by the reaction

$$n\,CO + 2n\,H_2 \rightarrow C_nH_{2n} + n\,H_2O$$

The reaction is promoted by an iron-based catalyst which also promotes the water-gas shift reaction

$$CO + H_2O \rightarrow CO_2 + H_2$$

The lighter hydrocarbon products leave the slurry-bed reactor in the vapor phase, are cooled and the condensed liquid collected. The heavier hydrocarbons are removed as liquids from the reactor, separated from the suspended catalyst, cooled, and combined with the lighter products to form the liquid fuel precursors product.

In order to maintain a constant catalyst activity, there is a continual addition of fresh catalyst and a continual withdrawal of used catalyst from the slurry-bed reactor. The fresh catalyst must be pretreated in a reducing atmosphere at an elevated temperature to activate it. The catalyst pretreating system consists of a similar vessel to the slurry-bed reactor, but without the internal cooling facilities.

As shown in Figure 3.3, the cooled sulfur-free syngas stream from the zinc oxide sulfur adsorption beds is mixed with 440ºF/375 psia steam before entering the slurry-bed F-T hydrocarbon synthesis reactor, 201R-1, where the hydrogen and carbon monoxide are converted to straight chain aliphatic hydrocarbons, carbon dioxide and water. The heat of reaction is removed from the slurry-bed F-T reactor by the generation of 440°F/375 psia steam inside tubes located within the slurry-bed reactor. Pump 201P-2 circulates boiler feed water (BFW) between the 201C-1 steam drum and the 201R-1 reactor to ensure that sufficient BFW always is flowing through the cooling tubes.

Cyclone 201T-1 removes entrained catalyst particles from the vapor stream leaving the top of the F-T reactor. The vapor stream then is cooled to 110°F in two exchangers, 201E-1 and 201E-2. The first exchanger (201E-1) cools the syngas to 130ºF by heating BFW, and the second exchanger (201E-2) cools the syngas to 110ºF with cooling water. The cooled syngas leaving the second exchanger enters the 201C-2 reactor overhead flash drum. The sour water from the boot of 201C-2 goes to the 201C-4 sour water flash drum. The vapor stream leaving the sour water flash drum goes to the incinerator, and the sour water is recycled to the gasifier.

The vapor stream from the reactor overhead vapor flash drum is washed in 201C-3 to remove any residual catalyst particles and heated from 110°F to 120°F in exchanger 201E-3 to prevent condensation of the heavy components during compression. Condensing 440°F/375 psia steam, that was generated in the slurry-bed F-T reactor, is the heating medium. The heated vapor stream is compressed to 380 psia in 201K-1 to produce a high-pressure fuel gas stream which is sent to the power block where it is mixed with the syngas and steam before entering the combustion turbine. This high-pressure fuel gas stream consists of unconverted syngas (carbon monoxide and hydrogen) and light hydrocarbons (primarily C1 through C3s) that are produced in the F-T reactor.

The liquid hydrocarbon stream leaving 201C-2 is mixed with the cooled liquid hydrocarbons from the slurry-bed F-T reactor and sent for upgrading into liquid transportation fuels.

The liquid stream leaving the slurry-bed F-T reactor passes through hydroclone 201T-2 to remove a majority of the entrained catalyst particles. The catalyst-rich hydroclone bottoms goes to mixing tank 201C-10 from which most of it is returned to the slurry-bed reactor by pump 201P-3. A portion of the hydroclone bottoms is withdrawn and sent to the catalyst withdrawal system shown in Figure 3.4. Residual catalyst particles are removed from the hydroclone overhead stream in the 201T-3 filter system.

The catalyst-free liquid leaving the filter system is reduced in pressure and flashed in drum 201C-5. The vapor stream is further cooled to 110°F in exchanger 201E-4 with cooling water and flashed in drum 201C-6. The vapor stream from drum 201C-6 is a low-pressure fuel gas which is used as fuel in the 200F-1 furnace.

The liquid leaving the 201C-5 flash drum is cooled to 200ºF in 201E-5 by preheating boiler feed water. The cooled liquid from 201C-5 is mixed with the liquid stream from the 201C-6 flash drum in drum 201C-9 and a cooled liquid recycle stream from 201C-8. This mixture now is cooled to 110ºF by cooling water in exchanger 201E-6 and sent to the 201C-8 liquid fuel flash drum along with the liquid from the 201C-2 reactor overhead vapor flash drum. The vapor leaving the 201C-8 liquid fuel flash drum is mixed with the vapor from the 201C-6 flash drum and is used as low-pressure fuel gas in the 200F-1 furnace.

The liquid from the 201C-8 flash drum is split into two streams. One of the liquid streams is recycled back to 201C-9 flash drum via pump 201P-1 to dilute the heavier hydrocarbons in order to control their viscosity as they are cooled in exchanger 201E-6. The other liquid stream is the liquid fuel precursors product which is sent to the adjacent petroleum refinery for upgrading into liquid transportation fuels (gasoline, diesel, etc.).

Figure 3.4 shows the catalyst withdrawal system. The hot catalyst-rich stream from the 201C-10 drum is cooled in exchanger 201E-7 and pumped by pump 201P-4 through the 201T-4 filters to remove the used catalyst which is collected and discarded. The catalyst free liquid is mixed with the liquid fuel precursors product stream from drum 201C-6 and sent to the adjacent petroleum refinery for upgrading.

The catalyst pretreatment system also is shown in Figure 3.4. The makeup catalyst is fed into the 201C-11 catalyst pretreater where it is combined with heated liquid product from storage. Recycle gas is circulated through the pretreater vessel via compressor 201K-2, exchanger 201E-9, and furnace 201F-1. Vapors leaving the pretreater vessel are cooled in exchangers 201E-9 and 201E-10 before being flashed in drum 201C-13. A portion of the vapor from 201C-13 is withdrawn and sent to the incinerator to remove inerts from the system. However, most of the vapors from 201C-13 are recycled to the pretreater after addition of some fresh syngas or hydrogen via the 201K-2 compressor, 201E-9 exchanger, and 201F-1 furnace. Pretreated catalyst is withdrawn from the pretreater vessel and stored in the heated 201C-12 mixing tank until it is injected into the slurry-bed F-T reactors via pump 201P-8.

3.3 Power Block

The major components of the power block include two gas turbine generators (GTG), two heat recovery steam generators (HRSG), one steam turbine generator (STG), and numerous supporting facilities.

Area 500 contains the gas turbines (GT), heat recovery steam generators (HRSG), and stacks. Each of the two combustion turbine generators are General Electric 7FA+e machines, each with a nominal output of 210 MW. Each GT utilizes moisturized syngas and steam injection for NO_x control. Combustion exhaust gases from each GT are routed to its associated HRSG and stack. Natural gas is used as back-up fuel for the gas turbine during startup, shutdown, and short duration transients in syngas supply. Optionally, each turbine can be fully fired on natural gas to generate power when syngas is unavailable.

The HRSG receives the GT exhaust gases and generates steam at the main steam and reheat steam energy levels. It generates high-pressure (HP) steam and provides condensate heating for both the combined cycle and the gasification facilities. The HRSG is a fully integrated system consisting of all required ductwork and boiler components. Each component is designed for pressurized operation.

The HRSG boiler includes a steam drum for proper steam purity and to reduce surge during cold start. Large unheated down comers assure proper circulation in each of the banks.

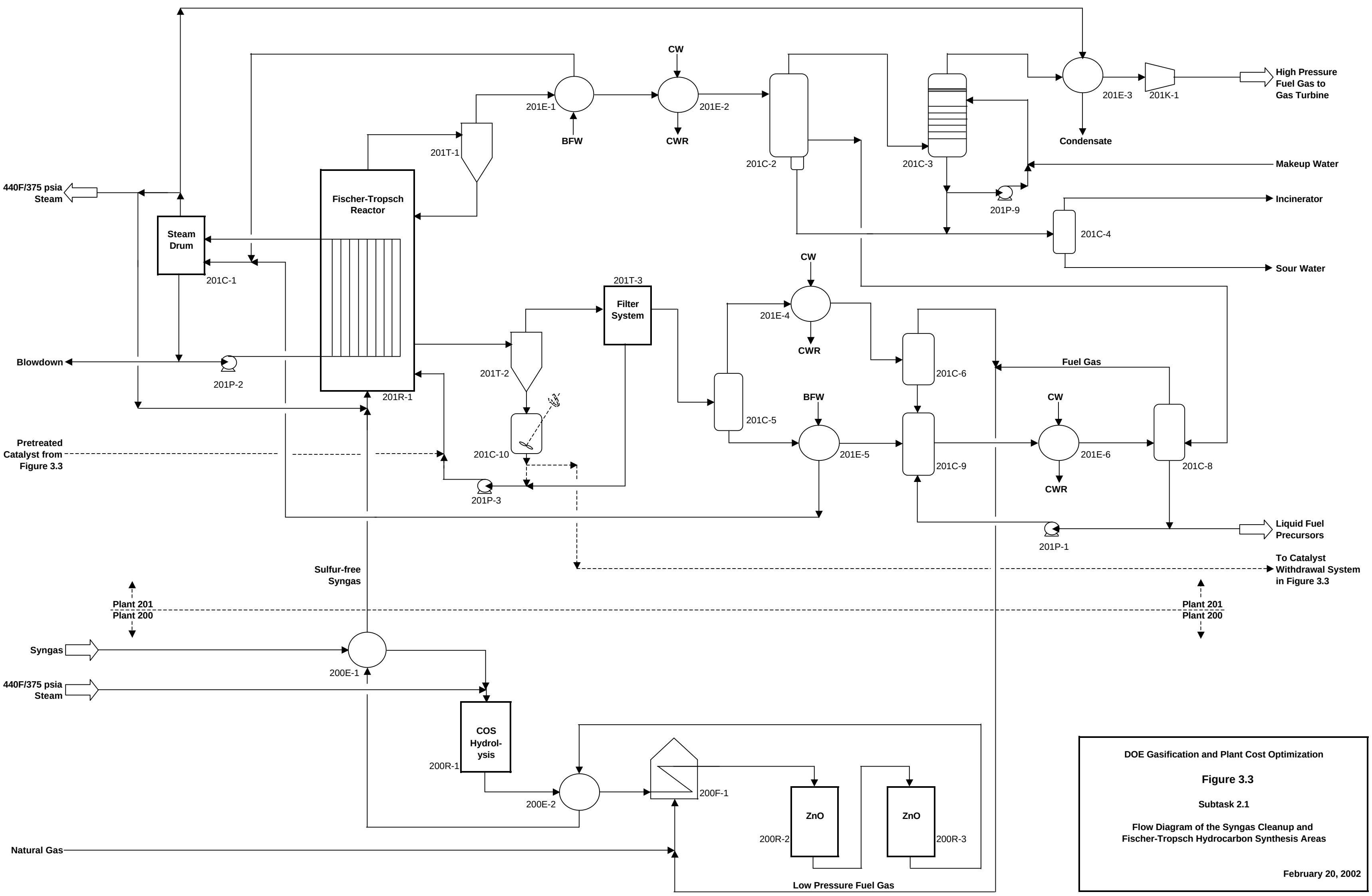
Each stack includes a continuous emission monitoring (CEM) system.

The Area 600 steam turbine (ST) is a reheat, condensing turbine that includes an integrated HP/IP opposed flow section and an axial flow LP section. Turbine exhaust steam is condensed in a surface condenser. The reheat design ensures high thermal efficiency and excellent reliability. The steam turbine produces 310 MW of electric power.

The power delivery system includes the GT generator output at 18 kilovolts (kV) with each connected through a generator breaker to its associated main power step-up transformer. A separate main step-up transformer and generator breaker is included for the ST generator. The HV switch yard receives the energy from the three generator step-up transformers at 230 kV.

Two auxiliary transformers are connected between the GTG breakers and the step-up transformers. Due to the large auxiliary load associated with the plant, internal power is distributed at 33 kV from the two auxiliary power transformers. The major motor loads in the ASUs are serviced by 33/13.8 kV transformers. Several substations serve the other internal loads with 33/4.16 kV transformers supplying a double-ended electrical bus.

Area 600 also includes a cooling water system similar to that in Area 250 and an emergency shutdown transformer to connect the 230 kV switch yard with essential safe shutdown loads.



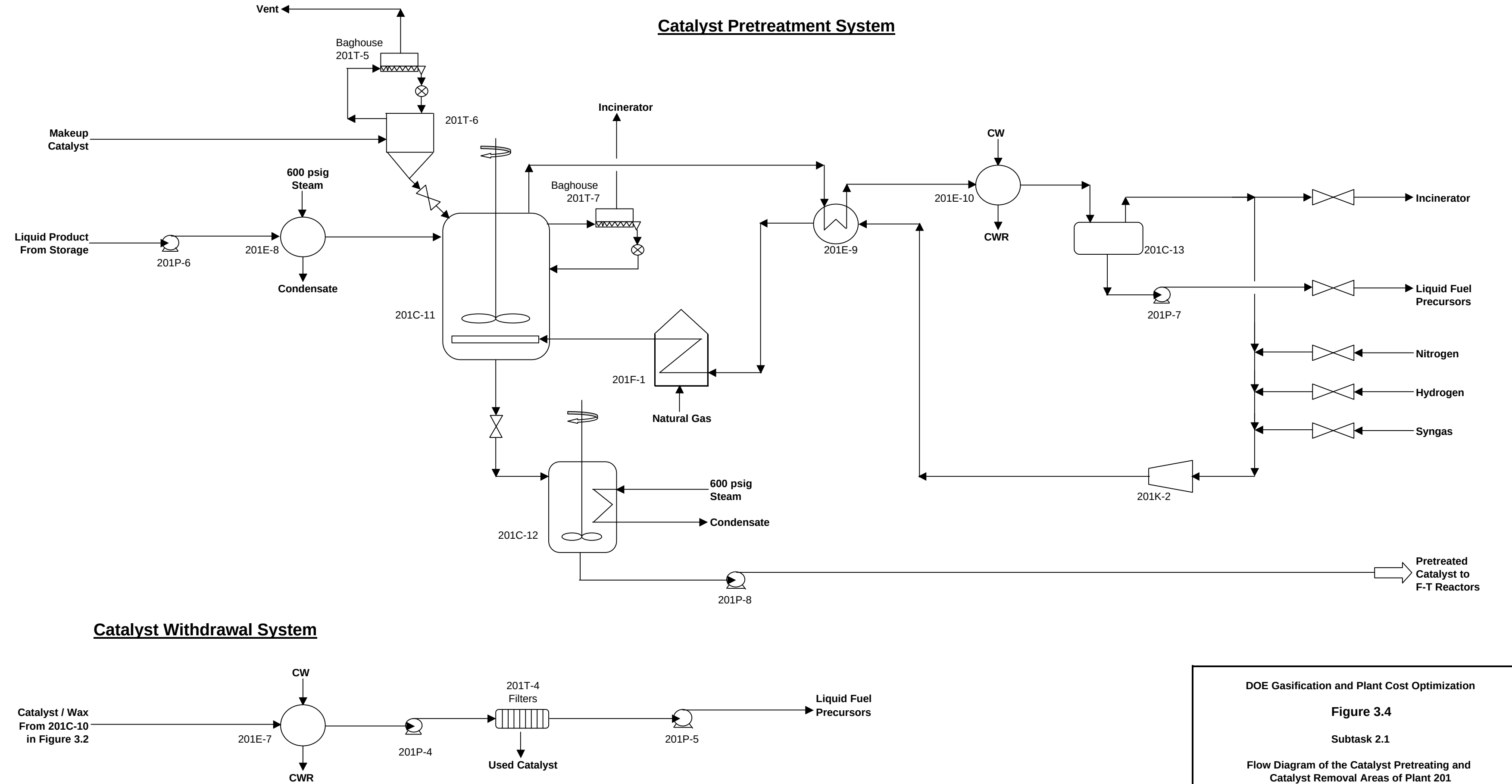
DOE Gasification and Plant Cost Optimization

Figure 3.3

Subtask 2.1

Flow Diagram of the Syngas Cleanup and
Fischer-Tropsch Hydrocarbon Synthesis Areas

February 20, 2002



DOE Gasification and Plant Cost Optimization

Figure 3.4

Subtask 2.1

Flow Diagram of the Catalyst Pretreating and Catalyst Removal Areas of Plant 201

December 4, 2002

3.4 Balance of Plant

The Area 900 balance of plant contains nine subsystems.

The fresh water supply system filters river water from an industrial water supply network for the fresh makeup water supply. A demineralizer supplies demineralized water for boiler water makeup. The demineralizer regeneration wastewater is sent to a process waste collection tank, where it is neutralized before discharge.

The fire and service water system includes a loop around the principal facilities with fire hydrants located for easy access. It also includes an onsite water storage tank. A jockey pump maintains line pressure in the loop during stand-by periods. During periods of high water usage, motor and diesel driven pumps are available.

The waste water management system processes both clear wastewater and storm water from a clean water collection pond. Clear wastewater includes water treatment effluent, cooling water blowdown, flushes and purges from equipment maintenance, filtered water from the ammonia stripper column (in Area 400), clarifier overflow, and sewage treatment overflow. Storm water is collected in a storm-water pond before going to the clean water collection pond. The water in the clean water collection pond is analyzed and treated, as required, until it meets permitted outfall specifications for discharge through the refinery waste water system.

The service and instrument air system provides compressed air and dried instrument air to users throughout the plant. The system consists of air compressors, air receivers, hose stations, and piping distribution for each unit.

The incineration system destroys the tank vent streams from various in-process storage tanks and drums that may contain small amounts of hydrocarbons and other gases such as ammonia and acid gas. During process upsets of SRU, tail gas streams also can be processed in the high temperature incinerator. The high temperature produced in the incinerator thermally destroys any residual hydrogen sulfide before the gas is vented to the atmosphere. The incinerator exhaust feeds into a heat recovery boiler to produce process steam.

The flare system provides for safe disposal of syngas during startup or short term upsets. The flare includes a natural gas fired pilot flame to ensure that the flare is continually operating.

The instrumentation and control system provides data acquisition, monitoring, alarming and control by a digital distributed control system (DCS). The DCS allows the plant to be operated from the central control room using the DCS as the control platforms. The gas and steam turbines, and the coke handling programmable logic controllers will continue to execute all permissive, protective, and sequence control related to their respective equipment. They will be controlled either locally using the turbine vendor man machine interface system, or from the DCS.

Other balance of plant equipment such as air compressors, condenser vacuum pumps, and water treatment facilities can be controlled by either local PLCs, or from contact and relay control cabinets. All remaining plant components are exclusively controlled by the DCS including the HRSG, the gasifier, ASU, hydrogen plant, electrical distribution, and other power block and gasification support systems.

The plant has a central building housing the main control room, office, training, other administration areas, and a warehouse/maintenance area. Other buildings are provided for water

treatment equipment and the motor control centers. The buildings, with the exception of water treatment, are heated and air-conditioned to provide a climate-controlled environment for personnel and electrical control equipment.

A series of strategically placed safety showers are located throughout the facility.

3.5 Plant Cost

Table III.2 shows the "overnight" EPC cost for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and compares it with that of the Subtask 1.3 Next Plant IGCC Coproduction Plant. These costs are on a mid-year 2000 basis; the same basis as those of Subtask 1.3 Next Plant and the other Task 1 plant costs.[7]

The Subtask 2.1 EPC cost was developed form the Subtask 1.3 Next Plant EPC cost by subtracting the cost of the hydrogen production and compression facilities, and then adding the cost of the F-T hydrocarbon synthesis area and the additional costs associated with the greater power production in the larger steam turbine. No adjustments were made to the costs of the solids handling, ASU, and gasification areas of the facility even though the amount of coke processed is 0.75% less than that of the reference case. This is a conservative assumption. Furthermore, no adjustments were made to the balance of plant area to account for the larger water treating facilities required by the larger amount of fresh water that has to be processed. Because of the way the additional cost of the power block was estimated, this cost and the cost associated with the bigger power distribution facilities are included in the additional cost of the power block.

The cost of the F-T area was estimated from the processing equipment sizes using an appropriate installation factor that was developed from previous cost estimates for similar facilities. The estimated cost of the large F-T slurry-bed hydrocarbon synthesis reactor is over 60% of the total equipment cost in the F-T area, and consequently, it dominates the cost of this area. Until wider experience is obtained with the construction of these large reactors, their estimated cost cannot have a high degree of accuracy.

The accuracy of the total installed cost for the Subtask 1.3 Next Optimized Coke IGCC Coproduction Plant was estimated to be on the order of ±11%. This level of accuracy reflects a high degree of confidence based on the large number of vendor quotes that were obtained and that the power block costs are based on a current similar Gulf Coast power project. This accuracy applies only to the total plant cost and does not apply to the individual areas or parts.

The accuracy of the total installed cost for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction is not as good. The estimated cost of the F-T area is only an order of magnitude cost estimate (nominally ±30%) because of the manner in which it was developed. Thus, the over estimate accuracy for the Subtask 2.1 plant probably is in the ±15% range.

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[7] All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

Table III.2

Capital Cost Summary for the
Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction
and the
Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

Plant Area	Subtask 2.1 Power and Liquids Plant	Subtask 1.3 Next Plant
Solids Handling	8,012,000	8,012,000
Air Separation Unit	107,246,000	107,246,000
Gasification	312,591,000	312,591,000
Hydrogen Production	0	42,931,000
F-T Liquids Area	34,279,000	0
Power Block	276,414,000	237,045,000
Balance Of Plant	79,420,000	79,420,000
Total	817,962,000	787,246,000

Notes:

1 Because of rounding, the columns may not add to the total that is shown.

2. All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

Section 4

Availability Analysis

In all the previous Task 1 cases, an availability analysis was used to determine the daily average production rates for calculating the annual production rates and cash flow. This analysis showed that the inclusion of a spare gasification train in the Subtask 1.3 Next Plant was a worthwhile addition that increased the return on investment although it increased the plant cost.

The common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period, all are dependent on the project cash flow. The net cash flow is the sum of all project revenues and expenses. Depending upon the detail of the financial analysis, the cash flow streams usually are computed on 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. Therefore, the annual production rate is a key parameter in determining the financial performance of a project.

4.1 Use of Natural Gas

The gasification trains in the Subtask 2.1 plant are sized so that when both trains are operating they will fully load the two combustion turbines and the F-T hydrocarbon synthesis reactor. However, when only one gasification train is operating, there is insufficient syngas available to operate the F-T area at full capacity and fully fire one combustion turbine. Thus, in this situation, supplemental natural gas is used to augment the syngas and co-fire one or both combustion turbines. When this situation occurs, the power output from the combustion turbines is reduced. However, the internal power consumption in the plant also is reduced when one gasification train is not operating by the internal power it consumes and the power consumed by one air separation unit. The net effect of this combination of events is that there is a net reduction in the export power.

Depending upon the situation, two, one or no gasification trains can be producing syngas at any time, and simultaneously, two, one or no combustion turbines may be available for operation. Thus, there are three operating philosophies that will be considered in the subsequent availability analysis.

1. Base Case This case operates under the philosophy that F-T liquids production is to be maintained whenever syngas is available, and supplemental natural gas is to be used to co-fire one or both combustion turbines at maximum rate when they are available.
2. Minimum Natural Gas Use Case This case operates under the philosophy that F-T liquids production is to be maintained whenever syngas is available, and supplemental natural gas is to be used only to co-fire one combustion turbine at maximum rate when either of the two combustion turbines are available.
3. Maximum Power Case This case maximizes power production. It operates under the philosophy that the F-T liquids are only to be produced when both gasifiers are producing syngas. When syngas is only available from one gasifier, the F-T liquids production will be terminated and supplemental natural gas will be used to co-fire both combustion turbines at maximum rate when they are available.

In all of the above cases when both gasifiers are available and only one turbine is available, F-T liquids will be produced and the gasifiers either will be operated at a reduced rate and/or the excess syngas will be flared. Also, in the very rare situation when both gasifiers are unavailable, natural gas will only be used to fire one combustion turbine to produce power.

The average daily production and natural gas rates for each of the above three scenarios are calculated as part of the availability analysis and are shown later in this section. Natural gas usage during startup and during maintenance operations, such as for curing refractory, are not considered in the availability analysis calculations, but are included in the operating and maintenance costs during the financial analysis.

4.2 Availability Analysis

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.[1] 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 three adjustments, this data was used to estimate the availability of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction under the previously described operating scenarios. The first adjustment increased the availability of the air separation plant from the observed availability of 96.32% to the industry average availability of 98%. The second adjusted the availability of the first gasification stage to remove a slag tap plugging problem caused by an unexpected change in the coal blend to the gasifier. This adjustment is justified since a dedicated petroleum coke plant would be very unlikely to experience this problem. The third eliminated a short outage that occurred in the water treatment facility because this plant will have sufficient treated water storage to handle this type of outage.

Using the EPRI recommended procedure, availability estimates were calculated for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction Plant for the three operating scenarios described previously.[8] Table IV.1 defines the plant configurations for both the Subtask 2.1 and Subtask 1.3 Next Plant. This table shows that the two plants have essentially the same configuration except for byproduct production. Both plants contain a spare gasification train. Because the Low Temperature Heat Recovery/Acid Gas Removal (LTHR/AGR) area and Sulfur Recovery Unit (SRU) are highly reliable, these facilities are not spared. However, they only contain sufficient capacity so that only two of the three parallel gasification trains can be operated simultaneously.

For each plant, the potential syngas availability from two of the three gasification trains at full design capacity is 86.5% of the time, and from only one of the two trains, it is 99.63%. The equivalent syngas availability is 93.24% of the design capacity.

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

Recent data presented at the 2002 Gasification Technologies Council conference by Clifton Keeler show further reliability improvements in the on-stream performance of the Wabash River Repowering Project.[9] However, the following availability and financial analyses will be based on the data reported in the final repowering project report for consistency with the Task 1 results. This will cause the following results to be somewhat conservative.

Table IV.2 shows the equipment status for of the Subtask 2.1 plant for each of the three operating scenarios; the Base Case, the Minimum Natural Gas Use Case, and the Maximum Power Case. Also shown in this table is the feed that each piece of equipment would be using in each situation when it is operating. In addition, below the case identification is shown the expected annual percentage of time the gasification and power blocks would be operating under these conditions. For example, in Case A under all scenarios, both gasifiers would be operating at design capacity on coke, both combustion turbines would be operating at design capacity on syngas and F-T fuel gas (unconverted syngas and light hydrocarbons), and the F-T liquefaction would be operating at design capacity on syngas.

Case C is a different situation that is expected to occur only about 4.44% of the time. In all scenarios, only one gasifier is operating at design capacity. In the Base Case, the F-T area is operating at design capacity on syngas and both combustion Turbines are operating at maximum (almost design) capacity with natural gas added to the syngas and F-T gas to fully fire both combustion turbines. The Minimum Natural Gas Use Case is similar to the Base Case except that less natural gas is brought into the plant so that only one combustion turbine is fired at its maximum capacity. In the Maximum Power Case, the F-T area is not operating and both combustion turbines are fired at their maximum capacity on syngas and natural gas.

The remaining 0.19% combined operating percentage for the gasification blocks and the combustion turbines that not shown in Cases A through E include the rare situations when only syngas is available (and the F-T area and gas turbines are not) or when all equipment is down.

Table IV.3 shows the design and annual average feed and product rates for all three operating scenarios. All three cases always use the gasification block to its maximum capacity, consume the same amount of coke and flux, and produce the same amount of byproduct slag. The Base Case and the Minimum Natural Gas Use Case both have the object of producing F-T liquids whenever syngas is available, and therefore, both cases produce the same amount F-T liquid fuel precursors (3,938 bpd). However, the Minimum Natural Gas Use Case only uses supplemental natural gas to co-fire one combustion turbine to its maximum capacity even when both turbines are available, and the Base Case always fires both turbines to their maximum capacity when they are available. Thus, the Minimum Natural Gas Use Case consumes on average much less natural gas than the Base Case (3,794 MMBtu/day vs. 8,830 MMBtu/day).

The Maximum Power Case discontinues F-T liquids production when insufficient syngas is available to fully fire both gasifiers and to produce the F-T liquids in order to maximize power production with minimal natural gas use. Thus, this case brings in supplemental natural gas to fully fire both gasifiers when they are available just as the Base Case does. Since the Maximum Power Case shuts down F-T liquids production when only one gasifier is operating, it produces less F-T liquids on average (3,273 bpd vs. 3,938 bpd). However, the Maximum Power Case exports slightly more power than the Base Case because it consumes slightly less power when the F-T plant is not operating (572.8 MW vs. 572.5 MW).

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

Table IV.1

Plant Configurations and Sections Capacities of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

Case Identification	**Subtask 2.1**	**Subtask 1.3**
Case Description	**Coke Power & Liquids Plant**	**Next Plant**
	Number of Trains and Section Capacity (Note 1)	
Gasification & Power Blocks		
Air Separation Unit (ASU)	2x50	2x50
Coke Handling	1x100	1x100
Slurry Preparation	2x60	2x60
Slurry Feed	3x50	3x50
Gasification (though HTHRU)	3x50	3x50
Slag Handling	1x100	1x100
Dry Particulate Removal		
Cyclone	3x50	3x50
Particulate Filters	3x50	3x50
Wet Scrubbing System	2x50	2x50
LTHR/AGR	2x50	2x50
SRU	2x50	2x50
Combustion Turbine	2x50	2x50
Steam Turbine	1x100	1x100
Byproduct Production		
Hydrogen	----	2x50
Fischer-Tropsch Process Area	1x100	----
Number of Fischer-Tropsch Reactors	1	0
Scheduled Outages per Gasification Train	15.34%	15.34%
Spare Gasifier Vessels (1 per train)	No	No
Potential Syngas Availability, % (note 2)		
From 2 of 3 Gasification Trains (@100% rate)	86.85%	86.85%
From 1 of 3 Gasification Trains (@50% rate)	99.63%	99.63%
Equivalent Availability (note 3)	93.24%	93.24%

Notes:
1. Capacity percentages are based on the total plant design capacity.
2. This is the clean syngas availability (including scheduled outages) without any downstream constraints on the consumption or use of the syngas; e.g., when exporting syngas to a pipline.
3. Equivalent availability is the annual average capacity expressed as a fraction of the design capacity.

Table IV.2

Equipment Status of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction for the Three Operating Scenarios

Case & % of Time+	Equipment	Base Case Full Gas Backup	Minimal Natural Gas Use Case	Maximum Power Case
		Feed or Status	Feed or Status	Fuel or Status
A 79.43%	Gasifier 1	Coke	Coke	Coke
	Gasifier 2	Coke	Coke	Coke
	Gas Turbine 1	Syngas & F-T Gas	Syngas & F-T Gas	Syngas & F-T Gas
	Gas Turbine 2	Syngas & F-T Gas	Syngas & F-T Gas	Syngas & F-T Gas
	F-T Liquids Area	Syngas	Syngas	Syngas
B 7.25%	Gasifier 1	Coke*	Coke*	Coke*
	Gasifier 2	Coke*	Coke*	Coke*
	Gas Turbine 1	Syngas & F-T Gas	Syngas & F-T Gas	Syngas & F-T Gas
	Gas Turbine 2	Not Operating	Not Operating	Not Operating
	F-T Liquids Area	Syngas	Syngas	Syngas
C 4.44%	Gasifier 1	Coke	Coke	Coke
	Gasifier 2	Not Operating	Not Operating	Not Operating
	Gas Turbine 1	Syngas, F-T Gas and Natural Gas	Syngas, F-T Gas and Natural Gas	Syngas & Natural Gas
	Gas Turbine 2	Syngas, F-T Gas and Natural Gas	Not Operating	Syngas & Natural Gas
	F-T Liquids Area	Syngas	Syngas	Not Operating
D 8.32%	Gasifier 1	Coke	Coke	Coke*
	Gasifier 2	Not Operating	Not Operating	Not Operating
	Gas Turbine 1	Syngas, F-T Gas and Natural Gas	Syngas, F-T Gas and Natural Gas	Syngas
	Gas Turbine 2	Not Operating	Not Operating	Not Operating
	F-T Liquids Area	Syngas	Syngas	Not Operating
E 0.37%	Gasifier 1	Not Operating	Not Operating	Not Operating
	Gasifier 2	Not Operating	Not Operating	Not Operating
	Gas Turbine 1	Natural Gas	Natural Gas	Natural Gas
	Gas Turbine 2	Not Operating	Not Operating	Not Operating
	F-T Liquids Area	Not Operating	Not Operating	Not Operating

+ Percent of operating time is based only on the combined gasification and power blocks.
* Operations at reduced capacity and/or excess syngas is flared.

Table IV.3

Design and Daily Average Feed and Product Rates of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction for the Three Operating Scenarios

Case	Design	Base Case	Minimum Natural Gas Use Case	Maximum Power Case
Feed Rates				
Coke, dry TPD	5,375	4,805	4,805	4,805
Flux, TPD	109.7	98.1	98.1	98.1
Natural Gas, MMBtu/hr	23.2	369	159	168
Product Rates				
Power, MW	617.0	572.5	554.7	572.8
Sulfur, TPD	370.6	331.3	331.3	331.3
Slag, TPD	193.6	173.1	173.1	173.1
F-T Liquids, bpd	4,125	3,938	3,938	3,273

Section 5

Financial Analysis

The following financial analysis was performed using the discounted cash flow (DCF) model that was developed by Bechtel Technology and Consulting (now Nexant Inc.) for the DOE as part of the Integrated Gasification Combined Cycle (IGCC) Economic and Capital Budgeting Practices Task.[10] This model calculates investment decision criteria used by industrial end-users and project developers to evaluate the economic feasibility of IGCC projects.

5.1 Financial Model Input Data

The required input information to the DCF financial model is organized into two distinct input areas that are called the Plant Input Sheet and the Scenario Input Sheet. The Plant Input Sheet contains data that are directly related to the specific plant as follows.

Data Contained on the Plant Input Sheet

- Project summary information
- Plant output and operating data
- Capital costs
- Operating costs and expenses

The Scenario Input Sheet contains data that are related to the general economic environment that is associated with the plant as well as some data that are plant related. The data on the Scenario Input Sheet are shown below.

Data Contained on the Scenario Input Sheet

- Financial and economic data
- Fuel data
- Tariff assumptions
- Construction schedule data
- Startup information

For all cases, the EPC spending pattern was adjusted to reflect forward escalation during the construction period since the EPC cost estimate is an “overnight” cost estimate based on mid-year 2000 costs.

Items that were excluded in the cost estimate, such as spares, owner’s cost, contingency and risk are included in the financial analysis.

Table V.1 summarizes the basic input parameters to the financial model. The daily average plant input and output flow rates are given in Table IV.3

[10] Nexant, Inc., “Financial Model User’s Guide – IGCC Economic and Capital Budgeting Evaluation”, Report for the U. S. Department of Energy, Contract DE-AMO1-98FE64778, May 2000.

Table V.1

Basic Financial Model Input Parameters

Parameter	Value
Financial Parameters	
Owner's Contingency (% of EPC Cost)	5.0%
Development Fee (% of EPC Cost)	1.23%
Start-up Cost (% of EPC Cost)	1.50%
Additional Financing Cost, EPC Contingency, Risk and Fees, etc.	5.0%
Loan Amount (% of Cost)	80%
Loan Interest Rate	10% & 8%
Loan Financing Fee	3.0%
Loan Repayment Term, years	15
Income Tax Rate	40%
Construction Period, years	15
Start Up	
First Year's Average Capacity	60%
Prices	
Coke, $/dry ton	0.00
Flux, $/ton	5.00
Natural Gas, $/MMBtu HHV or $/Mscf *	2.60
Fischer-Tropsch Liquids, $/bbl	30.00
Electric Power, $/MW	27.00
Sulfur, $/ton	30.00
Slag, $/ton (15% water)	0.00
Annual Inflation Rates	
Coke $/dry ton	1.2%
Natural Gas, $/HHV MMBtu	3.9%
Fischer-Tropsch Liquids	3.1%
Electric Power, $/MW	1.7%
Sulfur, $/ton	0.0%
Slag, $/ton (15% water)	0.0%

* Natural gas is assumed to have a HHV Btu content of 1,000 Btu/scf.

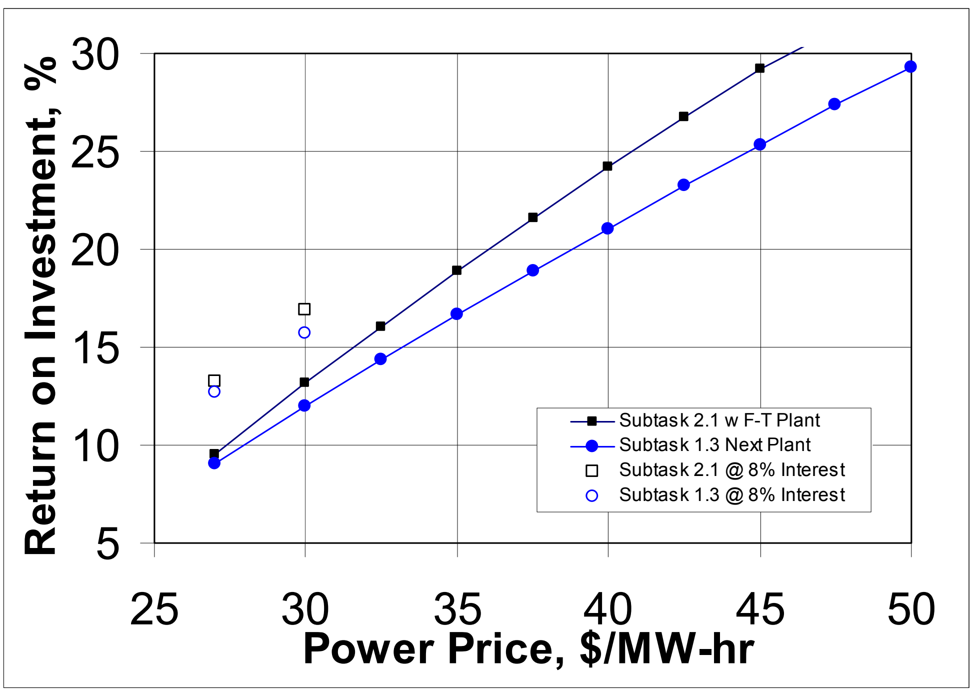
5.2 Financial Model Results

Figure 5.1 shows the return on investment (ROI) as a function of the export power price for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction under the Base Case (full natural gas backup) operating scenario and compares it with the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. The solid lines are with a 10% loan interest rate. The Subtask 2.1 plant has a 12.0% ROI at an export power price of 29.04 $/MW-hr, and the Subtask 3.1 Next Plant has a 12.0% ROI at a power price of 30.02 $/MW-hr. At a 27 $/MW-hr export power price, the Subtask 2.1 plant has a 9.50% ROI, and the Subtask 1.3 plant has a 9.05% ROI.

An 8% loan interest rate significant increases the return on investment by about 3.7 ROI percent as shown by the open symbols on the figure. At an 8% loan interest rate, and a 27 $/MW-hr power selling price, the Subtask 2.1 plant has a 13.24% ROI under the Base Case operating scenario, and the Subtask 1.3 Next Plant has a 12.70% ROI. These results also are given in tabular form in Table V.3 later in the report when the three alternate operating scenarios are compared.

Figure 5.1

Return on Investment vs. Power Price for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant



It is difficult to predict the future value of either power, natural gas and/or the liquid fuel precursors. The power price is related to the natural gas price which can be highly variable. The liquid fuel precursors price is related to the crude oil price which also can be highly variable both because of market forces and the influence of international politics. Various studies have been made which attempt to relate the value of the F-T liquids to that of crude oil by replacing crude oil in the refinery feed stream with the F-T liquids. The resulting values for the F-T liquids generally are above the crude oil values, but the specific amount can range from only a couple of $/bbl up to 10 $/bbl depending upon the refinery configuration, the crude oils being replaced, and the required refinery product mix.[11]

Figure 5.2 shows the effect of the liquid fuel precursors selling price on the return on investment versus the power selling price for the Subtask 2.1 plant under the Base Case operating scenario with a 10% loan interest rate. The dashed 30 $/bbl line is the same line as shown on the previous figure for the Subtask 2.1 plant. At a 27 $/MW-hr power selling price, reducing the liquid fuel precursors selling price by 10 $/bbl reduces the return on investment by about 4.4 ROI percent, and increasing the liquids fuel precursors selling price by 10 $/bbl increases the return on investment by about 4.0 ROI percent. As the power price increases, the effect of the liquid fuel precursors selling price decreases to 2.5 to 3.0 ROI percent at a 45 $/MW-hr power selling price. This is because the portion of the revenue generated from the liquids now is a smaller portion of the total revenue generated by the liquids and power sales.

Figure 5.2

Return on Investment vs. Power Price
Showing the Effect of the Liquid Fuel Precursors Price for the
Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction

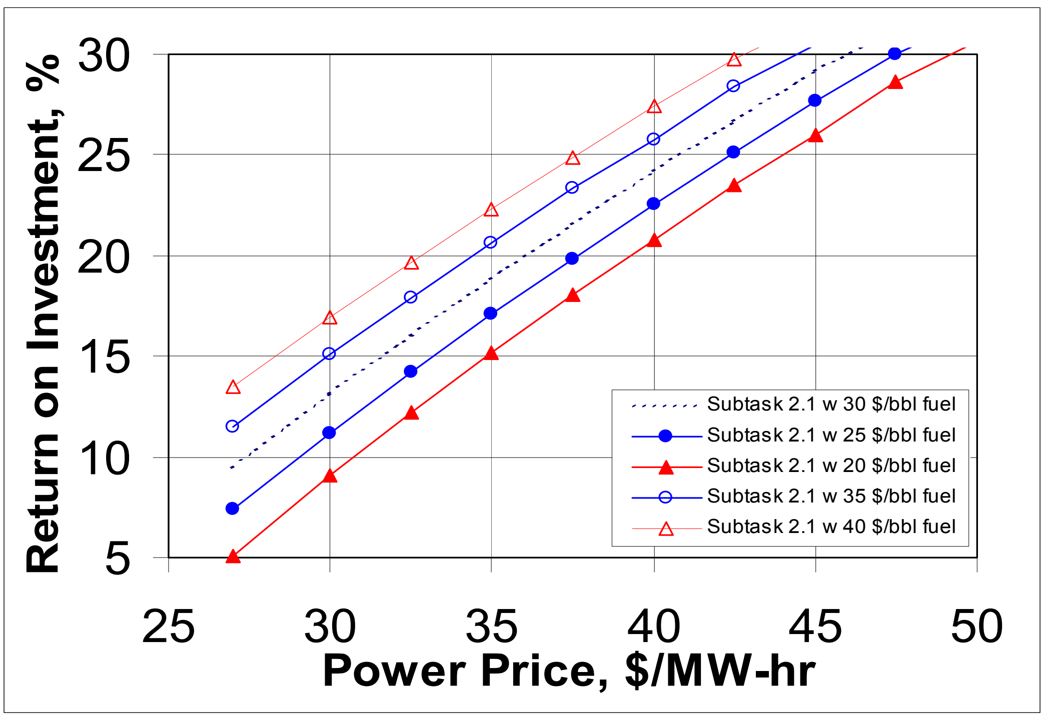


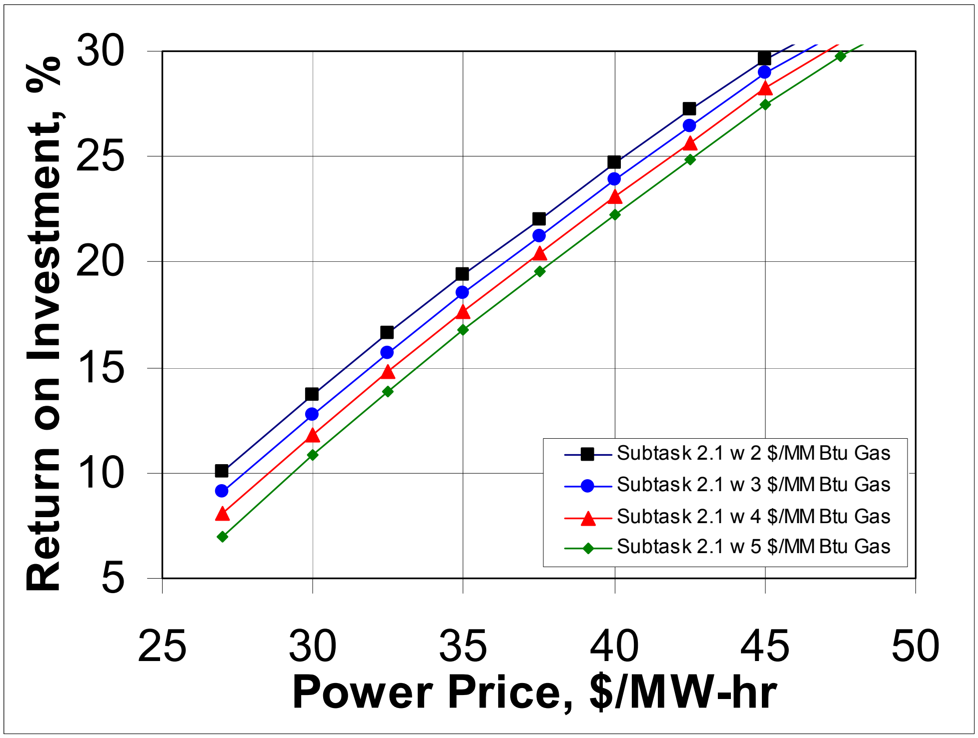
Figure 5.3 shows the effect of the natural gas price on the return on investment versus the power selling price for the Subtask 2.1 plant under the Base Case operating scenario with a 10% loan interest rate. The 2.6 $/MMBtu HHV natural gas price line is not shown on this figure to avoid congestion, but if it were shown, it would lie about midway between the 2.0 and 3.0 $/MMBtu lines (the upper two lines). Increasing the natural gas price from 2.0 to 5.0 $/MMBtu lowers the ROI by about 2.9 ROI percent at a 27 $/MW-hr power selling price. As the power price increases,

[11] Marano, J. J., Rogers, S., Choi, G. N., and Kramer, S. J., "Product Valuation of Fischer-Tropsch Derived Fuels," ACS National Meeting, Washington, D. C., August 21-6, 1994.

the effect of the natural gas price decreases to about 2.1 ROI percent at a 45 $/MW-hr power selling price for the same natural gas selling price difference.

Figure 5.3

Return on Investment vs. Power Price
Showing the Effect of the Natural Gas Price for the
Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction



After commissioning all plants undergo a “learning curve” during which problem areas are corrected, inadequate equipment is modified or replaced, and adjustments are made. Consequently, performance improves as measured by increased capacity and/or improved on-stream factors. Figure 5.4 shows the effect of improved syngas availability on the return on investment. As the syngas availability improves, the amount of natural gas is reduced until it almost disappears at the unattainable 100% syngas availability since a small amount of natural gas still is used as furnace fuel in the F-T area. At the expected 86.85% syngas availability, the Subtask 2.1 plant has a ROI of 9.50%, and the Subtask 1.3 Next Plant has a ROI of 9.05%. At 90% syngas availability, the ROI of the Subtask 2.1 plant increases to about 10.8%, and that of the Subtask 1.3 Next Plant increases to 10.1%. At the unattainable syngas availability of 100%, the Subtask 2.1 plant will have an expected ROI of 14.9%, and the Subtask 1.3 Next Plant will have an expected 13.4% ROI.

Figure 5.4

Return on Investment vs. Syngas Availability for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

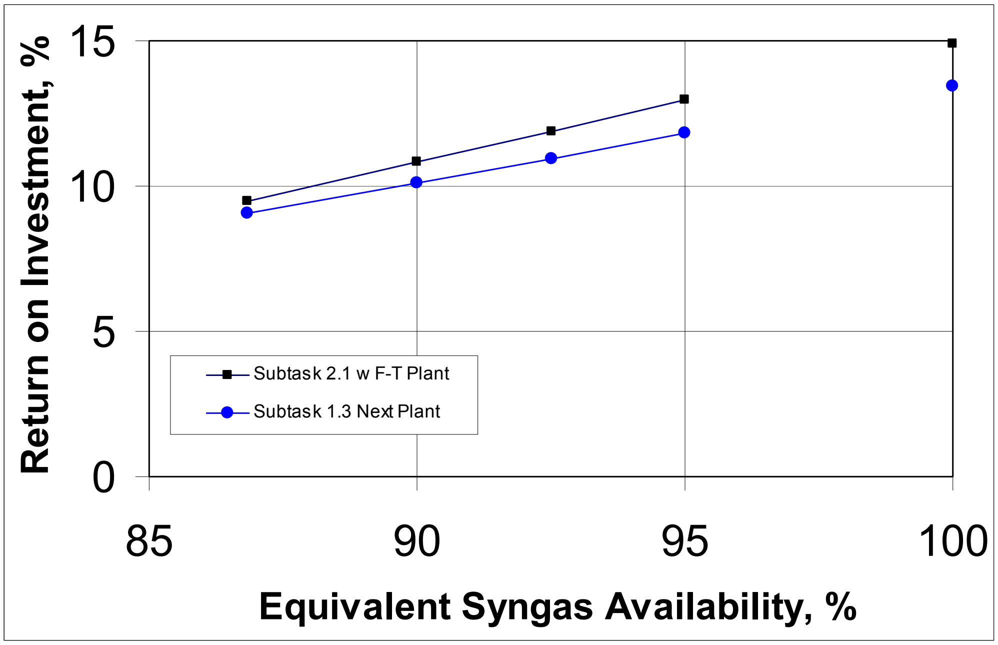


Figure 5.5

Required Power Selling Price for a 12% ROI vs. Syngas Availability for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

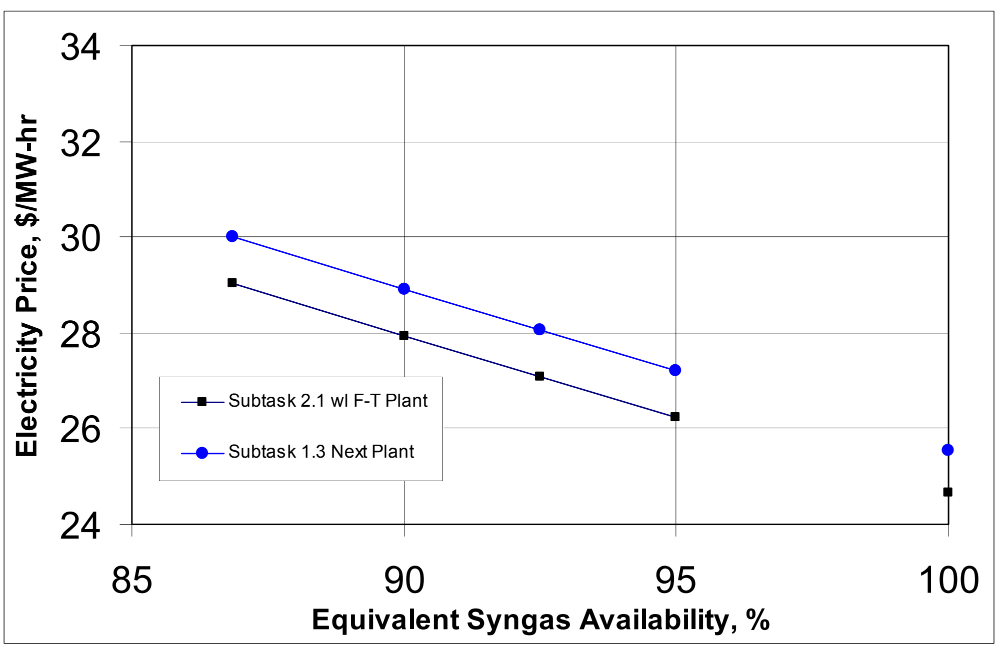


Figure 5.5 shows the effect of improved syngas availability on the required power selling price for a 12% ROI. At the expected 86.85% syngas availability, the Subtask 2.1 plant will require an export power selling price of 29.04 $/MW-hr, and the Subtask 3.1 Next Plant will require a power price of power price of 30.02 $/MW-hr. At 90% syngas availability, the Subtask 2.1 plant will require a power selling price of 27.94 $/MW-hr, and the Subtask 1.3 Next Plant will require a power price of 28.91 $/MW-hr. At the unattainable syngas availability of 100%, the Subtask 2.1 plant will require a power selling price of 24.64 $/MW-hr, and the Subtask 1.3 Next Plant will require a power price of 25.56 $/MW-hr to generate a 12% ROI.

Table V.2 shows the sensitivity of some individual component prices and financial parameters for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction starting from a 12% ROI (with a power price of 29.04 $/MW-hr and a 30.0 $/bbl liquids price). Each item was varied individually without affecting any other item. Most sensitivities are based on a ±10% change from the base value except when either a larger or smaller change is used because it either makes more sense or it is needed to show a meaningful result. The power selling price is the most sensitive product price with a 10% increase resulting in a 3.43% increase in the ROI to 15.43%, and a 10% decrease resulting in a 3.57% decrease in the ROI to 8.43%. A 10% increase in the F-T liquids price to 33.0 $/bbl will cause a 1.18% increase in the ROI to 13.18%, and a 10% decrease in the liquids price to 27.0 $/bbl will result in a 1.20% decrease in the ROI to 10.80%. Changes in the sulfur and slag prices only have a small influence on the ROI.

All the above economic cases were developed with a long-term coke netback price of zero at the refinery gate; i.e., the revenue obtained from the sale of the coke is the same as the expense of transporting it to a site where it is consumed. A decrease in the coke price of 5 $/ton to a negative 5.0 $/ton will increase the ROI by 1.82% to 13.82% and a 5 $/ton increase in the coal price will lower the ROI by 1.82% to 10.18%. A 10% decrease in the natural gas price of 0.26 $/MMBtu from the base natural gas price of 2.60 $/MMBtu will increase the ROI by 0.25% to 12.25%, and a 0.26 $/MMBtu increase in the gas price will lower the ROI by 0.25% to 11.75%.

A 5% decrease in the plant EPC cost from 818 MM$ to 770 MM$ will increase the ROI by 2.32% to 14.32%, and a 5% increase in the plant cost to 859 MM$ will decrease the ROI by 2.15% to 9.85%.

In today's unsettled financial situation, the loan interest rate and project financing conditions can be uncertain. A 20% decrease in the loan interest rate to 8% from the base interest rate of 10% will increase the ROI to 15.75% from 12.00%, and a 20% increase in the interest rate to 12% will lower the ROI by 3.8% to 8.20%. A 20% decrease in the loan amount from 80% to 72% will lower the ROI by 0.57% to 11.43%, and a 20% increase in the loan amount to 88% will increase the ROI by 0.97% to 12.97%. Decreasing the income tax rate by 10% from 40% to 36% will increase the ROI by 0.48% to 12.48%, and a 10% increase in the tax rate to 44% will lower the ROI by 0.51% to 11.49%.

Alternate Operating Scenarios

As discussed previously in Section 4 and Table IV.2, two alternate operating scenarios were proposed in addition to the Base Case scenario. These alternate scenarios are the Minimum Natural Gas Use Case and the Maximum Power Case. In the Base Case supplemental natural gas is used to fully fire both combustion turbines whenever possible when insufficient syngas is available. In the Minimum Natural Gas Use Case, supplemental natural gas is brought in only to fully fire one combustion turbine when insufficient syngas is available. In the Maximum Power Case, F-T liquids production is suspended when insufficient syngas is available to fully fire both combustion turbines. In all cases, when both gasifiers are not operating, supplemental natural gas is used to fire only one combustion turbine when it is available.

Table V.3 compares these two alternate operating scenarios with the Subtask 2.1 Base Case and Subtask 1.3 Next Plant. At the basic economic conditions shown in Table V.1, the Minimum Natural Gas Use Case has the highest ROI of 9.94% followed by the Base Case at 9.50% and the Maximum Power Case at 8.86%. Alternately, the Minimum Natural Gas Use Case has the lowest required power selling price of 28.75 $/MW-hr for a 12% ROI followed by the Base Case at 29.04$/MW-hr and the Maximum Power Case at 29.56 $/Mw-hr. The best case uses the least amount of natural gas followed by the other two cases in increasing order of natural gas use. This order is caused by the low power prices and relative high natural gas price so that the cost of the natural gas used to make the additional power cannot be recovered at the low power price.

The bottom of Table V.3 provides the same comparison between the three operating scenarios but with an 8% loan interest rate. The ROIs are higher and the required power selling prices are reduced, but there is no change in the relative ranking of the three cases.

Figure 5.6 shows the return on investment vs. export power selling price for the three operating scenarios with a 2.60 $/MMBtu natural gas price. The smooth line is the Base Case operating scenario. It is the same line as shown on Figure 5.1. At the low power price of 27 $/MW-hr, the Minimum Natural Gas Use Case has the highest ROI followed by the other two in order of increasing natural gas use. As the export power price increases, the Base Case overtakes the Minimum Natural Gas Use Case at about 34 $/MW-hr, and at higher power prices it has the highest ROI. Somewhere at a higher power price than shown on the graph, the Maximum Power Case will overtake the Base Case and have the highest ROI.

Thus, the best operating scenario (i.e.; the one that produces the highest ROI) for the plant depends upon the relationship between the natural gas price and the power price. At relatively low power prices and higher natural gas prices, the best operating scenario is the one that uses the minimum amount of natural gas. As the power prices increase relative to the natural gas price, the Base Case scenario will be the best one, and at very high relative power prices, the Maximum Power Production Case will be the best one.

Figure 5.7 shows the return on investment vs. export power selling price for the three operating scenarios with a 5.00 $/MMBtu natural gas price. Again, the smooth line is the Base Case operating scenario. Under these conditions, the Minimum Natural Gas Use Case is the best one because it uses the least amount of the high priced gas. The Maximum Power Production Case is the next best case because it reduces liquids production, and the liquids are relatively low priced compared to the natural gas price. The Base Case is the worst case.

Historically, on a $/Btu basis, natural gas has been less costly than liquid fuels derived from petroleum. The 5.00 $/MMBtu natural gas price is close to that of the liquids at 30 $/bbl. Assuming the liquids have a HHV Btu content of about 5.6 MMBtu/bbl, then they are worth about 5.36 $/MMBtu. Thus, at this price differential, it is not economically attractive to convert natural gas to liquid fuel precursors in this capital-intensive process.

However, when one or both gasifiers are unavailable, the decision as to which operating scenario (Base Case, Minimum Natural Gas Use Case, or Maximum Power Case) will be used to maximize revenue should be based on the relative values of the current feed and product prices. Detailed calculations, that are given in Appendix A, have shown that for the case when one gasifier is unavailable (Cases C and D of Table IV.2), the maximum return generally can be obtained by the Maximum Power Case scenario when the liquids are 30 $/bbl, the power prices are 27 $/MW-hr or above, and natural gas prices are less than 5.0 $/MMBtu. In this scenario the F-T liquids area is shut down to keep both combustion turbines operating on the maximum amount of syngas supplemented by the minimum amount of natural gas. At higher gas prices or lower power prices, the Minimum Natural Gas Use Case scenario may generate more net revenue.

In the situation where both gasifiers are not operating when natural gas prices are so high that the gas costs more than the revenue that are generated from selling the export power, the best operating scenario may be to shut down and not produce any power. For example, with 4.0 $/MMBtu gas, power prices have to be above 30 $/MW-hr to justify operations, and with 5.0 $/MMBtu gas, power prices have to be above about 38 $/MW-hr to show a positive return.

Figure 5.6

Return on Investment vs. Power Price for the Three Operating Scenarios with a 2.60 $/MMBtu Natural Gas Price

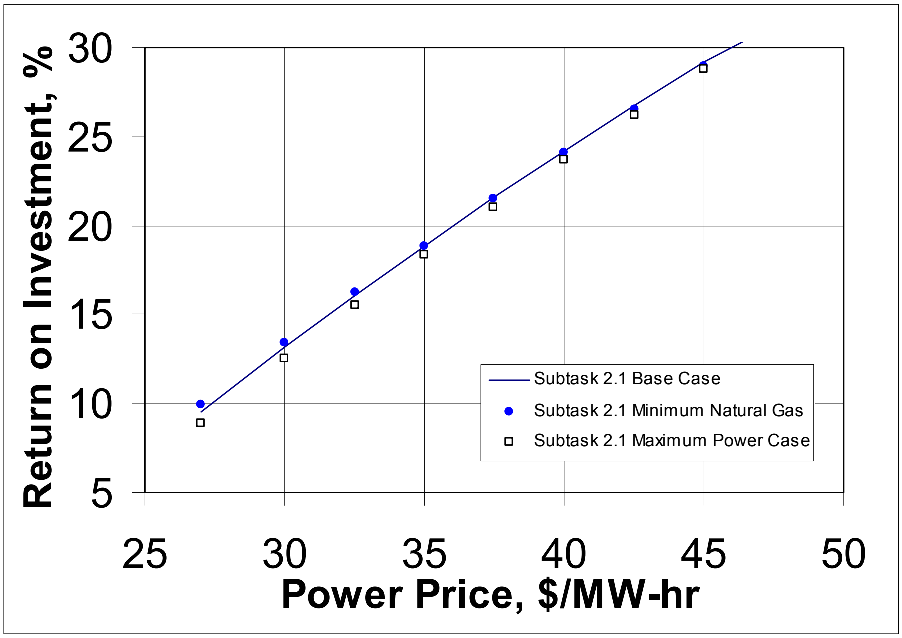


Figure 5.7

Return on Investment vs. Power Price for the Three Operating Scenarios with a 5.00 $/MMBtu Natural Gas Price

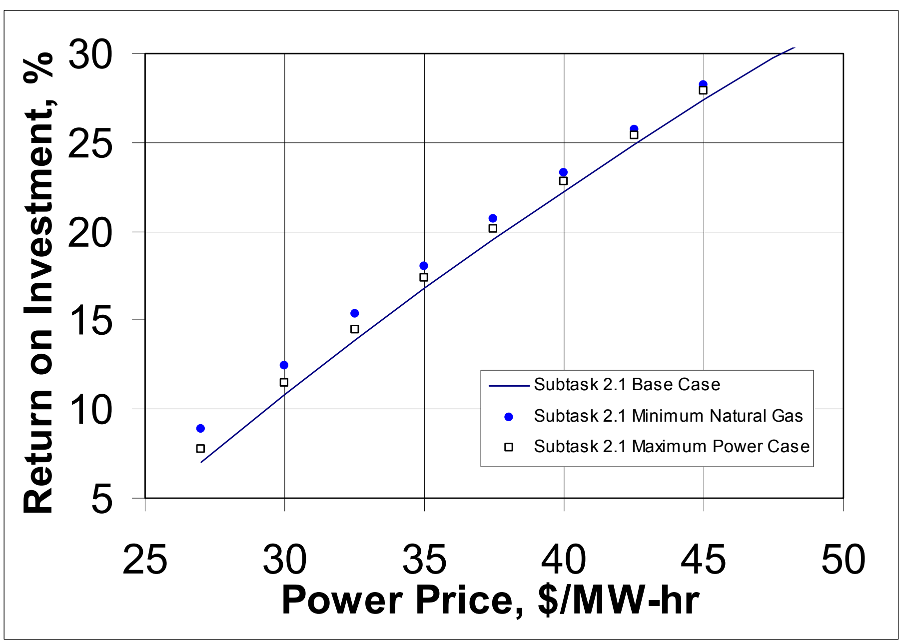


Table V.2

Sensitivity of Individual Component Prices on the Return on Investment
For the Subtask 2.1 Small F-T Case with a 12% ROI (at a Power Price of 29.04 $/MW-hr)

	Decrease				Increase		
	ROI	Value	% Change	Base Value	% Change	Value	ROI
Products							
Power	8.43%	26.14 $/MW-hr	-10%	**29.04 $/MW-hr**	+10%	31.94 $/MW-hr	15.43%
F-T Liquids	10.80%	27.0 $/bbl	-10%	**30.0 $/bbl**	+10%	33.0 $/bbl	13.18%
Sulfur	11.93%	27.0 $/ton	-10%	**30.0 $/ton**	+10%	33.0 $/ton	12.07%
Slag	11.94%	-5.0 $/ton	---	**0 $/ton**	---	5.0 $/ton	12.06%
Feeds							
Coke	13.82%	-5.0 $/ton	---	**0 $/ton**	---	5.0 $/ton	10.18%
Natural Gas	12.25%	2.34 $/MMBtu	-10%	**2.60 $/MMBtu**	+10%	2.86 $/MMBtu	11.75%
Flux	12.04%	0 $/ton	100%	**5.0 $/ton**	+100%	10.0 $/ton	11.96%
Financial							
EPC Cost	13.14%	797.5 MM$	-2.5%	**818 MM$**	+2.5%	838.4 MM$	10.91%
EPC Cost	14.32%	770.1 MM$	-5.0%	**818 MM$**	+5.0%	858.8 MM$	9.85%
Interest Rate	15.75%	8%	-20%	**10%**	+20%	12%	8.20%
Loan Amount	11.43%	72%	-20%	**80%**	+20%	88%	12.97%
Tax Rate	12.48%	36%	10%	**40%**	+10%	44%	11.49%

Note: Products and Feeds each are listed in decreasing sensitivity.

Table V.3

Return on Investments and Required Product Selling Prices for the Three Subtask 2.1 Operating Scenarios and the Subtask 1.3 Next Plant (with a Natural Gas Price of 2.60 $/MMbtu)

	Subtask 2.1 with F-T Liquids Plant			
	Base Case	Minimum Natural Gas Use Case	Maximum Power Case	Subtask 1.3 Next Plant
With a 10% Loan Interest Rate				
Return on Investment with a 27 $/MW-hr Power Price	9.50%	9.94%	8.86%	9.05%
Required Power Selling Price for a 12% ROI, $/MW-hr	29.04	28.75	29.56	30.02
With a 8% Loan Interest Rate				
Return on Investment with a 27 $/MW-hr Power Price	13.24%	13.66%	12.58%	12.70%
Required Power Selling Price for a 12% ROI, $/MW-hr	26.04	25.65	26.55	26.32

Section 6

Summary

A design for a coke gasification power plant with liquid fuel precursors coproduction using Fischer-Tropsch technology has been developed. The plant consumes 5,375 tpd of coke (dry basis), 110 tpd of flux and 23.2 MMBtu/hr (HHV) of natural gas to produce 617 MW of export power and 4,125 bpd of liquid fuel precursors. It also produces 371 tpd of elemental sulfur and 194 tpd of slag. The plant is located on the U.S. Gulf Coast adjacent to a petroleum refinery.

The design of the plant was developed from the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant that processes about the same amount of coke and produces 80 MMscfd of 99% hydrogen and 980,000 lb/hr of 750°F/700 psia steam for the adjacent refinery in addition to 474 MW-hr of export power. It also produces about the same amounts of byproduct sulfur and slag. The Subtask 1.3 design was modified by replacing the hydrogen facilities with the F-T hydrocarbon synthesis area. The unconverted syngas and light hydrocarbon products from the F-T area are compressed and sent to the combustion turbines to generate power. A larger steam turbine is used to generate additional power from the steam that previously was sent to the refinery.

The Fischer-Tropsch hydrocarbon synthesis area basically consists of three sections; final sulfur removal, slurry-bed F-T reactor, and product recovery sections. The final sulfur removal section contains a second COS hydrolysis reactor which converts the residual COS in the syngas to H_2S. The H_2S is removed from the syngas by ZnO adsorption beds. The sulfur-free syngas is fed to the slurry-bed F-T reactor which converts syngas to hydrocarbons over an iron-based catalyst. The heat of reaction is removed by generation of 440°F/375 psia steam inside tubes that are placed within the slurry-bed. The lighter hydrocarbon products and unconverted syngas leave the reactor as vapors and are cooled by cooling water to condense hydrocarbons and recover them as liquids. The unconverted syngas and non-condensable light hydrocarbons (primarily C1 through C3s) are compressed and sent to the combustion turbines for power generation. The heavier products are removed from the reactor as liquids, separated from the entrained catalyst by filtration, cooled, mixed with the lighter hydrocarbons, and sent to the adjacent petroleum refinery for separation and incorporation into liquid transportation fuels.

The F-T liquid fuel precursors essentially are a bottomless, sulfur-free crude oil. Basically they are straight-chain 1-olefins and paraffins without any aromatics. The diesel fraction has a very high cetane number (>70) and is a premium diesel fuel blending component. The naphtha fraction is a low octane material that requires further upgrading for use as a gasoline blending component. However, it is an excellent feedstock for an ethylene cracker. Linear programming studies have shown that the F-T liquid fuel precursors may be worth up to 10 $/bbl more than crude oil depending upon the specific refinery configuration and product demands.

This coke plant with liquid fuel precursors coproduction has a slightly better return on investment than the Subtask 1.3 Next Plant. At export power prices below 30 $/MW-hr, this plant can have a return on investment greater than 12% when the liquid fuel precursors are worth 30 $/bbl.

During development of this design, several ideas were generated for optimizing the process, such as using activated carbon to remove the sulfur from the syngas going to the F-T reactor, using a refrigeration system to increase the liquids recovery, and using a larger F-T reactor to produce more liquids at the expense of power production. These ideas will be incorporated in the Subtask 2.2 plant design.

Appendix A – Attachment

Subtask 2.1

Revenue Calculations for the Three Operating Scenarios

Appendix A - Attachment

Revenue Calculations for the Three Operating Scenarios

Tables A-I through A-III contain daily net revenue calculations for the three operating scenarios for the design basis (Case A), an annual average operating day and cases C, D and E of Table IV.2. Table A-I contains the daily net revenue projections for the Base Case operating scenario which uses as much supplemental natural gas as needed to keep both combustion turbines operating at full capacity whenever possible. Table A-II contains the daily net revenue projections for the Minimum Natural Gas Use Case scenario which only uses the minimum amount of natural gas to keep only one combustion turbine operating at maximum capacity whenever possible. Table A-III contains the daily net revenue projections for the Maximum Power Case operating scenario which maximizes the amount of syngas going to the combustion turbines by suppressing liquids production (and minimizes supplemental natural gas use) whenever one gasifier is not operating.

The top of each table contains a description of the operating units for each case under that operating scenario. Below these items is a listing of the feed and product flow rates for each case under that operating scenario. The remainder of that table lists the daily net revenue that would be produced given the feed and product prices at the left side of the table. All cases have the same fuels price of 30 $/bbl, sulfur price of 30 $/ton, and flux price of 5 $/ton. Only the power and natural gas prices change. Since the coke and slag have a net price of zero, they have been omitted from these tables.

Below this are seven groups of daily net revenue calculations for various power and natural gas price combinations. All revenue calculations are based on operating the combustion turbines at maximum capacity on syngas and/or supplemental natural gas, when possible. Under certain circumstances where the revenue calculation shows a net loss, it would be best to shut down the operation until conditions improve. However, there may be circumstances when this cannot be done (such as a contractual obligation), or the syngas outage will be of a very short duration where the shutdown and start-up operations would be almost back-to-back. Another alternative that could be considered is operating at reduced capacity. For simplicity, operations at reduced capacity are not considered in Tables A-I through A-III.

The net daily revenues for design case (Case A), Annual Average Case, and Case E are the same in all three operating scenario tables. Case E operations can have negative net daily revenues at low product prices and high natural gas prices. When the net revenue under the Case E operating scenario is negative, consideration should be given to shutting down, or at least, running at reduced capacity, if possible.

Revenue calculations for Case B are not shown because they involve either scaling back gasifier operations to reduce excess syngas production and/or flaring excess syngas. The method of operation would be dependent upon how long the gas turbine is expected to be unavailable. Another scenario that is not shown is the case when the F-T liquids fuel production area is not operating. Under these conditions with both gasifiers operating, syngas production would have to be reduced and/or syngas would have to be flared. Operations for these cases would be the same under all three operating scenarios and are not considered in Tables A-I through A-III.

Cases C and D (with only one operating Gasifiers) are the interesting cases. Under the Base Case operating scenario, supplemental natural gas would be consumed to maximize power production in order to fully load one or both combustion turbines, if available. In Case C, both combustion turbines are available, and in Case D, only one turbine is operable. The boxed in revenues show the larger net revenue for either Cases C or D. Under conditions of high natural gas costs and low power prices, increased net revenues can be obtained by only operating one of the two available combustion turbines, and shutting down the other one. For example, looking at the fourth group of net revenue calculations (with a 4 $/MMBtu natural gas price), it is best to operate only one gas turbine even when both are available when the power price is less than 35 $/MW-hr. At higher power prices, additional supplemental natural gas should be used to fire both turbines at maximum capacity because this will produce higher net revenues.

Cases C and D are the same under the Minimum Natural Gas Use scenario as shown in Table A-II. They also are the same as for Case D for the Base Case as shown in Table A-I. Under this operating scenario, when only one gasifier is available, F-T liquids production will be maintained and only one combustion turbine will be operating even if two are available in order to minimize supplemental natural gas use.

Under the Maximum Power Case operating scenario, Case C always has larger net revenues than Case D as shown in Table A-III. In this scenario, the F-T liquids production is shut down when only one gasifier is available, and all syngas is used for power production. When only one turbine is available, either the gasifier is operated at reduced capacity and/or excess syngas is flared. When both combustion turbines are available, the maximum net revenue is obtained when supplemental natural gas is used to operate both of them at maximum capacity. However, at relatively low power prices compared to the natural gas price, reducing the power output by cutting back on the amount of supplemental natural gas may increase the daily net revenue.

The boxed in cases in Table A-III have larger daily net revenues than the corresponding cases in either of the previous two tables. Thus, at high power prices, shutting down the F-T liquid fuels area and maximizing the power production from syngas can increase the daily net revenues.

In a similar manner, additional tables similar to Tables A-I though A-III can be constructed to develop the operating scenario with the highest daily net revenue for other feed and product price combinations.

Table A-I
Daily Net Revenue Calculations for the Base Case Operating Scenario

BASE CASE OPERATING SCENARIO	**Case A Design**	**Annual Average**	**Case C**	**Case D**	**Case E**
Equipment Status					
Number of Operating Gasifiers	2	Mixed	1	1	0
Number of Operating Combustion Turbines	2	Mixed	2	1	1
F-T Fuels Production	Running	Mixed	Running	Running	Down
Feed and Product Flow Rates					
Export Power, MW-hr/day	14,808	13,740	10,668	5,532	6,112.8
Liquid Fuels, bpd	4,125	3,939	4,125	4,125	0
Sulfur, tpd	370.6	331.3	185.3	185.3	0
Natural Gas, Mscfd	535	8,830	64,303	17,623	46,416
Flux, tpd	109.7	98.1	54.85	54.85	0

Feed and Product Prices									
Power $/MW-hr	Fuels $/bbl	Sulfur $/ton	Gas $/MMBtu	Flux $/ton	Net Revenue, M$/day	Net Revenue, M$/day	Net Revenue, M$/day	Net Revenue, M$/day	Net Revenue, M$/day
20	30	30	2	5	429.4	384.8	213.8	$\boxed{204.4}$	29.4
27	30	30	2	5	533.1	480.9	$\boxed{288.5}$	243.2	72.2
30	30	30	2	5	577.5	522.2	$\boxed{320.5}$	259.7	90.6
35	30	30	2	5	651.5	590.9	$\boxed{373.8}$	287.4	121.1
40	30	30	2	5	725.6	659.6	$\boxed{427.1}$	315.1	151.7
45	30	30	2	5	799.6	728.3	$\boxed{480.5}$	342.7	182.2
50	30	30	2	5	873.6	797.0	$\boxed{533.8}$	370.4	212.8
20	30	30	2.6	5	429.1	379.5	175.2	$\boxed{193.9}$	1.6
27	30	30	2.6	5	532.7	475.6	$\boxed{249.9}$	232.6	44.4
30	30	30	2.6	5	577.2	516.9	$\boxed{281.9}$	249.2	62.7
35	30	30	2.6	5	651.2	585.6	$\boxed{335.2}$	276.8	93.3
40	30	30	2.6	5	725.2	654.3	$\boxed{388.6}$	304.5	123.8
45	30	30	2.6	5	799.3	723.0	$\boxed{441.9}$	332.2	154.4
50	30	30	2.6	5	873.3	791.7	$\boxed{495.2}$	359.8	185.0
20	30	30	3	5	428.9	375.9	149.5	$\boxed{186.8}$	-17.0
27	30	30	3	5	532.5	472.1	224.2	$\boxed{225.5}$	25.8
30	30	30	3	5	577.0	513.3	$\boxed{256.2}$	242.1	44.1
35	30	30	3	5	651.0	582.0	$\boxed{309.5}$	269.8	74.7
40	30	30	3	5	725.0	650.7	$\boxed{362.8}$	297.4	105.3
45	30	30	3	5	799.1	719.4	$\boxed{416.2}$	325.1	135.8
50	30	30	3	5	873.1	788.1	$\boxed{469.5}$	352.8	166.4
20	30	30	4	5	428.3	367.1	85.2	$\boxed{169.2}$	-63.4
27	30	30	4	5	532.0	463.3	159.9	$\boxed{207.9}$	-20.6
30	30	30	4	5	576.4	504.5	191.9	$\boxed{224.5}$	-2.3
35	30	30	4	5	650.5	573.2	245.2	$\boxed{252.2}$	28.3
40	30	30	4	5	724.5	641.9	$\boxed{298.5}$	279.8	58.8
45	30	30	4	5	798.5	710.6	$\boxed{351.9}$	307.5	89.4
50	30	30	4	5	872.6	779.3	$\boxed{405.2}$	335.1	120.0
20	30	30	5	5	427.8	358.3	20.9	$\boxed{151.6}$	-109.8
27	30	30	5	5	531.5	454.4	95.6	$\boxed{190.3}$	-67.0
30	30	30	5	5	575.9	495.7	127.6	$\boxed{206.9}$	-48.7
35	30	30	5	5	649.9	564.4	180.9	$\boxed{234.5}$	-18.1
40	30	30	5	5	724.0	633.1	234.2	$\boxed{262.2}$	12.4
45	30	30	5	5	798.0	701.8	287.6	$\boxed{289.9}$	43.0
50	30	30	5	5	872.0	770.5	$\boxed{340.9}$	317.5	73.6
20	30	30	6	5	427.3	349.4	-43.4	$\boxed{133.9}$	-156.2
27	30	30	6	5	530.9	445.6	31.3	$\boxed{172.7}$	-113.5
30	30	30	6	5	575.3	486.8	63.3	$\boxed{189.3}$	-95.1
35	30	30	6	5	649.4	555.5	116.6	$\boxed{216.9}$	-64.5
40	30	30	6	5	723.4	624.2	169.9	$\boxed{244.6}$	-34.0
45	30	30	6	5	797.5	692.9	223.3	$\boxed{272.2}$	-3.4
50	30	30	6	5	871.5	761.6	276.6	$\boxed{299.9}$	27.1
20	30	30	7	5	426.7	340.6	-107.7	$\boxed{116.3}$	-202.7
27	30	30	7	5	530.4	436.8	-33.1	$\boxed{155.0}$	-159.9
30	30	30	7	5	574.8	478.0	-1.0	$\boxed{171.6}$	-141.5
35	30	30	7	5	648.9	546.7	52.3	$\boxed{199.3}$	-111.0
40	30	30	7	5	722.9	615.4	105.6	$\boxed{227.0}$	-80.4
45	30	30	7	5	796.9	684.1	159.0	$\boxed{254.6}$	-49.8
50	30	30	7	5	871.0	752.8	212.3	$\boxed{282.3}$	-19.3

$\boxed{}$ = Best of the C and D cases (cases with only one gasifier operating).

Table A-II
Daily Net Revenue Calculations for the
Minimum Natural Gas Use Case Operating Scenario

MINIMUM GAS USE CASE OPERATING SCENARIO					Case A Design	Annual Average	Case C	Case D	Case E
Equipment Status									
Number of Operating Gasifiers					2	Mixed	1	1	0
Number of Operating Combustion Turbines					2	Mixed	2	1	1
F-T Fuels Production					Running	Mixed	Running	Running	Down
Feed and Product Flow Rates									
Export Power, MW-hr/day					14,808	13,740	5,532	5,532	6,112.8
Liquid Fuels, bpd					4,125	3,939	4,125	4,125	0
Sulfur, tpd					370.6	331.3	185.3	185.3	0
Natural Gas, Mscfd					535	8,830	17,623	17,623	46,416
Flux, tpd					109.7	98.1	54.85	54.85	0
Feed and Product Prices									
Power $/MW-hr	Fuels $/bbl	Sulfur $/ton	Gas $/MMBtu	Flux $/ton	Net Revenue, M$/day	Net Revenue, M$/day	Net Revenue, M$/day	Net Revenue, M$/day	Net Revenue, M$/day
20	30	30	2	5	429.4	384.8	204.4	204.4	29.4
27	30	30	2	5	533.1	480.9	243.2	243.2	72.2
30	30	30	2	5	577.5	522.2	259.7	259.7	90.6
35	30	30	2	5	651.5	590.9	287.4	287.4	121.1
40	30	30	2	5	725.6	659.6	315.1	315.1	151.7
45	30	30	2	5	799.6	728.3	342.7	342.7	182.2
50	30	30	2	5	873.6	797.0	370.4	370.4	212.8
20	30	30	2.6	5	429.1	379.5	193.9	193.9	1.6
27	30	30	2.6	5	532.7	475.6	232.6	232.6	44.4
30	30	30	2.6	5	577.2	516.9	249.2	249.2	62.7
35	30	30	2.6	5	651.2	585.6	276.8	276.8	93.3
40	30	30	2.6	5	725.2	654.3	304.5	304.5	123.8
45	30	30	2.6	5	799.3	723.0	332.2	332.2	154.4
50	30	30	2.6	5	873.3	791.7	359.8	359.8	185.0
20	30	30	3	5	428.9	375.9	186.8	186.8	-17.0
27	30	30	3	5	532.5	472.1	225.5	225.5	25.8
30	30	30	3	5	577.0	513.3	242.1	242.1	44.1
35	30	30	3	5	651.0	582.0	269.8	269.8	74.7
40	30	30	3	5	725.0	650.7	297.4	297.4	105.3
45	30	30	3	5	799.1	719.4	325.1	325.1	135.8
50	30	30	3	5	873.1	788.1	352.8	352.8	166.4
20	30	30	4	5	428.3	367.1	169.2	169.2	-63.4
27	30	30	4	5	532.0	463.3	207.9	207.9	-20.6
30	30	30	4	5	576.4	504.5	224.5	224.5	-2.3
35	30	30	4	5	650.5	573.2	252.2	252.2	28.3
40	30	30	4	5	724.5	641.9	279.8	279.8	58.8
45	30	30	4	5	798.5	710.6	307.5	307.5	89.4
50	30	30	4	5	872.6	779.3	335.1	335.1	120.0
20	30	30	5	5	427.8	358.3	151.6	151.6	-109.8
27	30	30	5	5	531.5	454.4	190.3	190.3	-67.0
30	30	30	5	5	575.9	495.7	206.9	206.9	-48.7
35	30	30	5	5	649.9	564.4	234.5	234.5	-18.1
40	30	30	5	5	724.0	633.1	262.2	262.2	12.4
45	30	30	5	5	798.0	701.8	289.9	289.9	43.0
50	30	30	5	5	872.0	770.5	317.5	317.5	73.6
20	30	30	6	5	427.3	349.4	133.9	133.9	-156.2
27	30	30	6	5	530.9	445.6	172.7	172.7	-113.5
30	30	30	6	5	575.3	486.8	189.3	189.3	-95.1
35	30	30	6	5	649.4	555.5	216.9	216.9	-64.5
40	30	30	6	5	723.4	624.2	244.6	244.6	-34.0
45	30	30	6	5	797.5	692.9	272.2	272.2	-3.4
50	30	30	6	5	871.5	761.6	299.9	299.9	27.1
20	30	30	7	5	426.7	340.6	116.3	116.3	-202.7
27	30	30	7	5	530.4	436.8	155.0	155.0	-159.9
30	30	30	7	5	574.8	478.0	171.6	171.6	-141.5
35	30	30	7	5	648.9	546.7	199.3	199.3	-111.0
40	30	30	7	5	722.9	615.4	227.0	227.0	-80.4
45	30	30	7	5	796.9	684.1	254.6	254.6	-49.8
50	30	30	7	5	871.0	752.8	282.3	282.3	-19.3

Table A-III
Daily Net Revenue Calculations for the Maximum Power Case Operating Scenario

MAXIMUM POWER CASE OPERATING SCENARIO	Case A Design	Annual Average	Case C	Case D	Case E
Equipment Status					
Number of Operating Gasifiers	2	Mixed	1	1	0
Number of Operating Combustion Turbines	2	Mixed	2	1	1
F-T Fuels Production	Running	Mixed	Down	Down	Down
Feed and Product Flow Rates					
Export Power, MW-hr/day	14,808	13,740	12,182	6,528	6,112.8
Liquid Fuels, bpd	4,125	3,939	0	0	0
Sulfur, tpd	370.6	331.3	185.3	185.3	0
Natural Gas, Mscfd	535	8,830	29,767	17,623	46,416
Flux, tpd	109.7	98.1	54.85	54.85	0

Feed and Product Prices: Power $/MW-hr	Fuels $/bbl	Sulfur $/ton	Gas $/MMBtu	Flux $/ton	Case A Design Net Revenue, M$/day	Annual Average Net Revenue, M$/day	Case C Net Revenue, M$/day	Case D Net Revenue, M$/day	Case E Net Revenue, M$/day
20	30	30	2	5	429.4	384.8	189.4	100.6	29.4
27	30	30	2	5	533.1	480.9	274.7	146.3	72.2
30	30	30	2	5	577.5	522.2	311.2	165.9	90.6
35	30	30	2	5	651.5	590.9	372.1	198.5	121.1
40	30	30	2	5	725.6	659.6	433.0	231.2	151.7
45	30	30	2	5	799.6	728.3	493.9	263.8	182.2
50	30	30	2	5	873.6	797.0	554.9	296.4	212.8
20	30	30	2.6	5	429.1	379.5	171.5	90.0	1.6
27	30	30	2.6	5	532.7	475.6	256.8	135.7	44.4
30	30	30	2.6	5	577.2	516.9	293.4	155.3	62.7
35	30	30	2.6	5	651.2	585.6	354.3	187.9	93.3
40	30	30	2.6	5	725.2	654.3	415.2	220.6	123.8
45	30	30	2.6	5	799.3	723.0	476.1	253.2	154.4
50	30	30	2.6	5	873.3	791.7	537.0	285.9	185.0
20	30	30	3	5	428.9	375.9	159.6	83.0	-17.0
27	30	30	3	5	532.5	472.1	244.9	128.7	25.8
30	30	30	3	5	577.0	513.3	281.4	148.3	44.1
35	30	30	3	5	651.0	582.0	342.4	180.9	74.7
40	30	30	3	5	725.0	650.7	403.3	213.5	105.3
45	30	30	3	5	799.1	719.4	464.2	246.2	135.8
50	30	30	3	5	873.1	788.1	525.1	278.8	166.4
20	30	30	4	5	428.3	367.1	129.9	65.4	-63.4
27	30	30	4	5	532.0	463.3	215.1	111.0	-20.6
30	30	30	4	5	576.4	504.5	251.7	130.6	-2.3
35	30	30	4	5	650.5	573.2	312.6	163.3	28.3
40	30	30	4	5	724.5	641.9	373.5	195.9	58.8
45	30	30	4	5	798.5	710.6	434.4	228.6	89.4
50	30	30	4	5	872.6	779.3	495.3	261.2	120.0
20	30	30	5	5	427.8	358.3	100.1	47.7	-109.8
27	30	30	5	5	531.5	454.4	185.4	93.4	-67.0
30	30	30	5	5	575.9	495.7	221.9	113.0	-48.7
35	30	30	5	5	649.9	564.4	282.8	145.6	-18.1
40	30	30	5	5	724.0	633.1	343.7	178.3	12.4
45	30	30	5	5	798.0	701.8	404.6	210.9	43.0
50	30	30	5	5	872.0	770.5	465.5	243.6	73.6
20	30	30	6	5	427.3	349.4	70.3	30.1	-156.2
27	30	30	6	5	530.9	445.6	155.6	75.8	-113.5
30	30	30	6	5	575.3	486.8	192.1	95.4	-95.1
35	30	30	6	5	649.4	555.5	253.1	128.0	-64.5
40	30	30	6	5	723.4	624.2	314.0	160.7	-34.0
45	30	30	6	5	797.5	692.9	374.9	193.3	-3.4
50	30	30	6	5	871.5	761.6	435.8	225.9	27.1
20	30	30	7	5	426.7	340.6	40.6	12.5	-202.7
27	30	30	7	5	530.4	436.8	125.8	58.2	-159.9
30	30	30	7	5	574.8	478.0	162.4	77.8	-141.5
35	30	30	7	5	648.9	546.7	223.3	110.4	-111.0
40	30	30	7	5	722.9	615.4	284.2	143.0	-80.4
45	30	30	7	5	796.9	684.1	345.1	175.7	-49.8
50	30	30	7	5	871.0	752.8	406.0	208.3	-19.3

= Best OVERALL case with only one gasifier operating (Cases C and D).

Appendix B

Subtask 2.2

Optimized Petroleum Coke Gasification

Power Plant with Liquids Coproduction

Subtask 2.2

Executive Summary

This report describes an optimized petroleum coke gasification power plant with liquid fuel precursors coproduction using Fischer-Tropsch technology. The plant consumes 5,417 tpd of coke (dry basis) and 111 tpd of flux to produce 367 MW of export power and 10,450 bpd of liquid fuel precursors. It also produces 373 tpd of elemental sulfur and 195 tpd of slag. The plant is located on the U.S. Gulf Coast adjacent to a petroleum refinery.

The Subtask 2.2 optimized design was developed from the Subtask 2.1 non-optimized design by enlarging and optimizing the F-T liquids production facilities. These changes increased the production of the F-T liquid fuel precursors from 4,125 bpd to 10,450 bpd at the expense of the production of export power, which was reduced from 617 MW to 367 MW. In contrast to the non-optimized Subtask 2.1 plant, the optimized Subtask 2.2 plant does not consume any natural gas during normal operations.

The Fischer-Tropsch hydrocarbon synthesis area basically consists of three sections; final sulfur removal, slurry-bed F-T reactor, and product recovery sections. The final sulfur removal section consists of three regenerable activated carbon beds in series. This is a much simpler and less costly design then that of the non-optimized plant design, which used a hydrolysis reactor followed by a non-regenerable ZnO adsorbent.

The sulfur-free syngas is fed to the slurry-bed F-T reactor which converts it to hydrocarbons over an iron-based catalyst. The heat of reaction is removed by generation of 440°F/375 psia steam. The lighter hydrocarbon products and unconverted syngas leave the reactor as vapors and are cooled by refrigeration to condense and recover the hydrocarbons as liquids. The unconverted syngas and non-condensable light hydrocarbons (primarily C1 through C3s) are compressed, moisturized, and sent to the power block. The heavier products are removed from the reactor as liquids, separated from the entrained catalyst, cooled, mixed with the lighter hydrocarbons, and sent to the adjacent petroleum refinery for separation, upgrading and incorporation into liquid transportation fuels.

The F-T liquid fuel precursors essentially are a bottomless, sulfur-free crude oil. Basically they are straight-chain 1-olefins and paraffins without any aromatics. The diesel fraction has a very high cetane number (>70) and is a premium blending component for diesel fuel. The naphtha fraction is a low octane material that requires further upgrading for use as a gasoline blending component. However, it is an excellent feedstock for an ethylene cracker.

The combined-cycle power block includes one GE7FAe+ combustion turbine, one heat recovery steam generator (HRSG), and a non-reheat steam turbine. The gross power output of the combined-cycle system is 474 MW (199 MW from the gas turbine and 275 MW from the steam turbine) resulting in 367 MW of net export power.

The Subtask 2.2 optimized plant has a LHV thermal efficiency of 55.1% and an HHV thermal efficiency of 56.7%, both of which are based on the heating value of the F-T liquids, the byproduct sulfur and the equivalent energy of the export power. These efficiencies are 7 to 9% greater than those of the Subtask 2.1 non-optimized plant, and about 12 to 14% greater than that of a coke IGCC power plant which was developed previously.

The Subtask 2.2 Optimized Coke Power Plant with Liquids Coproduction produces about 2½ times as much liquids and about 60% as much export power at a lower EPC cost (735 MM versus 818 MM mid-year 2000$ dollars) than the Subtask 2.1 non-optimized plant. When power prices are low and the liquid fuel precursors are worth 25 $/bbl or more, the Subtask 2.2 optimized plant has a substantially better return on investment than the Subtask 2.1 plant.

The improvements generated during the development of this Subtask 2.2 optimized design will be applied to the Subtask 2.3 design which will use coal rather than petroleum coke as the gasifier feedstock.

Subtask 2.2
Table of Contents

Attachment

Section 1

Introduction

The objective of this Gasification Plant Cost and Performance Optimization Project is to develop optimized engineering designs and costs for several Integrated Gasification Combined Cycle (IGCC) plant configurations. These optimized IGCC plant systems build on the commercial demonstration cost data and operational experience from the Wabash River Coal Gasification Repowering Project.[1] The Wabash River Repowering Project consists of a nominal 2,500 TPD E-GAS™ gasifier producing clean syngas for a GE 7A gas turbine and steam for repowering an existing steam turbine.

Task 1 of this IGCC Plant Cost and Performance Optimization study consists of the following nine subtasks:

- Subtask 1.1 – Expand the Wabash River Project facility design to a greenfield unit
- Subtask 1.2 – Petroleum Coke based IGCC plant with the coproduction of steam and hydrogen
- Subtask 1.3 – Optimized petroleum coke based IGCC plant with the coproduction of steam and hydrogen
- Subtask 1.4 – Optimized coal to power IGCC plant
- Subtask 1.5 – Comparison between single-train coal and coke fueled IGCC power plants
- Subtask 1.6 – Optimized coal fueled 1,000 MW IGCC power plant
- Subtask 1.7 – Optimized single-train coal to hydrogen plant
- Subtask 1.8 – Review the status of warm gas clean-up technology applicable to IGCC plants
- Subtask 1.9 – Discuss the Value Improving Practices availability and reliability optimization program

Task 1 has been completed. The Task 1 Topical Report was issued to the Department of Energy on May 30, 2002.[2]

Task 2 has the objectives of developing optimum plant configurations for IGCC power plants with the coproduction of liquid fuel precursors. Task 2 is divided into the three subtasks.

Subtask 2.1 – [Non-Optimum] Petroleum Coke Gasification Power Plant with Liquids Coproduction Starting with the same petroleum coke and gasification plant designs generated in the Subtask 1.3 Next Optimized Coke IGCC Coproduction Plant, a design shall be developed for a coke gasification power plant co-producing liquid transportation fuel precursors containing a single-train, once-through Fischer-Tropsch (F-T) gas-to-liquids (GTL) plant. The liquid hydrocarbons from the F-T hydrocarbon synthesis section will be recovered and sent to the adjacent petroleum refinery for upgrading and blending into premium liquid transportation fuels. The unconverted syngas and non-condensable hydrocarbons from the F-T hydrocarbon synthesis section will be used for power production in the combined-cycle power block.

[1] "Wabash River Coal Gasification Repowering Project, Final Technical Report", U. S. Department of Energy, Contract Agreement DE-FC21-92MC29310, August 2000.
[2] "Task 1 Topical Report – IGCC Plant Cost Optimization", Gasification Plant Cost and Performance Optimization, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, May 30, 2002.

Subtask 2.2 – Optimum Petroleum Coke Gasification Power Plant with Liquids Coproduction The Subtask 2.1 plant shall be optimized to develop an optimized coke gasification plant co-producing liquid transportation fuel precursors. Optimization activities primarily will be concerned with the F-T area and overall plant integration. Since the Subtask 2.1 gasification area was developed from an optimized IGCC petroleum coke gasification coproduction plant, a review of the plant is appropriate at this time to ensure that the previous modifications are still applicable to this case.

Subtask 2.3 - Optimum Coal Gasification Power Plant with Liquids Coproduction The Subtask 2.2 plant shall be converted to a coal-fueled gasification unit using Illinois No. 6 coal, retaining the optimized portions and incorporating those optimizations developed in Subtask 1.6. This will involve combining the optimized coal gasification plant developed in Subtask 1.6 with the Subtask 2.2 plant, to develop an optimized coal gasification power plant co-producing liquid transportation fuel precursors. Because of differences in the syngas generation area and resulting syngas composition, the F-T hydrocarbon synthesis area, and overall plant integration probably will require modification.

This report describes the design, performance and economics of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. Section 2 provides background information on two previous studies that are the basis for this current study. Section 2.1 describes the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. Section 2.2 briefly describes the F-T hydrocarbon synthesis reaction and provides an overview of the indirect coal liquefaction plant that was developed during the previous Department of Energy study.

Section 3 described the Subtask 2.2 Optimized Petroleum Coke Gasification Power Plant with Liquids Coproduction that was developed in this study and compares it with the Subtask 2.1 non-optimized plant. Section 4 provides an availability analysis. Section 5 contains a financial analysis of plant performance and compares it with the previously developed Subtask 2.1 plant and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. Section 6 contains a summary of this subtask.

An Attachment contains a listing of the major equipment in the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction.

Section 2

Background

During Task 1, several designs were developed for petroleum coke IGCC coproduction plants that supplied an adjacent petroleum refinery with 750ºF/750 psig steam and hydrogen.[2] Subtask 1.2 developed a first pass design and cost estimate for a petroleum coke IGCC coproduction plant located on the Gulf Coast based on the Subtask 1.1 Wabash River Greenfield Plant. Subtask 1.3 developed designs for three optimized petroleum coke IGCC coproduction plants, each with the same design capacity. The only difference between these plants was the amount of spare equipment inside the gasification block. The Base Case design contains two gasification trains with each gasification train having a spare gasification reactor vessel that can be placed in service when the other reactor requires refractory replacement. The minimum cost case eliminated the spare gasification vessel in each train. The spare gasification train case has three complete gasification trains beginning with the slurry feed pumps and continuing through the syngas particulate removal systems. The remainder of the facility is sized so that only two gasification trains can operate simultaneously at design capacity. The Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant is based on the Subtask 1.3 spare train case because that case has the highest return on investment. Section 2.1 describes the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant.

In 1991, Bechtel with Amoco as the main subcontractor was awarded DOE contract DE-AC22-91PC90027 to develop designs and computer process simulation models for indirect coal liquefaction plants using advanced Fischer-Tropsch Technology.[3] Subsequently, the simulation model was improved by adding additional components.[4] The Fischer-Tropsch (F-T) hydrocarbon synthesis section of this ASPEN process simulation model was used to develop the design of the F-T hydrocarbon synthesis section of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction. Section 2.2 briefly describes the F-T hydrocarbon synthesis reaction and presents an overview of the entire facility.

2.1 Subtask 2.1 Next Optimized Petroleum Coke IGCC Coproduction Plant

The Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant was developed by applying nine Value Improving Practices (VIPs) to the Subtask 2.1 plant to reduce costs and improve operability.[5] As a result of this effort, plant performance was improved, the plant cost was reduced, and the return on investment was significantly improved. The results of this VIP and optimization study included:

[3] "Topical Report – Volume I, Process Design – Illinois No. 6 Coal Case with Conventional Refining", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.
"Topical Report – Volume IV, Process Flowsheet (PFS) Models", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.
[4] "Topical Report VI – Natural Gas Fischer-Tropsch Case, Volume II, Plant Design and Aspen Process Simulation Model", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, August, 1996.
[5] "Task 1 Topical Report – IGCC Plant Cost Optimization", Gasification Plant Cost and Performance Optimization, Chapter II, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, May 30, 2002

- Simplified solids handling system
- Removal of the feed heaters and spare pumps
- Maximum use of slurry quench
- Maximum syngas moisturization
- Use of a cyclone and a dry particulate removal system to clean the syngas
- Removal of the T-120 post reactor residence vessel
- Simplified Claus plant, amine and sour water stripper
- Use of state-of-the-art GE 7FA+e gas turbines with 210 MW output and lower NOx
- Use of steam diluent in the gas turbines
- Development of a compact plant layout to minimize the use of large bore piping
- Used Bechtel's advanced construction techniques to reduce costs
- Added design features to reduce O&M costs and increase syngas availability

Table II.1 shows the design input and output streams for the Subtask 1.3 Next Plant. The plant processes 5,417 tpd of dry petroleum coke and produces 474 MW of export power. In addition, the plant exports 980,000 lb/hr of 750ºF/700 psia steam and 80 MMscfd of hydrogen to the adjacent petroleum refinery. It also produces 373.4 tpd of sulfur and 195.1 tpd of slag. No natural gas is consumed during design operations. However, the plant does use natural gas during startup and as a supplementary fuel to fire the combustion turbines when insufficient syngas is available.

The resulting design configuration for the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant is shown in Figure 2.1. The plant basically is a two train facility, but with three gasification trains, two operating and one spare. The two-train sections of the plant are sized so that they only have sufficient capacity to process the output from two gasifiers simultaneously operating at design capacity.

Table II.1

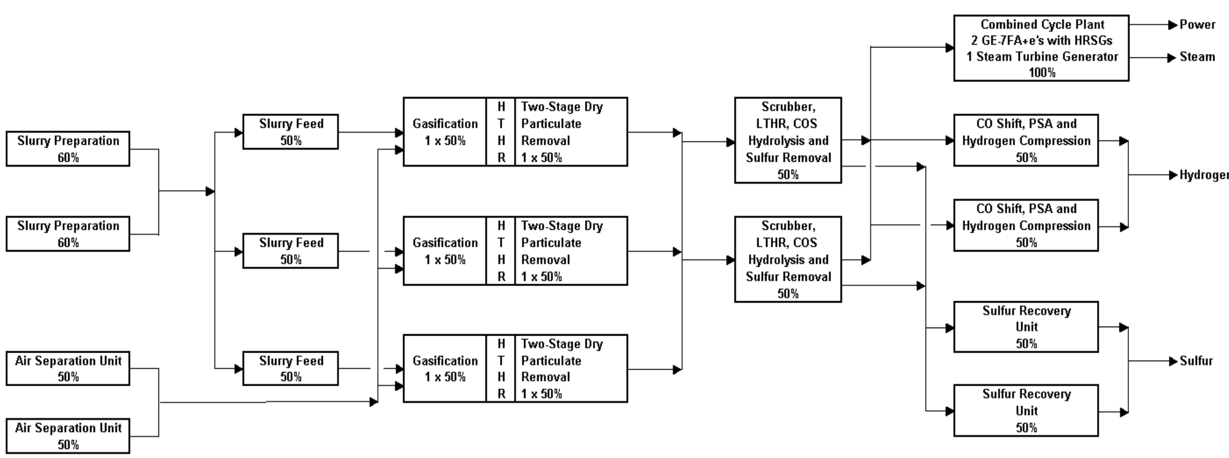
Design Input and Output Streams for the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

	Subtask 1.3 Next Plant
Plant Inputs	
Coke Feed, as received TPD	5,692
Dry Coke Feed to Gasifiers, TPD	5,417
Oxygen Production, TPD of 95% O_2	5,954
Total Fresh Water Consumption, gpm	5,223
Condensate Return from the Refinery, lb/hr	686,000
Flux, TPD	110.6
Natural Gas, MMBtu/hr	0
Plant Outputs	
Net Power Output, MW	474.0
Sulfur, TPD	373.4
Slag, TPD (15% moisture)	195.1
Hydrogen, MMscfd	80
HP Steam, 750°F/700 psia	980,000
Fuel Gas Export, MMBtu/hr	0

Figure 2.1

Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

Simplified Block Train Diagram



Notes: Capacity percentages are based on total plant capacity.

2.2 Fischer-Tropsch Hydrocarbon Synthesis Process

The Fischer-Tropsch hydrocarbon synthesis process is an old process in which synthesis gas or syngas (carbon monoxide and hydrogen) react over a catalyst to produce aliphatic hydrocarbons (principally normal paraffins and straight chain 1-olefins). It was used by Germany during the Second World War to make liquid fuels for military use. Subsequent cost reductions may have made F-T processes competitive in certain situations. Currently, there is a lot of interest in using the F-T process to monetize remote natural gas by converting it into an easily transportable synthetic crude oil that can be upgraded to liquid transportation fuels.

In general, the F-T hydrocarbon synthesis reactions for olefins and normal paraffins can be written as

$$n\ CO + 2n\ H_2 \rightarrow C_nH_{2n} + n\ H_2O$$

$$n\ CO + (2n+2)\ H_2 \rightarrow H\text{-}(CH_2)_n\text{-}H + n\ H_2O$$

As seen from the above reaction stoichiometry, the ideal syngas composition is just over 2 moles of hydrogen for each mole of carbon monoxide.

The reaction is very exothermic. Traditionally, at a large scale the reaction has been performed over solid catalyst that is placed in small diameter tubes immersed in a cooling medium (such as boiling water) to remove the heat of reaction. The hydrocarbon product yield distribution can be characterized by a Schultz-Flory distribution in which the molar ratio of a component containing n carbon atoms to one with n+1 carbon atoms is a constant called alpha (α). As the reaction temperature increases, the yield distribution shifts to lighter hydrocarbons; i.e., the α parameter gets smaller. As time has progressed, more sophisticated mathematical yield models using multiple α parameters have been developed to represent the F-T reaction yields.

In the 1950s, the slurry-bed reactor was developed in which fine catalyst particles are suspended in a liquid, and the reactant syngas is bubbled up through the catalyst/liquid mixture. Steam is generated within cooling coils immersed in the slurry-bed to remove the heat of reaction. This system has a high heat transfer rate resulting in a cheaper reactor with a higher productivity rate than catalyst particles packed in tubes. The lighter hydrocarbon products and unconverted syngas are withdrawn as vapor from the top of the reactor. Slurry is withdrawn from the reactor and pumped through a hydroclone and filter system which separates the clarified liquid products from the catalyst. The concentrated catalyst/slurry stream is returned to the reactor. A constant (steady-state) catalyst activity is maintained by continually withdrawing a small portion of catalyst from the reactor and replacing it with fresh catalyst.

Iron-based and promoted cobalt-based catalysts are the two primary catalysts currently used for F-T synthesis. Iron-based catalysts promote the water gas shift reaction which produces hydrogen from carbon monoxide and water; whereas cobalt catalysts generally do not. Therefore, for a syngas with a low hydrogen to carbon molar ratio, an iron based catalyst is preferred because it will produce hydrogen within the slurry-bed F-T synthesis reactor; whereas with a cobalt based catalyst, additional hydrogen has to be produced externally to the F-T synthesis reactor.

In the early 1990s, Bechtel developed several designs for indirect coal liquefaction plants using Fischer-Tropsch technology (references 3 and 4). Table II.2 shows the major input and output streams for the Baseline plant. The plant consumes 20,323 tpd of ROM Illinois No. 6 coal (8.6

wt% water) and 3,119 bpsd of normal butane to produce at total of 50,491 bpsd of petroleum products (1,921 bpsd of C3 LPG, 23,915 bpsd of gasoline, and 24,655 bpsd of distillate fuels). The plant is divided into three processing areas.

Area 100	Clean Syngas Production Area
Area 200	Fischer-Tropsch Synthesis Loop
Area 300	Product Upgrading and Refining Area

The Area 100 Clean Syngas Production Area grinds and dries the coal, gasifies the coal in six Shell gasifiers (five operating and one spare), scrubs and cleans the syngas, and recovers 46.69 Mlb/hr of sulfur for sale.

The Area 200 Fischer-Tropsch Synthesis Loop obtains a high CO conversion by recycling the unconverted syngas after recovering hydrogen and removing CO_2. The F-T hydrocarbon synthesis section contains 25 slurry-bed reactors (24 operating and one spare) arranged in eight parallel trains with each train having three reactors in parallel.

The Area 300 Product Upgrading and Refining Area essentially is a small refinery that upgrades the F-T products into liquid transportation fuels. It contains a saturated gas plant, C3/C4/C5 alkylation unit, C4 isomerization unit, C5/C6 isomerization unit, catalytic reformer, naphtha hydrotreater, distillate hydrotreater, and a wax hydrocracker. To increase the gasoline yield, normal butanes are purchased, isomerized to isobutene, and used to alkylate the C3, C4 and C5 olefins to make a high-octane gasoline blending component.

Table II.2

Design Input and Output Streams for the Baseline Indirect Coal Liquefaction Plant

Plant Inputs				
Illinois No. 6 ROM Coal*	1,693.6	Mlb/hr	20,323	TPD
Electric Power	54.36	MW		
Normal Butane	26.50	Mlb/hr	3,119	bpsd
Raw Water	10,042	gpm		
Plant Outputs				
C3 LPG	14.22	Mlb/hr	1,921	bpsd
Gasoline	251.44	Mlb/hr	23,915	bpsd
Diesel	278.21	Mlb/hr	24,655	bpsd
Sulfur	46.69	Mlb/hr		
Slag	187.03	Mlb/hr		

* As received coal containing 8.6 wt% water

2.3 Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction

The Subtask 2.1 [non-optimized] Coke Gasification Power Plant with Liquids Coproduction was developed from the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant which produces 80 MMscfd of hydrogen and 980,000 lb/hr of 750ºF/700 psig steam for the adjacent petroleum refinery. This plant has been described previously.[6] Therefore, this section only provides an overview of the entire facility. However, the F-T area is described in more detail to provide a basis for comparison with the subsequently developed optimized plant design.

Starting from the Subtask 1.3 Next Plant, the Subtask 2.1 plant was developed by eliminating the export steam production and hydrogen production facilities and replacing them with a single-train, once-through Fischer-Tropsch hydrocarbon synthesis plant. The energy that was used to produce the export steam now is used to generate additional power. Even with almost the same coke feed rate to the gasifiers, the process changes required adjustments to the steam and water flows both in and between the gasification block and the power generation block in order to effectively balance the systems.

Table II.3 compares the design input and output stream flows for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. The Subtask 2.1 plant processes slightly less petroleum coke (5,376 vs. 5,417 dry tpd) than the Subtask 1.3 Next Plant. It also has a higher fresh water consumption of 6,472 gpm vs. 5,223 gpm. Furthermore, it consumes a small amount of natural gas, 23.2 MMBtu HHV/hr. Because the Subtask 2.1 plant does not export any hydrogen or steam, it produces more export power than the previous case (617 MW vs. 474 MW) in addition to 4,125 bpd of liquid fuel precursors from the F-T area.

On a lower heating value (LHV) basis, the plant has a thermal efficiency 47.8% when the heating value of the byproduct sulfur is included and 45.9% when the byproduct sulfur is not included. On a higher heating value (HHV) basis, the plant has a thermal efficiency 47.9% when the heating value of the byproduct sulfur is included and 46.0% when the byproduct sulfur is not included. These thermal efficiencies are higher than those that would be obtained from a coke IGCC power plant of a similar design because it includes the heating value of the liquid fuel that is produced. Since the second law of thermodynamics states this liquid fuel cannot be used at a 100% thermal efficiency, the thermal efficiency of the plant will be somewhat lower when the final disposition of the liquid fuel is considered.

Figure 2.2 shows the train configuration of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction. The plant basically is a two train facility. However, there are three gasification trains, two operating and one spare, and only a single train F-T hydrocarbon synthesis section. The two-train sections of the plant are sized so that they only have sufficient capacity to process the output from two gasifiers simultaneously operating at design capacity.

Figure 2.3 is a simplified block flow diagram of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction. This plant can be considered to consist of three distinct main processing areas and a balance of plant area (Area 900).

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[6] Task 2 Progress Report – Subtask 2.1, A Coke Gasification Power Plant with Liquid Fuels Coproduction," Gasification Plant Costs and Performance Optimization, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, Draft Report of February, 2003.

- The gasification island and air separation unit (Areas 100, 150, 250, 300, 350, 400, 420, and 800)
- The F-T hydrocarbon synthesis area (Areas 200 and 201)
- The power block (Areas 500 and 600)

The Fischer-Tropsch hydrocarbon production area consists of two plants; Area 200, the Final Syngas Cleanup Area, and Area 201, the Fischer-Tropsch Slurry-bed Reactor Area.

2.3.1 Area 200, Final Syngas Cleanup Area

The Final Syngas Cleanup Area, Area 200, reduces the sulfur concentration of the cleaned syngas from the acid gas removal area of the gasification block to less than 0.1 ppm of sulfur. This is accomplished by hydrolyzing the small residual amounts of carbonyl sulfide (COS) and trace amounts of other light organic sulfur compounds (such as CS_2) to hydrogen sulfide (H_2S), and removing the H_2S by reacting it with zinc oxide (ZnO) to produce solid zinc sulfide (ZnS) and water. The ZnO is permanently consumed, and the ZnS is discarded.

Süd-Chemie’s G-41P RS hydrolysis catalyst is used at a 300ºF operating temperature to hydrolyze the COS to H_2S and H_2O. This is a potassium chromate on aluminum oxide catalyst and is provided in 1/8 inch extrudates. At the design volumetric hourly space velocity of 3,000 vol/vol-hr, the expected catalyst life is greater than 60 months.

Süd-Chemie’s G-72E ZnO catalyst/sulfur adsorbent is used to capture the sulfur and reduce the residual syngas sulfur content at 650ºF. In order to provide continuous H_2S removal, the process design uses a two bed reactor configuration with the two beds in series. Necessary piping is provided so that these two beds can be switched, and the spent adsorbent can be replaced without any interruption of service. When H_2S breakthrough occurs in the first bed (lead bed), it is taken out of service for adsorbent replacement, and the other bed (lag bed) is in service alone. After the adsorbent has been replaced, the bed with the freshly loaded adsorbent is put back in service as the lag bed. The two bed in series operation continues until H_2S breakthrough occurs in the other bed, and it is taken out of service for adsorbent replacement. The operating cycle repeats. Each catalyst bed is sized for a six month cycle length.

Figure 2.4 contains a schematic flow diagram of the Final Syngas Cleanup Area, Area 200, and the F-T slurry-bed reactor and product recovery area, Area 201. The cleaned syngas from the gasification block is preheated to 292ºF in heat exchanger 201E-1 with hotter sulfur-free syngas from exchanger 200E-2. The preheated syngas leaving the 200E-1 heat exchanger is mixed with 440ºF/375 psia stream that was generated in the slurry-bed F-T reactor and fed to the 200R-1 COS Hydrolysis Reactor where the following chemical reaction converts the COS to H_2S.

$$COS + H_2O \rightarrow H_2S + CO_2$$

The syngas leaving the 200R-1 reactor is heated to 520ºF in exchanger 200E-2 with the hot sulfur-free syngas leaving the 200R-2 and/or 200R-3 ZnO sulfur adsorbers. The syngas then is heated to 650ºF in the 200F-1 furnace. The furnace fuel is a mixture of natural gas and the low-pressure fuel gas recovered from the F-T reaction products. This furnace is oversized for startup.

The heated syngas then enters the ZnO sulfur adsorption beds, 200R-2 and 200R-3. Although it is not shown in the drawing, these two beds are arranged in a lead-lag configuration so that one bed may be taken off line for ZnO replacement while the other remains in service.

The sulfur-free syngas leaving the ZnO beds is cooled to 257ºF by preheating the entering cleaned syngas in the 200E-1 and 200E-2 heat exchangers. This sulfur-free syngas stream is the feed stream to the 201R-1 slurry-bed F-T hydrocarbon synthesis reactor.

2.3.2 Fischer-Tropsch Slurry-bed Reactor Area

The Fischer-Tropsch slurry-bed reactor converts the sulfur-free syngas primarily into olefinic hydrocarbons by the reaction

$$n\,CO + 2n\,H_2 \rightarrow C_nH_{2n} + n\,H_2O$$

The reaction is promoted by an iron-based catalyst which also promotes the water-gas shift reaction

$$CO + H_2O \rightarrow CO_2 + H_2$$

The lighter hydrocarbon products leave the slurry-bed reactor in the vapor phase, are cooled and the condensed liquid collected. The heavier hydrocarbons are removed as liquids from the reactor, separated from the suspended catalyst, cooled, and combined with the lighter products to form the liquid fuel precursors product.

In order to maintain a constant catalyst activity, there is a continual addition of fresh catalyst and a continual withdrawal of used catalyst from the slurry-bed catalyst. The fresh catalyst must be pretreated in a reducing atmosphere at an elevated temperature to activate it. The catalyst pretreating system consists of a similar vessel to the slurry-bed reactor, but without the internal cooling facilities.

As shown in Figure 2.4, the cooled sulfur-free syngas stream from the zinc oxide sulfur adsorption beds is mixed with 440ºF/375 psia steam before entering the slurry-bed F-T hydrocarbon synthesis reactor, 201R-1, where the hydrogen and carbon monoxide are converted to straight chain aliphatic hydrocarbons, carbon dioxide and water. The heat of reaction is removed from the slurry-bed F-T reactor by the generation of 440°F/375 psia steam inside tubes located within the slurry-bed reactor. Pump 201P-1 circulates boiler feed water (BFW) between the 201C-1 steam drum and the 201R-1 reactor to ensure that sufficient BFW always is flowing through the cooling tubes.

Cyclone 201T-1 removes entrained catalyst particles from the vapor stream leaving the top of the F-T reactor. The vapor stream then is cooled to 110°F in two exchangers, 201E-1 and 201E-2. The first exchanger (201E-1) cools the syngas to 130ºF by heating BFW, and the second exchanger (201E-2) cools the syngas to 110ºF with cooling water. The cooled syngas leaving the second exchanger enters the 201C-2 reactor overhead flash drum. The sour water from the boot of 201C-2 goes to the 201C-4 sour water flash drum. The vapor stream leaving the sour water flash drum goes to the incinerator, and the sour water is recycled to the gasifier.

The vapor stream from the reactor overhead vapor flash drum is washed in 201C-3 to remove any residual catalyst particles and heated from 110°F to 120°F in exchanger 201E-3 to prevent condensation of the heavy components during compression. Condensing 440°F/375 psia steam, which was generated in the slurry-bed F-T reactor, is the heating medium. The heated vapor stream is compressed to 380 psia in 201K-1 to produce a high-pressure fuel gas stream which is sent to the power block where it is mixed with the syngas and steam before entering the combustion turbine. This high-pressure fuel gas stream consists of unconverted syngas (carbon monoxide and hydrogen) and light hydrocarbons (primarily C1 through C3s) that are produced in the F-T reactor.

The liquid hydrocarbon stream leaving 201C-2 is mixed with the cooled liquid hydrocarbons from the slurry-bed F-T reactor and sent for upgrading into liquid transportation fuels.

The liquid stream leaving the slurry-bed F-T reactor passes through hydroclone 201T-2 to remove a majority of the entrained catalyst particles. The catalyst-rich hydroclone bottoms goes to mixing tank 201C-10 from which most of it is returned to the slurry-bed reactor by pump 201P-3. A portion of the hydroclone bottoms is withdrawn and sent to the catalyst withdrawal system shown in Figure 3.3. Residual catalyst particles are removed from the hydroclone overhead stream in the 201T-3 filter system.

The catalyst-free liquid leaving the filter system is reduced in pressure and flashed in drum 201C-5. The vapor stream is further cooled to 110°F in exchanger 201E-4 with cooling water and flashed in drum 201C-6. The vapor stream from drum 201C-6 is a low-pressure fuel gas which is used as fuel in the 200F-1 furnace.

The liquid leaving the 201C-5 flash drum is cooled to 200ºF in 201E-5 by preheating boiler feed water. The cooled liquid from 201C-5 is mixed with the liquid stream from the 201C-6 flash drum in drum 201C-9 and a cooled liquid recycle stream from 201C-8. This mixture now is cooled to 110ºF by cooling water in exchanger 201E-6 and sent to the 201C-8 liquid fuel flash drum along with the liquid from the 201C-2 reactor overhead vapor flash drum. The vapor leaving the 201C-8 liquid fuel flash drum is mixed with the vapor from the 201C-6 flash drum and is used as low-pressure fuel gas in the 200F-1 furnace.

The liquid from the 201C-8 flash drum is split into two streams. One of the liquid streams is recycled back to 201C-9 flash drum via pump 201P-1 to dilute the heavier hydrocarbons in order to control their viscosity as they are cooled in exchanger 201E-6. The other liquid stream is the liquid fuel precursors product which is sent to the adjacent petroleum refinery for upgrading into liquid transportation fuels (gasoline, diesel, etc.).

Figure 2.5 shows the catalyst withdrawal system. The hot catalyst-rich stream from the 201C-10 drum is cooled in exchanger 201E-7 and pumped by pump 201P-4 through the 201T-4 filters to remove the used catalyst which is collected and discarded. The catalyst free liquid is mixed with the liquid fuel precursors product stream from drum 201C-6 and sent to the adjacent petroleum refinery for upgrading.

The catalyst pretreatment system also is shown in Figure 2.5. The makeup catalyst is fed into the 201C-11 catalyst pretreater where it is combined with heated liquid product from storage. Recycle gas is circulated through the pretreater vessel via compressor 201K-2, exchanger 201E-9, and furnace 201F-1. Vapors leaving the pretreater vessel are cooled in exchangers 201E-9 and 201E-10 before being flashed in drum 201C-13. A portion of the vapor from 201C-13 is withdrawn and sent to the incinerator to remove inerts from the system. However, most of the vapors from 201C-13 are recycled to the pretreater after addition of some fresh syngas or hydrogen via the 201K-2 compressor, 201E-9 exchanger, and 201F-1 furnace. Pretreated catalyst is withdrawn from the pretreater vessel and stored in the heated 201C-12 mixing tank until it is injected into the slurry-bed F-T reactors via pump 201P-8.

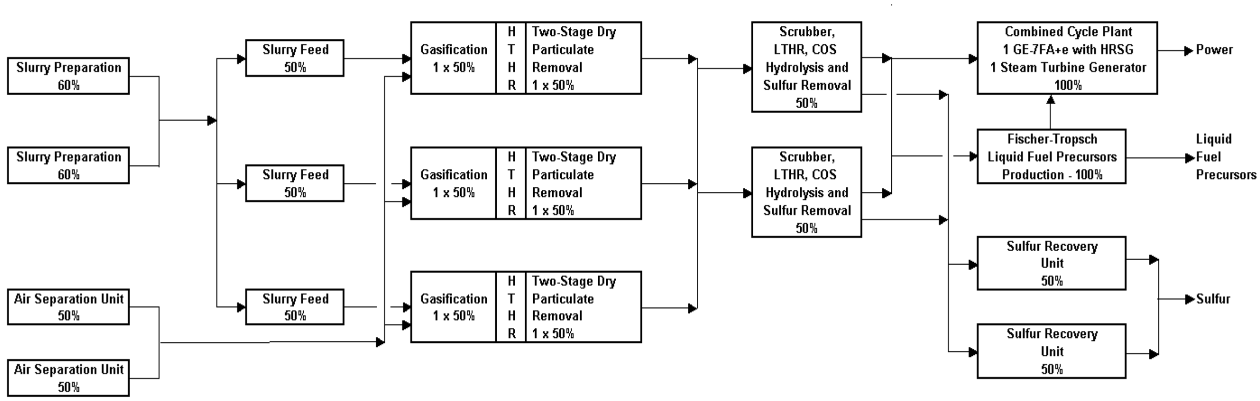
Table II.3

Design Input and Output Streams for the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

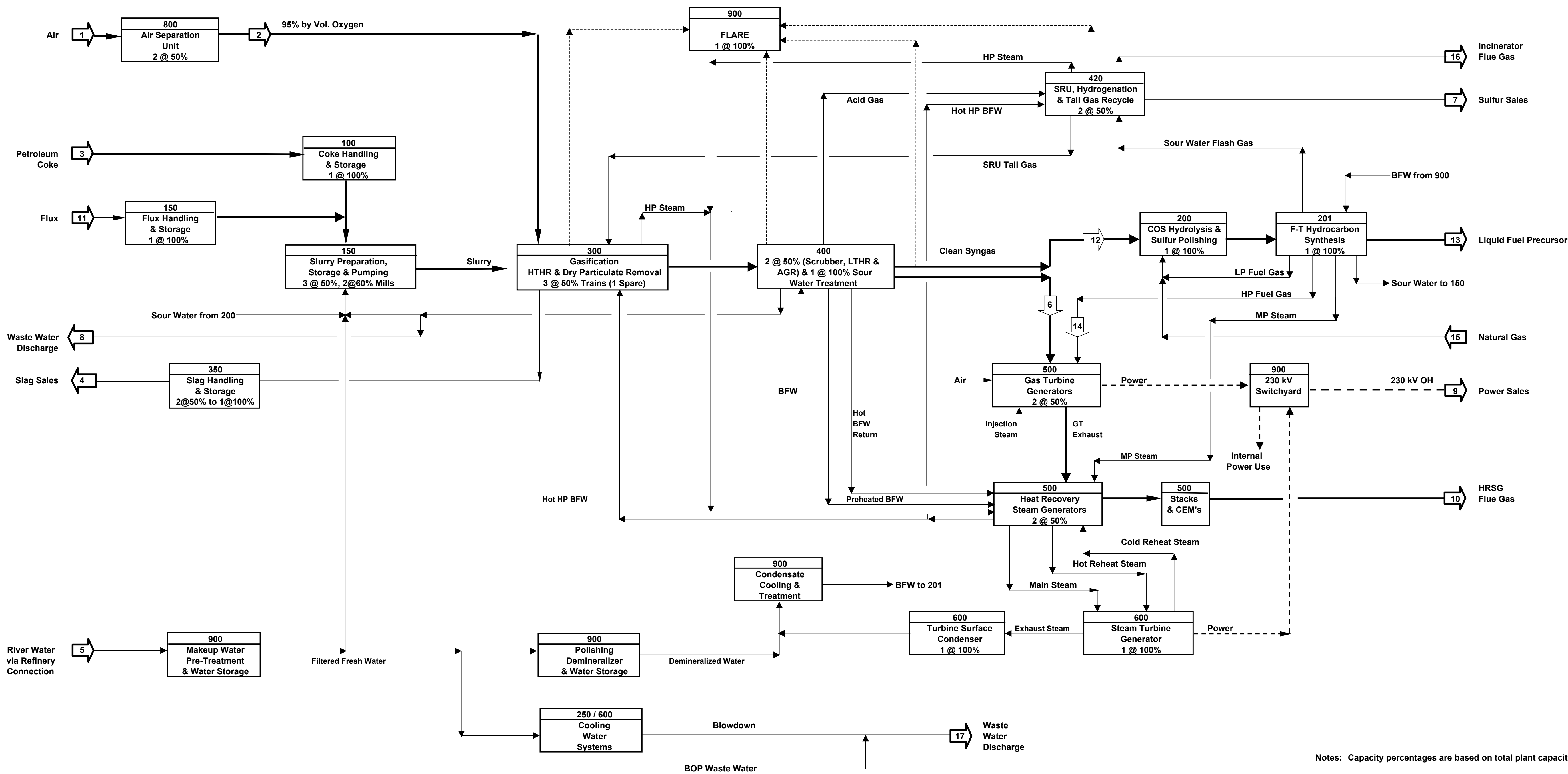
	Subtask 1.3 Next Plant	Subtask 2.1 Power and Liquids Plant
Plant Inputs		
Coke Feed, as received TPD	5,692	5,649
Dry Coke Feed to Gasifiers, TPD	5,417	5,376
Oxygen Production, TPD of 95% O_2	5,954	5,919
Total Fresh Water Consumption, gpm	5,223	6,472
Condensate Return from the Refinery, lb/hr	686,000	0
Flux, TPD	110.6	109.7
Natural Gas, MMBtu/hr	0	23.2
Plant Outputs		
Net Power Output, MW	474.0	617.0
Sulfur, TPD	373.4	370.6
Slag, TPD (15% moisture)	195.1	193.6
Hydrogen, MMscfd	80	0
HP Steam, 750°F/700 psia	980,000	0
Liquid Fuel Precursors, bpd	0	4,125

Figure 2.2

Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction Simplified Block Train Diagram



Notes: Capacity percentages are based on total plant capacity.



Notes: Capacity percentages are based on total plant capacity.

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21
	Air	Oxygen	Coke	Slag	Water	Syngas	Sulfur	Water	Power	Flue Gas	Flux	Syngas	Liq Fuel	Fuel Gas	Nat Gas	Flue Gas	Water				
Flow	25,808 Tons/Day	5,919 Tons/Day	5,376 Tons/Day	193.6 Tons/Day	3,236,000 Lb/Hr	722,540 Lb/Hr	370.6 Tons/Day	52,000 Lb/Hr	617,000 kWe	7,966,800 Lb/Hr	109.7 Tons/Day	403,502 Lb/Hr	48,897 Lb/hr	370,255 Lb/hr	23.2 MMBtu/hr	21,672 Lb/Hr	711,000 Lb/Hr				
Nominal Pressure - psig	Atmos.	609	NA	NA	50	355	NA	62	NA	Atmos.	NA	360	50	365	50	Atmos.	Atmos.				
Temperature - F	70	240	Ambient	180	70	532	332	80	NA	265	NA	100	100	168	100	500	71				
HHV Btu/lb	NA	NA	14,848	NA	NA	4,325	3,983	NA	NA	NA	NA	4,997	19,777	1,852	1,000	NA	NA				
LHV Btu/lb	NA	NA	14,548	NA	NA	4,101	3,983	NA	NA	NA	NA	4,738	18,297	1,698	910	NA	NA				
Energy - MM HHV/hr	NA	NA	6,652	NA	NA	3,125	123	NA	NA	NA	NA	2,016	967.0	685.6	23.2	NA	NA				
Energy - MM LHV/hr	NA	NA	6,518	NA	NA	2,963	123	NA	NA	NA	NA	1,912	894.7	628.8	21.1	NA	NA				
Notes	Dry Basis	5,582 O2	Dry Basis	15%Wtr.	6,472 GPM	To GT	Sales	104 GPM	230 kV			For F-T	4,125 bpd				1,422 GPM				

DOE Gasification Plant Cost and Performance Optimization

Figure 2.3

Subtask 2.1

NON-OPTIMIZED COKE GASIFICATION POWER PLANT WITH LIQUIDS COPRODUCTION

BLOCK FLOW DIAGRAM

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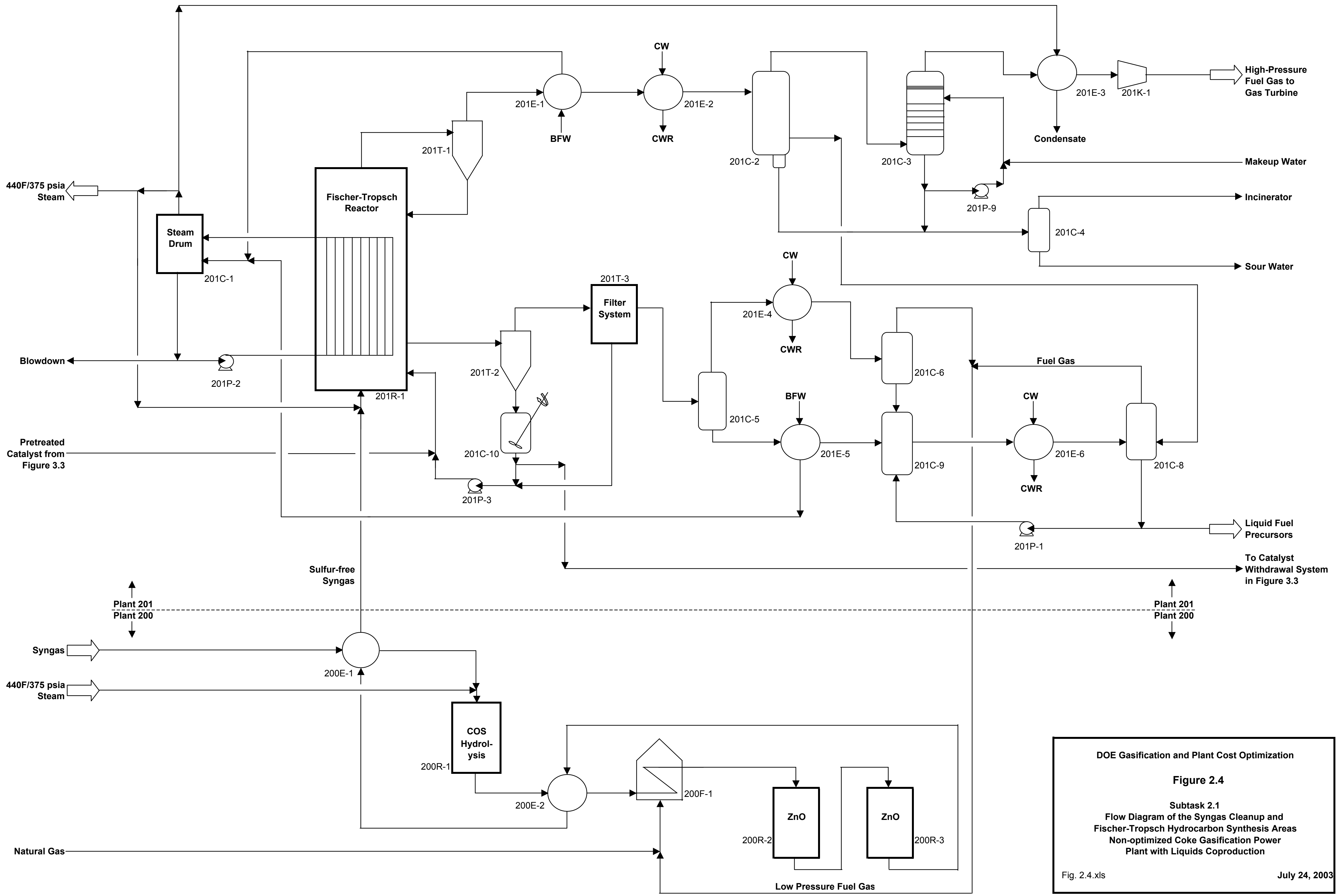
July 24, 2003



DOE Gasification and Plant Cost Optimization

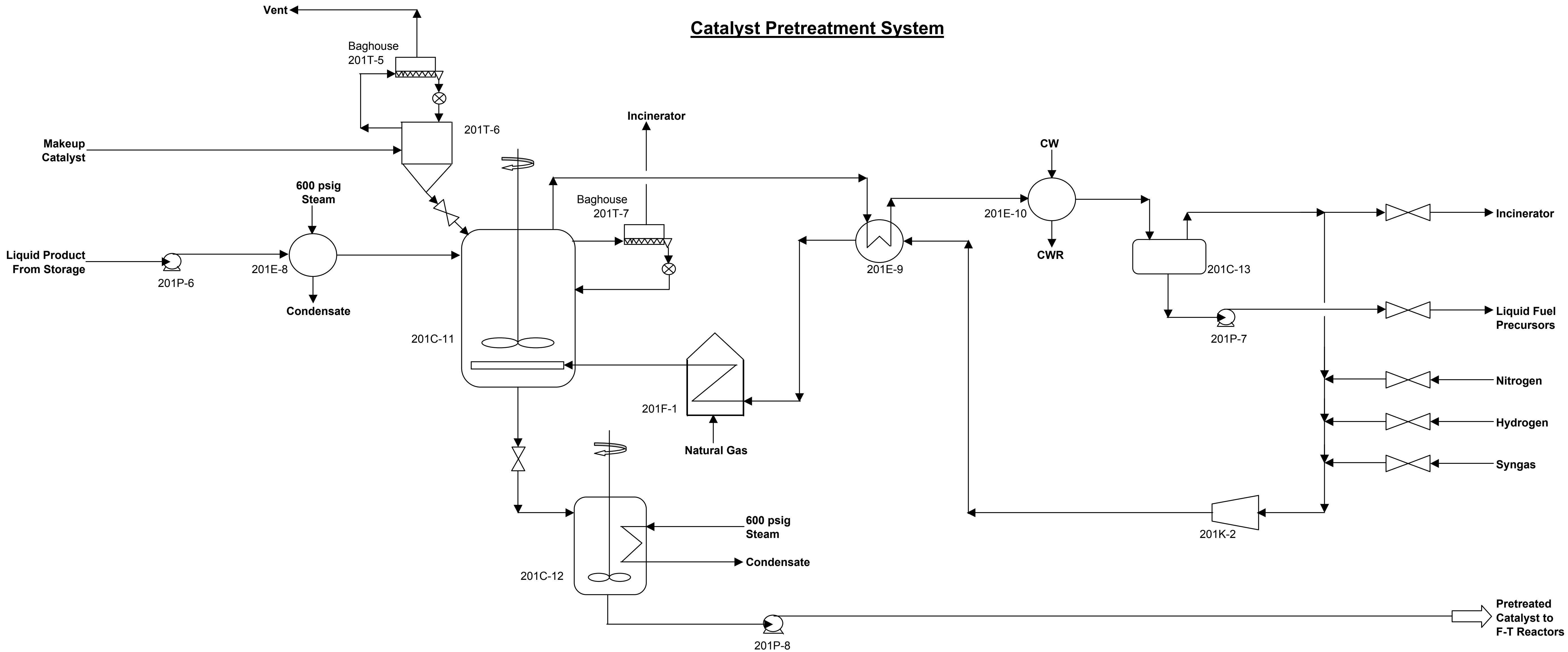
Figure 2.4

Subtask 2.1
Flow Diagram of the Syngas Cleanup and
Fischer-Tropsch Hydrocarbon Synthesis Areas
Non-optimized Coke Gasification Power
Plant with Liquids Coproduction

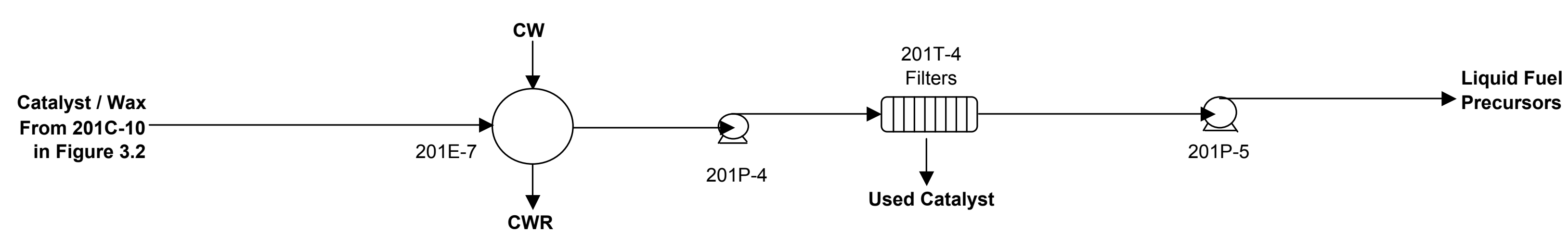
Fig. 2.4.xls

July 24, 2003

Catalyst Pretreatment System



Catalyst Withdrawal System



DOE Gasification and Plant Cost Optimization

Figure 2.5

Subtask 2.1
Flow Diagram of the Catalyst Pretreating and Catalyst Removal Areas of Plant 201
Non-optimized Cole Gasificatiopn Power Plant with Liquid Fuels Coproduction

Fig. 2.5.xls **December 4, 2002**

Section 3

Description of the Subtask 2.2 Optimized Coke Gasification Power Plant With Liquids Coproduction

The Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction was developed from the Subtask 2.1 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction design. Because of a limited amount of coke supply, the coke gasification capacity of the plant was kept about the same; i. e., that amount which could be processed in two gasification trains. However, the F-T liquids production was maximized, and power production was reduced by using only one combined cycle train. The unconverted syngas and light hydrocarbons from the F-T synthesis section is compressed and combined with the small amount of syngas bypassing the F-T area to provide the fuel for the combustion turbine. Figure 3.1 shows the train configuration of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction.

Table III.1 compares the design input and output stream flows for the, Subtask 2.2 and Subtask 2.1 Coke Gasification Power Plants with Liquids Coproduction facilities and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. From 5,417 tpd of dry petroleum coke, the plant produces 10,450 bpd of liquid fuel precursors, 366.9 MW of export power, and 195.1 tpd of sulfur. This is about 2½ times the liquid fuel production of the Subtask 2.1 non-optimized plant. However, the export power production is reduced to 366.9 MW from the 617 MW of the non-optimized plant.

The Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction was optimized through a joint VIP review session and by additional design reviews. As a result, the following improvements were considered:

- Use activated carbon for final/trace sulfur removal
- Addition of refrigeration to increase the light oil recovery from the F-T area
- Replacement of the fired heater in the F-T catalyst preparation area with steam heating using high-pressure steam from the gasification block to eliminate the use of natural gas during normal operation
- CO_2 removal to enrich F-T offgas going to the combustion turbine by an amine system
- CO_2 removal to enrich F-T offgas going to the combustion turbine by a compression/ vaporization scheme
- Use of a modified Ge7FA+e combustion turbine that can use a fuel gas with a heating value below 200 Btu/scf; i.e., direct combustion of the low Btu F-T off gas in the gas turbine.
- Moisturization of the combustion turbine fuel gas
- More effective utilization of steam in the F-T area

All of the above items were implemented in the development of the final design for the optimized Subtask 2.2 plant except the two CO_2 removal schemes. The compression/vaporization scheme was eliminated form consideration because it had significant hydrocarbon fuel losses in the CO_2 product. The CO_2 removal by amine adsorption design was rejected because that design was more expensive and has reduced yields compared to the case in which the combustion turbine was modified to use a leaner fuel gas.

On a lower heating value (LHV) basis, the plant has a thermal efficiency 55.1% when the heating value of the byproduct sulfur is included and 53.2% when the byproduct sulfur is not included. On a higher heating value (HHV) basis, the plant has a thermal efficiency 56.7% when the heating value of the byproduct sulfur is included and 54.9% when the byproduct sulfur is not included. These thermal efficiencies are higher than those that would be obtained from a coke IGCC power plant of a similar design because it includes the heating value of the liquid fuel that is produced. Since the second law of thermodynamics states this liquid fuel cannot be used at a 100% thermal efficiency, the thermal efficiency of the plant will be somewhat lower when the final disposition of the liquid fuel is considered.

The LHV thermal efficiencies of the optimized Subtask 2.2 plant are 7.2% higher than those of the Subtask 2.1 non-optimized plant, and the HHV thermal efficiencies are 8.7% higher. This increase is the result of the increased liquids production relative to power production.

Figure 3.2 is a simplified block flow diagram of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. This figure also contains the flow rates of the major plant input and output streams. This plant can be considered to consist of three distinct main processing areas:

- The gasification island and air separation unit (Areas 100, 150, 250, 300, 350, 400, 420, and 800)
- The F-T hydrocarbon synthesis area (Areas 200 and 201)
- The power block (Areas 500 and 600)

In addition there is a balance of plant area (Area 900). The remainder of this section describes the three main processing areas, the balance of plant area, and discusses the plant EPC cost.

Table III.1

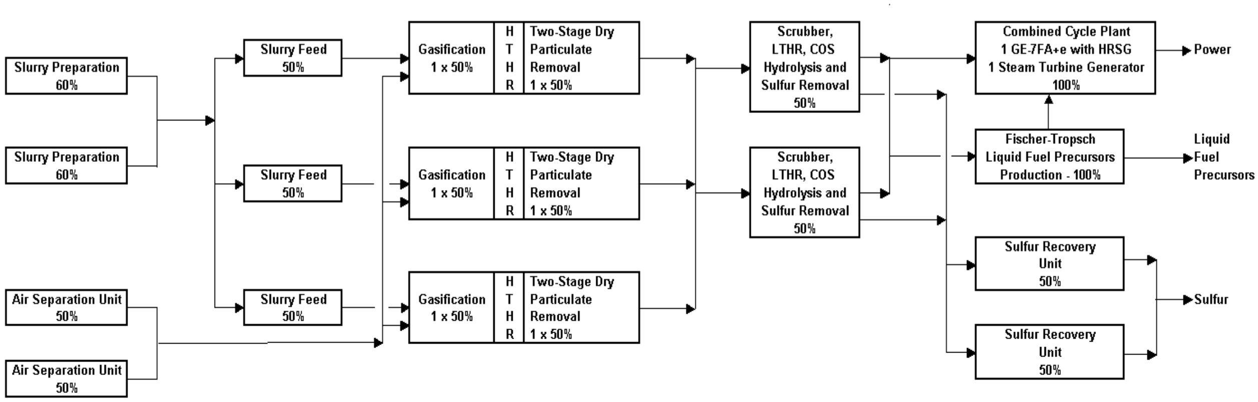
Design Input and Output Streams for the Subtask 2.1 and 2.2 Coke Gasification Power Plant with Liquids Coproduction Facilities and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

	Subtask 1.3 Next Plant	Subtask 2.1 Non-optimized Power and Liquids Plant	Subtask 2.2 Optimized Power and Liquids Plant
Plant Inputs			
Coke Feed, as received TPD	5,692	5,649	5,684
Dry Coke Feed to Gasifiers, TPD	5,417	5,376	5,417
Oxygen Production, TPD of 95% O_2	5,954	5,919	5,877
Total Fresh Water Consumption, gpm	5,223	6,472	5,693
Condensate Return from the Refinery, lb/hr	686,000	0	0
Flux, TPD	110.6	109.7	110.6
Natural Gas, MMBtu/hr	0	23.2	0
Plant Outputs			
Net Power Output, MW	474.0	617.0	366.9
Sulfur, TPD	373.4	370.6	373.4
Slag, TPD (15% moisture)	195.1	193.6	195.1
Hydrogen, MMscfd	80	0	0
HP Steam, 750°F/700 psia	980,000	0	0
Liquid Fuel Precursors, bpd	0	4,125	10,450

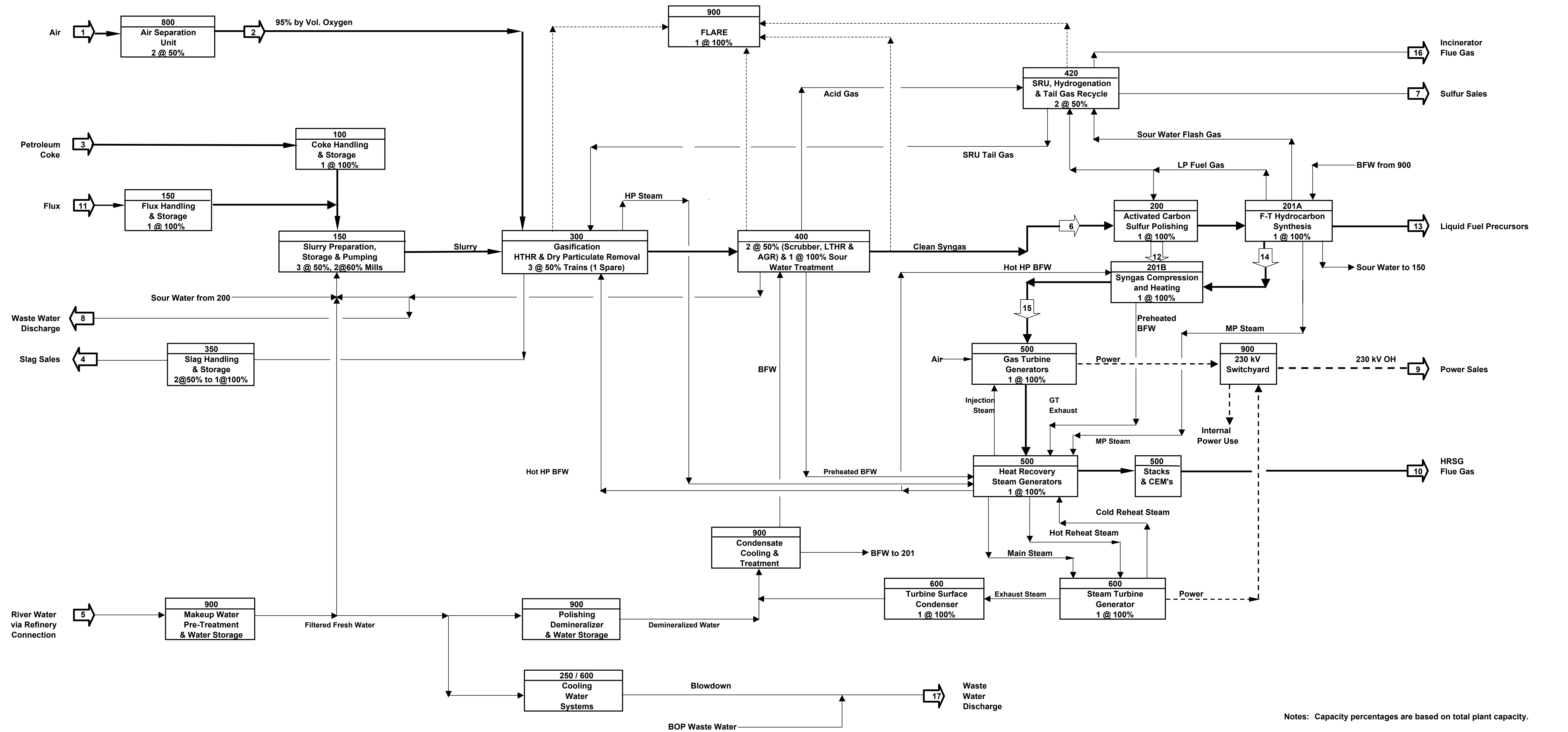
Figure 3.1

Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction

Simplified Block Train Diagram



Notes: Capacity percentages are based on total plant capacity.



Notes: Capacity percentages are based on total plant capacity.

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21
	Air	Oxygen	Coke	Slag	Water	Syngas	Sulfur	Water	Power	Flue Gas	Flux	Syngas	Liq Fuel	Fuel Gas	GT Fuel	Flue Gas	Water				
Flow	25,624	5,877	5,417	195.1	2,846,500	1,035,700	373.4	41,850	366,900	3,983,400	110.6	80,210	123,280	863,320	943,530	25,809	585,500				
	Tons/Day	Tons/Day	Tons/Day	Tons/Day	Lb/Hr	Lb/Hr	Tons/Day	Lb/Hr	kWe	Lb/Hr	Tons/Day	Lb/Hr	Lb/hr	Lb/hr	Lb/hr	Lb/Hr	Lb/Hr				
Nominal Pressure - psig	Atmos.	609	NA	NA	50	365	NA	62	NA	Atmos.	NA	360	50	314	445	Atmos.	Atmos.				
Temperature - F	70	240	Ambient	180	70	100	332	80	NA	227	NA	100	110	80	532	500	71				
HHV Btu/lb	NA	NA	14,848	NA	NA	4,847	3,983	NA	NA	NA	NA	4,847	19,689	1,744	2,008	NA	NA				
LHV Btu/lb	NA	NA	14,548	NA	NA	4,602	3,983	NA	NA	NA	NA	4,602	18,214	1,599	1,855	NA	NA				
Energy - MM HHV/hr	NA	NA	6,703	NA	NA	5,020	124	NA	NA	NA	NA	389	2,427	1,505	1,894	NA	NA				
Energy - MM LHV/hr	NA	NA	6,567	NA	NA	4,766	124	NA	NA	NA	NA	369	2,245	1,381	1,750	NA	NA				
Notes	Dry Basis	5,583 O2	Dry Basis	15%Wtr.	5693 GPM		Sales	84 GPM	230 kV			No S	10450 bpd				1171 GPM				

DOE Gasification Plant Cost and Performance Optimization

Figure 3.2

Subtask 2.2

OPTIMIZED COKE GASIFICATION POWER PLANT WITH LIQUIDS COPRODUCTION

BLOCK FLOW DIAGRAM

File: Fig 3.2.xls

July 24, 2003

3.1 Air Separation Unit and Gasification Island

The gasification island and the air separation unit basically are the same as those in the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant, as described in Appendix D of reference 2. Therefore, detailed descriptions of these areas will not be repeated here. The Gasification Island deals with the coke handling, gasification, syngas cooling and cleanup, sulfur production, and slag handling. These are Areas 100, 150, 250, 300, 350, 400, and 420. The Air Separation Unit has been renumbered to Area 800 for this case in order to allow the F-T hydrocarbon synthesis area to be Areas 200 and 201, which are more consistent with the nomenclature used in the indirect baseline study.

The fuel handling system (Area 100) provides the means to receive, unload, store, and convey the delayed petroleum coke to the storage facility. The coke and flux are mixed by the weigh belt feeders and transferred by coke feed conveyors to the day storage bins above the rod mills in the slurry preparation area.

The slurry preparation area also contains the flux receiving and storage facilities as well as the rod mills for grinding the coke. In order to produce the desired slurry solids concentration, coke is fed to each rod mill with water that is recycled from other areas of the gasification plant. Prepared slurry is stored in agitated tanks before being introduced into the first-stage of the gasifier.

The gasification, high temperature heat recovery, and particulate removal system (Area 300) is the heart of the Gasification Island. Global Energy's E-GAS™ gasification process consists of two stages, a slagging first-stage and an entrained flow non-slagging second-stage. The slagging section, or first-stage, is a horizontal refractory lined vessel into which the coke and flux slurry, recycle solids, and oxygen are atomized via opposing mixer nozzles. The coke and flux slurry, recycle solids, and oxygen are fed sub-stoichiometrically to the gasifier vessel at an elevated temperature and pressure to produce a high temperature syngas. The oxygen feed rate to the mixers is carefully controlled to maintain the gasification temperature above the ash fusion point; thereby ensuring good slag removal while producing high quality syngas.

The raw synthesis gas generated in the first-stage flows up from the horizontal section into the second-stage of the gasifier. The non-slagging second-stage is a vertical refractory-lined vessel into which additional coke slurry is injected via an atomizing nozzle to mix with the hot syngas stream exiting the first-stage. No oxygen is introduced into the second-stage. This additional slurry lowers the temperature of the gas exiting the first-stage by vaporizing the water in the slurry feed and by the endothermic nature of the steam and CO_2 reactions with carbon, thereby generating syngas and increasing cold gas efficiency.

The coke is almost totally gasified to form a synthetic fuel gas consisting primarily of hydrogen, carbon monoxide, carbon dioxide, and water. Sulfur in the coke is primarily converted to hydrogen sulfide (H_2S) with a small portion converted to carbonyl sulfide (COS); both of which are easily removed by downstream processing.

The gas and entrained particulate matter exiting the gasifier is further cooled in a firetube heat recovery boiler system to produce saturated steam at 1,650 psia which is superheated in the HRSG and used for power generation. The syngas leaving the high temperature heat recovery unit passes through a two-step cyclone/dry char filter particulate removal system to remove solids from the syngas. The recovered particulates are recycled to the gasifier. Water-soluble impurities are removed from the syngas in a wet scrubbing column following the dry char filters.

Mineral matter in the coke and flux form a molten slag which flows continuously through the tap hole into a water quench bath located below the first-stage. The slag then is crushed and removed through a continuous pressure let-down system as a slag/water slurry. This continuous slag removal technique eliminates high-maintenance, problem-prone lock hoppers and completely prevents the escape of raw gasification products to the atmosphere during slag removal.

The Area 350 slag handling and storage system processes and stores the slag. The slag slurry leaving the slag crushers at the outlet of the quench section of the gasifier flows continuously through the pressure let down system and into a dewatering bin. After passing through a settling tank to remove fine particles, the clear water is cooled in heat exchangers before it is returned to the gasifier quench section. The dewatered slag is loaded into trucks or rail cars for transport to market or to storage. The fines from the bottom of the settling tank are recycled to the slurry preparation area.

Area 400 contains the COS hydrolysis unit, low temperature heat recovery system, sour water treatment system, and the acid gas removal system.

Since COS is not removed efficiently by the downstream Acid Gas Removal (AGR) system, the COS must be converted to H_2S in order to obtain the desired high sulfur removal level. This is accomplished by the catalytic reaction of the COS with water vapor in the COS hydrolysis unit to create H_2S and CO_2. The H_2S is removed in the downstream AGR section and most of the CO_2 remains in the syngas.

Upon exiting the COS hydrolysis unit, the syngas is cooled in a series of shell and tube exchangers which condense water, ammonia, some carbon dioxide, and hydrogen sulfide in an aqueous solution. This water goes to the sour water treatment unit. Some of the cooled syngas goes to the syngas recycle compressor for use in various areas of the plant; such as for quench gas in the second-stage of the gasifier, particulate recycle, and for back pulsing the dry char filters.

The heat removed prior to the AGR unit provides moisturizing heat for the gas turbine fuel gas, steam for the AGR stripper, and condensate heat. Cooling water provides trim cooling to ensure that the syngas enters the AGR at a sufficiently low temperature.

The sour water treatment system removes the small amounts of dissolved gases (i.e., carbon dioxide, hydrogen sulfide, ammonia, and other trace contaminants) from the condensed water and any other process water. The gases are stripped out of the sour water in a two-step process. First, the acid gases are removed in the acid gas stripper column by steam stripping. The stripped gases go to the Sulfur Recovery Unit (SRU). The water from the acid gas stripper column, is cooled, and a major portion is recycled to slurry preparation. The remainder is treated in the ammonia stripper column to remove the ammonia, filtered to remove trace organics and solids, and then sent to the waste water management system. The stripped ammonia is combined with water that will be recycled back to the slurry mix tank after being cooled with cooling tower water.

The acid gas removal (AGR) system removes the H_2S from the syngas to produce a low sulfur syngas. The H_2S is removed from the sour syngas in an absorber column at high pressure and low temperature using a solvent, methydiethanolamine (MDEA). After the hydrogen sulfide removal, all of the un-moisturized syngas is sent to Area 200 for sulfur polishing before F-T synthesis.

The H_2S rich MDEA solution leaving the absorber goes to a stripper column where the H_2S is removed by steam stripping at a lower pressure. The concentrated H_2S exits the top of the stripper column and goes to the Sulfur Recovery Unit. The lean amine exits the bottom of the stripper, is cooled, and then recycled to the absorber. An online MDEA reclaim unit continuously removes impurities from the lean amine to improve system efficiency.

The Area 420 sulfur recovery unit (SRU) processes the concentrated H_2S from the AGR unit and the CO_2 and H_2S stripped from the sour water in a reaction furnace, a waste heat recovery boiler, and then in a series of Claus catalytic reaction stages where the H_2S is converted to elemental sulfur. The sulfur is recovered as a molten liquid and sold as a by-product.

The tail gas stream, composed of mostly carbon dioxide and nitrogen with trace amounts of sulfur dioxide, exits the last catalytic stage and goes to tail gas recycling. It is hydrogenated to convert all the remaining sulfur species to H_2S, cooled to condense the bulk of the water, compressed, and then injected into the gasifier. This allows for a very high sulfur removal efficiency with low recycle rates.

Area 800 contains two 50% capacity Air Separation Units (ASUs) to deliver the required oxygen for the coke gasification process. Each ASU consists of several subsystems and major pieces of equipment, including an air compressor, air cooling system, air purification system, cold box, and product handling and backup systems.

Gaseous oxygen leaves the cold boxes at moderate pressure and is compressed in centrifugal compressors and delivered to the gasifiers. Nitrogen tanks with steam vaporizers provide gaseous nitrogen for various in-plant uses such as purging vessels.

The Area 250 cooling water system provides cooling water to the gasification island and ASU. A second system provides the cooling duty for the power block.

The major components of the cooling water system consist of a cooling tower, circulating water pumps, and appropriate piping for distribution of the cooling water around the facility. Both cooling towers are multi-cell mechanically induced draft towers, sized to provide the design heat rejection at the ambient conditions corresponding to the maximum summer temperature. Chemical treatment systems, including metering pumps, storage tanks and unloading facilities provide the necessary biocide, pH treatment and corrosion inhibiting chemicals for the circulating water system. Cooling tower blowdown discharges to the wastewater management system.

3.2 Fischer-Tropsch Hydrocarbon Synthesis Area

The design for the Fischer-Tropsch Hydrocarbon Synthesis Area was developed based on the ASPEN Plus process flowsheet reactor model that was developed for the Baseline Design/Economics for Advanced Fischer-Tropsch Technology study.[4] The ASPEN Plus process flowsheet model of the Fischer-Tropsch (F-T) hydrocarbon synthesis area that was developed for this study does not include the following systems:

- Filter system (and associated hydrocarbon circulation loop) which removes the catalyst from the liquid product leaving the slurry-bed F-T reactor

- Used catalyst removal and disposal system
- Fresh catalyst handling and pretreatment systems

The designs for these systems were developed based on the previous Baseline F-T design study.[3]

The Fischer-Tropsch hydrocarbon synthesis area consists of two sub areas, Area 200 and Area 201. Area 200 is the Final Syngas Cleanup Area, which removes the final traces of sulfur from the syngas, before it is converted to hydrocarbons in Area 201, the Hydrocarbon Synthesis and Product Recovery Area.

3.2.1 Final Syngas Cleanup Area

Description

The purpose of the Final Syngas Cleanup Area, Area 200, is to reduce the sulfur concentration of the cleaned syngas from the acid gas removal area of the gasification block to less than 0.5 ppm of sulfur. This is accomplished by absorbing the small amounts hydrogen sulfide, carbonyl sulfide (COS) and trace amounts of other light organic sulfur compounds (such as CS_2) on metal impregnated activated carbon. The active bed is regenerated weekly with medium-pressure steam and a small quantity of air, and the off gas is sent to the sour water stripper (SWS) overhead cooling system to condense the steam prior to going to Claus sulfur recovery. After its useful life, the deactivated carbon is sent to the gasifier for destruction and conversion to syngas and slag. The metal activator is entrained in the slag, which is a non-hazardous waste.

In order to provide continuous H_2S removal, the process design uses a three bed reactor configuration with two beds in series to remove sulfur (the second bed is a guard bed). The third bed is in regeneration. Necessary piping is provided so that these beds can be switched into any position, and when necessary, the spent adsorbent can be replaced without any interruption of service. When H_2S breakthrough occurs in the first bed (lead bed), it is taken out of service for regeneration (or adsorbent replacement, when necessary), and the other bed (lag bed) is placed in the first position. The freshly regenerated bed now becomes the second bed. This two bed in series operation continues until H_2S breakthrough occurs in the first bed, and it is removed from service for regeneration causing the operating cycle to repeat. Each carbon bed is sized for a one week cycle. Each activated carbon bed has an expected life of about three years so that, on average, one bed should be replaced each year.

Figure 3.3 contains a schematic flow diagram of the Final Syngas Cleanup Area, Area 200 and the F-T slurry-bed reactor and product recovery area, Area 201.

The Area 200 Final syngas cleanup area is sized to treat all of the syngas produced by the gasification block. After treatment, about 92% of the cleaned syngas is set to the Area 201 Fischer-Tropsch slurry-bed reactor. The remaining cleaned syngas is mixed with the unconverted syngas and light hydrocarbons from the F-T area, compressed, moisturized, heated, and mixed with steam to become fuel gas for the combustion turbine.

Comparison with the Subtask 2.1 Design

The Subtask 2.2 design is a lot simpler and less expensive than the non-optimized Subtask 2.1 design. In the Subtask 2.2 design, the entire Final Syngas cleanup area consists of only three vessels containing the activated carbon absorbent and associated vales and piping. The two heat exchangers and furnace have been eliminated, and no fuel is required. The elimination of the furnace means that no natural gas is required during normal operation.

Furthermore, the activated carbon adsorbent is regenerable with medium-pressure steam. When its active life is expended after about three years, the active carbon can be mixed with the coke and fed to the gasifiers thereby eliminating the disposal problem that was associated with the ZnO adsorbent used in Subtask 2.1.

Another advantage of the Subtask 2.2 design is that all the syngas from the gasification block goes through the activated carbon beds so that the total sulfur content of the gas turbine fuel is less than 0.5 ppm of sulfur; whereas in the Subtask 2.1 design, the sulfur content of the fuel gas bypassing the F-T area was about 22 ppm.

3.2.2 Fischer-Tropsch Slurry-bed Reactor Area

Description

The Fischer-Tropsch slurry-bed reactor converts the sulfur-free syngas primarily into olefinic hydrocarbons by the reaction

$$n\,CO + 2n\,H_2 \rightarrow C_nH_{2n} + n\,H_2O$$

The reaction is promoted by an iron-based catalyst, which also promotes the water-gas shift reaction

$$CO + H_2O \rightarrow CO_2 + H_2$$

The lighter hydrocarbon products leave the slurry-bed reactor in the vapor phase, are cooled and the condensed liquid collected. The heavier hydrocarbons are removed as liquids from the reactor, separated from the suspended catalyst, cooled, and combined with the lighter products to form the liquid fuel precursors product.

In order to maintain a constant catalyst activity, there is a continual addition of fresh catalyst and a continual withdrawal of used catalyst from the slurry-bed reactor. The fresh catalyst must be pretreated in a reducing atmosphere at an elevated temperature to activate it. The catalyst pretreating system consists of a similar vessel to the slurry-bed reactor, but without the internal cooling facilities.

Figures 3.3 and 3.4 contain flow diagrams of the syngas and liquids processing areas of Plants 200 and 201. Figure 3.5 contains flow diagrams of the catalyst preparation and removal systems in the F-T processing area.

The cleaned syngas from the gasification block is preheated to 244ºF with low-pressure steam in heat exchanger 201E-21. The preheated syngas leaving the 201E-21 heat exchanger is mixed with 440ºF/375 psia stream that was generated in the slurry-bed F-T reactor and fed to the 200R-1 slurry-bed F-T hydrocarbon synthesis reactor.

The slurry-bed F-T hydrocarbon synthesis reactor, 201R-1, converts the hydrogen and carbon monoxide to straight chain aliphatic hydrocarbons, carbon dioxide and water. The heat of reaction is removed from the slurry-bed F-T reactor by the generation of 440°F/375 psia steam inside tubes located within the slurry-bed reactor. Pump 201P-2 circulates boiler feed water (BFW) between the 201C-1 steam drum and the 201R-1 reactor to ensure that sufficient BFW always is flowing through the cooling tubes.

Cyclone 201T-1 removes entrained catalyst particles from the vapor stream leaving the top of the F-T reactor. The vapor stream then is cooled to 40°F in four exchangers. The first exchanger (201E-1) cools the syngas to 130ºF by heating BFW. The cooled syngas leaving the first exchanger enters the 201C-22 reactor overhead flash drum to separate condensate. The next exchanger (201E-2) cools the syngas to 100ºF with cooling water. The next two exchangers (210E-19 and 201E-20) chill the syngas to 40ºF. The chilled syngas enters the 201C-2 reactor overhead flash drum. The sour water from the boot of 201C-2 goes to the 201C-4 sour water flash drum. The vapor stream leaving the sour water flash drum goes to the incinerator, and the sour water is recycled to the gasifier.

The vapor stream from the reactor overhead vapor flash drum is routed through 201E-19 to recover refrigeration and is washed in 201C-3 to remove any residual catalyst particles prior to compression. A propane refrigeration system provides the refrigerant used in 201E-20. The washed vapor stream is mixed with clean syngas from final gas cleanup and is compressed to 475 psia in 201K-1. This compressed stream is a high-pressure fuel gas stream which is sent to the power block where it is moisturized, heated to 425ºF with intermediate pressure (400 psia) steam in 201E-23 and is used for combustion turbine fuel. This stream consists of unconverted syngas (carbon monoxide and hydrogen) and light hydrocarbons (primarily C1 through C5s) that are produced in the F-T reactor.

The liquid hydrocarbon stream leaving 201C-2 is mixed with the cooled liquid hydrocarbons from the slurry-bed F-T reactor and sent for upgrading into liquid transportation fuels.

The liquid stream leaving the slurry-bed F-T reactor passes through hydroclone 201T-2 to remove a majority of the entrained catalyst particles. The catalyst-rich hydroclone bottoms goes to mixing tank 201C-10 from which most of it is returned to the slurry-bed reactor by pump 201P-3. A portion of the hydroclone bottoms is withdrawn and sent to the catalyst withdrawal system shown in Figure 3.3. Residual catalyst particles are removed from the hydroclone overhead stream in the 201T-3 filter system.

The catalyst-free liquid leaving the filter system is reduced in pressure and flashed in drum 201C-5. The vapor stream is further cooled to 100°F in exchanger 201E-4 with cooling water and flashed in drum 201C-6. The vapor stream from drum 201C-6 is a low-pressure fuel gas which is used as fuel in the 200F-1 furnace.

The liquid leaving the 201C-5 flash drum is cooled to 200ºF in 201E-5 by preheating boiler feed water. The cooled liquid from 201C-5 is mixed with the liquid stream from the 201C-6 flash drum in drum 201C-9 and a cooled liquid recycle stream from 201C-8. This mixture now is cooled to 110ºF by cooling water in exchanger 201E-6 and sent to the 201C-8 liquid fuel flash drum along with the liquid from the 201C-2 reactor overhead vapor flash drum. The vapor leaving the 201C-8 liquid fuel flash drum is mixed with the vapor from the 201C-6 flash drum and is used as low-pressure fuel gas in the gasification area.

The liquid from the 201C-8 flash drum is split into two streams. One of the liquid streams is recycled back to 201C-9 flash drum via pump 201P-1 to dilute the heavier hydrocarbons in order to control their viscosity as they are cooled in exchanger 201E-6. The other liquid stream is the liquid fuel precursors product which is sent to the adjacent petroleum refinery for upgrading into liquid transportation fuels (gasoline, diesel, etc.).

Figure 3.5 shows the catalyst withdrawal system. The hot catalyst-rich stream from the 201C-10 drum is cooled in exchanger 201E-7 and pumped by pump 201P-4 through the 201T-4 filters to remove the used catalyst which is collected and discarded. The catalyst free liquid is mixed with the liquid fuel precursors product stream from drum 201C-6 and sent to the adjacent petroleum refinery for upgrading.

The catalyst pretreatment system also is shown in Figure 3.5. The makeup catalyst is fed into the 201C-11 catalyst pretreater where it is combined with heated liquid product from storage. Recycle gas is circulated through the pretreater vessel via compressor 201K-2, exchanger 201E-9, and exchanger 201E-11, which uses high-pressure steam from the gasification plant as the heating media. Vapors leaving the pretreater vessel are cooled in exchangers 201E-9 and 201E-10 before being flashed in drum 201C-13. A portion of the vapor from 201C-13 is withdrawn and sent to the incinerator to remove inerts from the system. However, most of the vapors from 201C-13 are recycled to the pretreater after addition of some fresh syngas or hydrogen via the 201K-2 compressor, exchanger 201E-9, and exchanger 201E-11. Pretreated catalyst is withdrawn from the pretreater vessel and stored in the heated 201C-12 mixing tank until it is injected into the slurry-bed F-T reactors via pump 201P-8.

Comparison with the Subtask 2.1 Design

Several changes have been made to the Fischer-Tropsch Slurry-bed Reactor Area of the Subtask 2.2 design to improve it.

Since the final sulfur removal operation is now done at about room temperature, the 201E-21 syngas preheater was added to heat the syngas going top the F-T reactor. Low-pressure steam is used as the heating media.

Flash drum 201C-22 was added between the 201E-1 and 201E-2 heat exchangers to remove condensed water from the F-T reactor overhead vapor stream before it if further cooled. The removal of the water from the vapor at this point reduces the load on the downstream cooling systems.

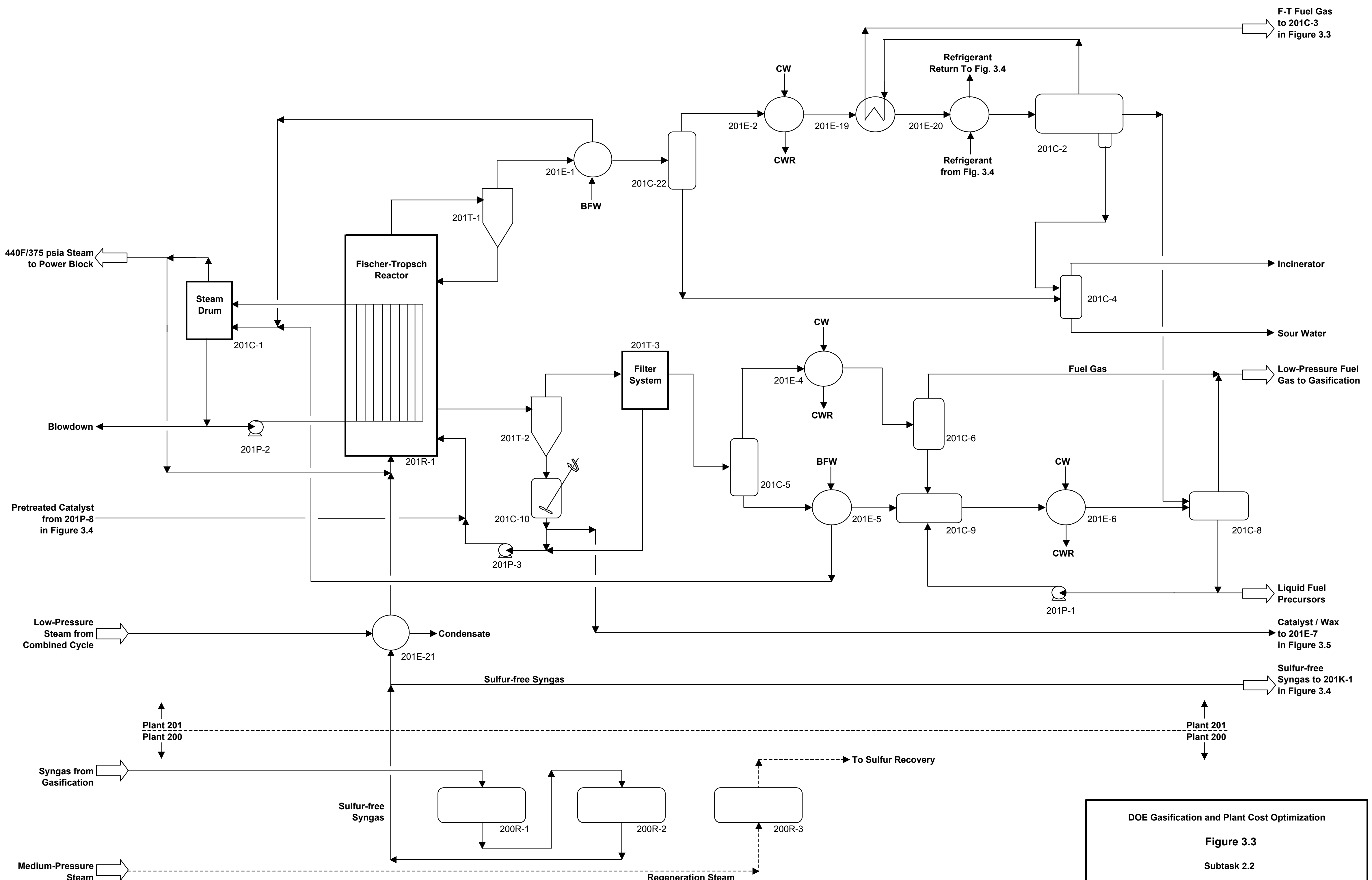
Cooling by a refrigeration system has been added between the 201E-2 exchanger and the 201C-2 flash drum to increase the recovery of liquid hydrocarbons produced in the F-T reactor rather than have them be consumed in the combustion turbine. To save refrigeration duty, the 201E-2 exchanger was enlarged so that the vapor stream is now cooled to 100ºF with cooling water rather than 110ºF in the Subtask 2.1 design. The refrigeration system required the addition of the 201E-18, 201E-19 and 201E-20 exchangers, the 201C-23 and 201C-24 drums, and the 201K-3 compressor.

As a result of the increased F-T liquids recovery, the combined combustion turbine fuel gas has a lower heating value of about 164 Btu/scf unmoisturized. Based on previous information from

General Electric, this fuel gas could be used in the GE7FA+e combustion turbine when moisturized to a lower heating value of 147 Btu/scf for NOx control but would require a higher inlet pressure. Thus, in the Subtask 2.2 design, the complete gas turbine fuel gas (F-T product gas and syngas bypassing the F-T area) is compressed in the 201K-1 compressor to 475 psia. It is then moisturized in the C-195 column of the gasification block using low-level heat from syngas cooling, and heated to 425ºF with intermediate pressure steam from the gasification block before going to the combustion turbine. For improved reliability, the 201K-1 compressor is spared to maximize syngas availability to the turbine.

The Subtask 2.1 design only compressed the F-T product gas to 380 psia, mixed it with the bypass syngas, and used steam as diluent for NOx control. In this design, more water is goes through the combustion turbine, HRSG and into the atmosphere requiring a larger makeup water purification system. In Subtask 2.1, the F-T product gas compressor was not spared because the power block could operate at reduced capacity during a compressor outage since the bypassed syngas provided most of the turbine fuel. In this situation, the F-T product gas either could be sent to the incinerator to generate steam, or it could be flared.

The Subtask 2.2 catalyst pretreatment system is essentially the same except that 201F-1 furnace has been replaced with a heat exchanger that uses 1400 psia stream from the gasification block. In addition, 1400 psia steam also is used in the 201E-8 heat exchanger. The elimination of the 201F-1 furnace (and the 200F-1 furnace) removes the need for natural gas during normal operation. The low-pressure fuel gas which was previously used to partially fire these furnaces now is sent to the incinerator in the gasification block where it is used to make intermediate pressure steam.



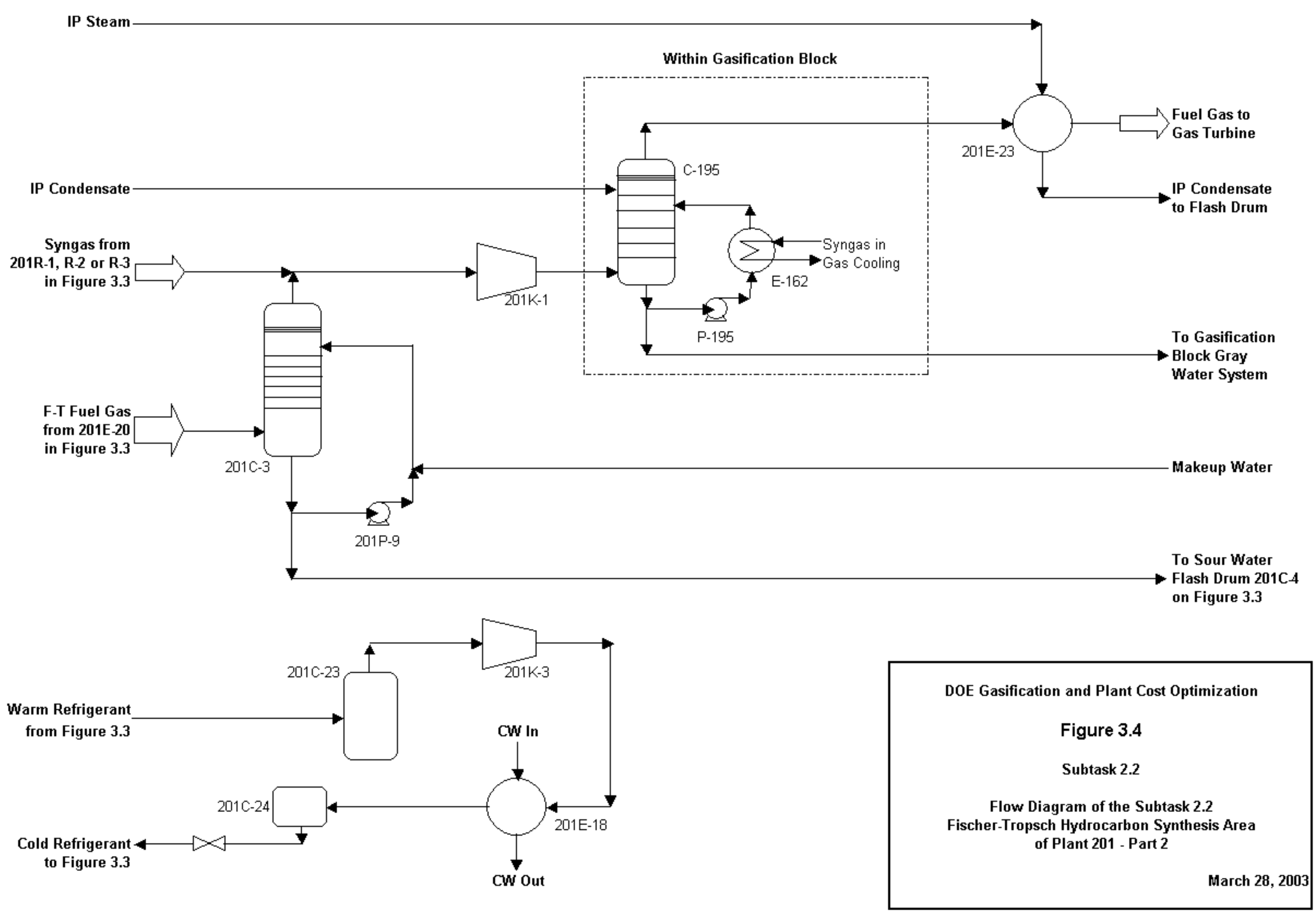
DOE Gasification and Plant Cost Optimization

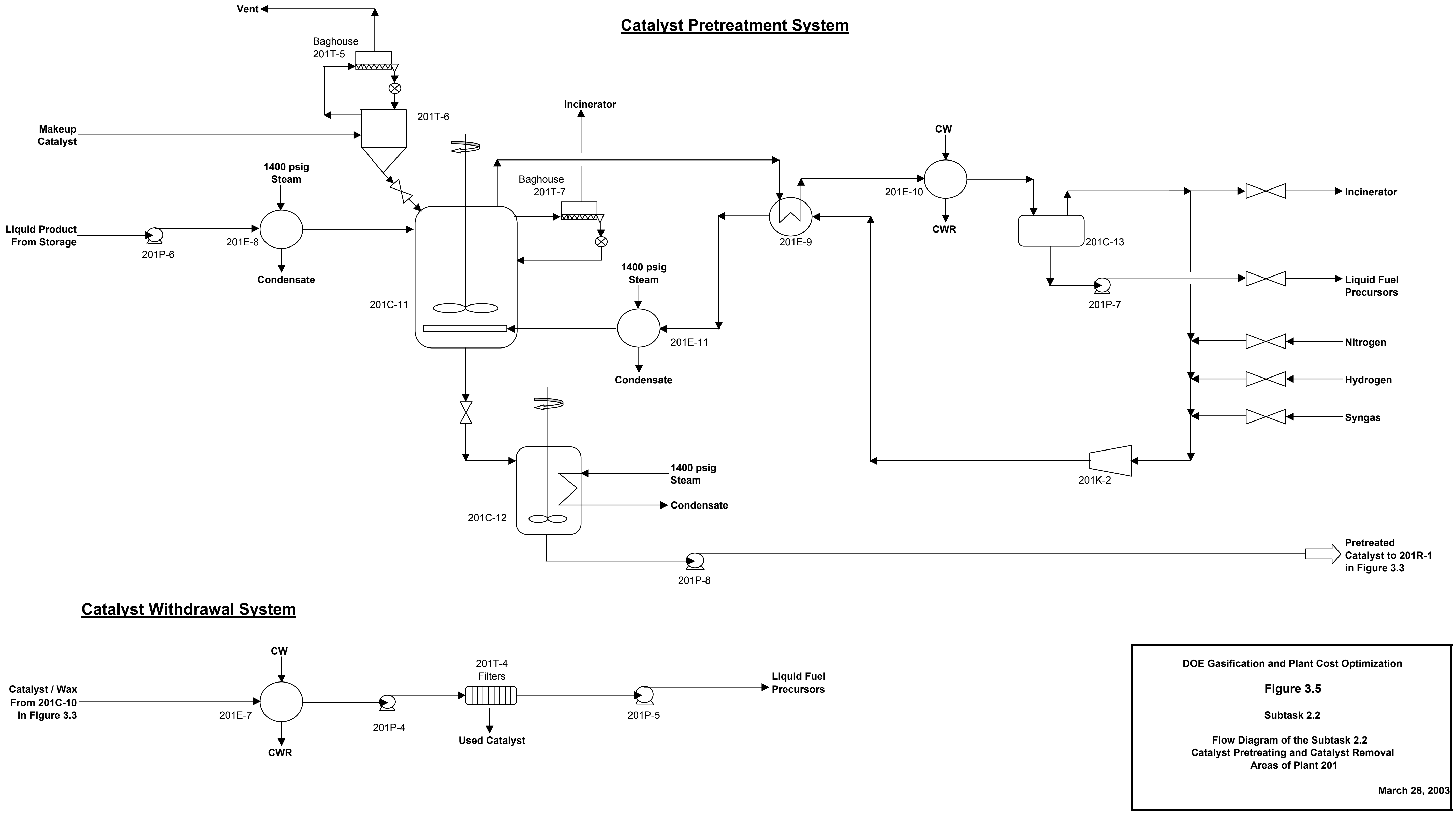
Figure 3.3

Subtask 2.2

Flow Diagram of the Subtask 2.2 Syngas Cleanup and Fischer-Tropsch Hydrocarbon Synthesis Areas of Plants 200 and 201

July 24, 2003


IP Steam
Within Gasification Block
Fuel Gas to
Gas Turbine
201E-23
C-195
IP Condensate
IP Condensate
to Flash Drum
Syngas from
201R-1, R-2 or R-3
in Figure 3.3
Syngas in
Gas Cooling
E-162
201K-1
P-195
To Gasification
Block Gray
Water System
F-T Fuel Gas
from 201E-20
in Figure 3.3
201C-3
Makeup Water
201P-9
To Sour Water
Flash Drum 201C-4
on Figure 3.3
201C-23
201K-3
Warm Refrigerant
from Figure 3.3
CW In
201C-24
201E-18
Cold Refrigerant
to Figure 3.3
CW Out
DOE Gasification and Plant Cost Optimization
Figure 3.4
Subtask 2.2
Flow Diagram of the Subtask 2.2
Fischer-Tropsch Hydrocarbon Synthesis Area
of Plant 201 - Part 2
March 28, 2003



DOE Gasification and Plant Cost Optimization

Figure 3.5

Subtask 2.2

Flow Diagram of the Subtask 2.2
Catalyst Pretreating and Catalyst Removal
Areas of Plant 201

March 28, 2003

3.3 Power Block

The major components of the power block include a gas turbine generator (GTG), a heat recovery steam generator (HRSG), a steam turbine generator (STG), and numerous supporting facilities.

Area 500 contains the gas turbine (GT), heat recovery steam generator (HRSG), and stack. The combustion turbine generator is a General Electric 7FA+e machine with a nominal output of 199.4 MW. The GT utilizes moisturized syngas at 147.1 LHV Btu/scf without additional diluent injection for NO_x control. Combustion exhaust gases from the GT are routed to the HRSG and stack. Natural gas is used as back-up fuel for the gas turbine during startup, shutdown, and short duration transients in syngas supply. Optionally, the gas turbine can be fully fired on natural gas to generate power when syngas is unavailable.

The HRSG receives the GT exhaust gases and generates steam at the main steam, medium-pressure steam and low-pressure steam energy levels (non-reheat). The HRSG provides superheating capability for all three steam pressure levels including superheating of process area steam. The HRSG also provides condensate and feedwater heating for both the combined cycle and the gasification facilities. The HRSG is a fully integrated system consisting of all required ductwork and boiler components. Each component is designed for pressurized operation.

The HRSG boilers include steam drums for proper steam purity and to reduce surge during cold start. Large unheated down comers assure proper circulation in each of the banks.

The stack includes a continuous emission monitoring (CEM) system.

The Area 600 steam turbine (ST) is a non-reheat, condensing turbine that includes a HP/MP section and a four-flow down-exhaust LP section. Turbine exhaust steam is condensed in a surface condenser. The steam turbine produces 274.9 MW of electric power.

The power delivery system includes the GT generator output at 18 kilovolts (kV) connected through a generator breaker to its associated main power step-up transformer. A separate main step-up transformer and generator breaker is included for the ST generator. The HV switch yard receives the energy from the two generator step-up transformers at 230 kV.

An auxiliary transformer is connected between the GTG breaker and the step-up transformers. Due to the large auxiliary load associated with the plant, internal power is distributed at 33 kV from the two auxiliary power transformers. The second auxiliary transformer is connected between the steam turbine generator at 24 kV and the steam turbine main step-up transformer. The major motor loads in the air separation units are serviced by 33/13.8 kV transformers. Several substations serve the other internal loads with 33/4.16 kV transformers supplying a double-ended electrical bus.

Area 600 also includes a cooling water system similar to that in Area 250 and an emergency shutdown transformer to connect the 230 kV switch yard with essential safe shutdown loads.

3.4 Balance of Plant

The Area 900 balance of plant contains nine subsystems.

The fresh water supply system filters river water from an industrial water supply network for the fresh makeup water supply. A demineralizer supplies demineralized water for boiler water makeup. The demineralizer regeneration wastewater is sent to a process waste collection tank, where it is neutralized before discharge.

The fire and service water system includes a loop around the principal facilities with fire hydrants located for easy access. It also includes an onsite water storage tank. A jockey pump maintains line pressure in the loop during stand-by periods. During periods of high water usage, motor and diesel driven pumps are available.

The waste water management system processes both clear wastewater and storm water from a clean water collection pond. Clear wastewater includes water treatment effluent, cooling water blowdown, flushes and purges from equipment maintenance, filtered water from the ammonia stripper column (in Area 400), clarifier overflow, and sewage treatment overflow. Storm water is collected in a storm-water pond before going to the clean water collection pond. The water in the clean water collection pond is analyzed and treated, as required, until it meets permitted outfall specifications for discharge through the refinery waste water system.

The service and instrument air system provides compressed air and dried instrument air to users throughout the plant. The system consists of air compressors, air receivers, hose stations, and piping distribution for each unit.

The incineration system destroys the tank vent streams from various in-process storage tanks and drums that may contain small amounts of hydrocarbons and other gases such as ammonia and acid gas. During process upsets of SRU, tail gas streams also can be processed in the high temperature incinerator. The high temperature produced in the incinerator thermally destroys any residual hydrogen sulfide before the gas is vented to the atmosphere. The incinerator exhaust feeds into a heat recovery boiler to produce process steam.

The flare system provides for safe disposal of syngas during startup or short term upsets. The flare includes a natural gas fired pilot flame to ensure that the flare is continually operating.

The instrumentation and control system provides data acquisition, monitoring, alarming and control by a digital distributed control system (DCS). The DCS allows the plant to be operated from the central control room using the DCS as the control platforms. The gas and steam turbines, and the coke handling programmable logic controllers will continue to execute all permissive, protective, and sequence control related to their respective equipment. They will be controlled either locally using the turbine vendor man machine interface system, or from the DCS.

Other balance of plant equipment such as air compressors, condenser vacuum pumps, and water treatment facilities can be controlled by either local PLCs, or from contact and relay control cabinets. All remaining plant components are exclusively controlled by the DCS including the HRSG, the gasifier, ASU, hydrogen plant, electrical distribution, and other power block and gasification support systems.

The plant has a central building housing the main control room, office, training, other administration areas, and a warehouse/maintenance area. Other buildings are provided for water

treatment equipment and the motor control centers. The buildings, with the exception of water treatment, are heated and air-conditioned to provide a climate-controlled environment for personnel and electrical control equipment.

A series of strategically placed safety showers are located throughout the facility.

3.5 Plant Layout

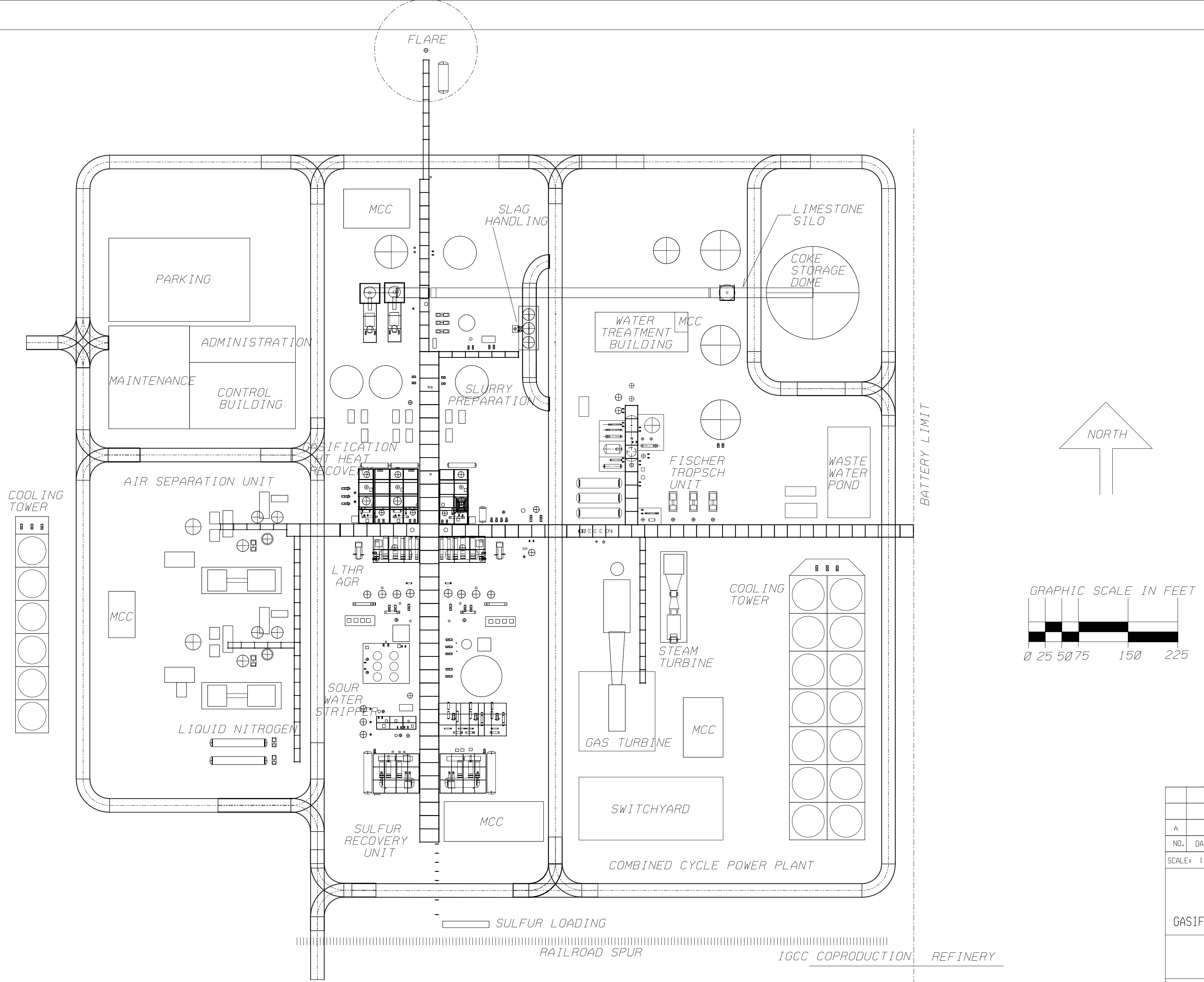
Figure 3.6 is a site plan of the Subtask 2.2 Optimized Coke Gasification Power Plant with F-T Liquids Coproduction. The plant occupies about 52 acres and is the same size as the site of the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant.

The site plan is very similar to that of the Subtask 1.3 Next Plant. However, in the Subtask 2.2 plant, the Area 200 and 201 F-T hydrocarbon synthesis areas replace the two hydrogen production facilities of the Subtask 1.3 Next Plant. Also the combined cycle power plant area of the Subtask 2.2 plant is smaller than that of the Subtask 1.3 Next Plant because one of the two combustion turbines is deleted. However, this space is now occupied by a much larger cooling tower containing 14 cells rather than the five cell tower in the Subtask 1.3 Next Plant. The larger cooling tower is needed to condense the additional high-pressure (750ºF/700 psig) steam that goes to the steam power cycle that was exported to the adjacent refinery in Subtask 2.2 and the medium-pressure (440ºF/360 psig) steam generated in the F-T area. Furthermore, the Subtask 2.3 steam cycle is a non-reheat cycle that is less efficient and has to reject more heat than the reheat steam cycle used in Subtask 1.3 Next Plant.

3.6 Thermal Efficiency

Table III.2 shows the thermal efficiencies of the Subtask 2.2 [Optimized] and Subtask 2.1 [Non-optimized] Coke Gasification Power Plants with Liquids Coproduction based on the energy content of the F-T liquid fuel, sulfur byproduct, and the equivalent energy in the export power. Also shown are the thermal efficiencies of the Subtask 1.5B Coke IGCC Power Plant. The thermal efficiencies of the Subtask 2.2 optimized plant are significantly higher than those of the Subtask 2.1 non-optimized plant because more of the energy leaves in the liquid F-T product. The efficiency of syngas conversion to liquids (excluding the syngas that becomes fuel gas) is about 75%. Most of the remaining 25% is recovered as steam for power generation. In comparison, the efficiency of a combined-cycle power plant is less than 60%, and the steam cycle efficiency is less than 30%. Therefore, increasing the production of F-T liquids increases the overall plant efficiency. It is for this reason, that both the Subtask 2.1 and 2.2 plants have higher thermal efficiencies than the Subtask 1.5B Coke IGCC Power Plant.

On an LHV basis, the thermal efficiency of the Subtask 2.2 optimized plant is 55.14% which is 7.38% higher than the thermal efficiency of the Subtask 2.1 non-optimized plant. Most of this increase is because the Subtask 2.2 plant produces a larger portion of F-T liquids than the Subtask 2.1 plant. Only a small portion is attributable to the improved processing efficiency of the Subtask 2.2 plant. On an HHV basis, the thermal efficiency of the Subtask 2.2 optimized plant is 56.74% which is 8.87% higher than that of the non-optimized plant. Excluding the energy content of the byproduct sulfur would lower the reported efficiencies by 1.8 to 1.9%.



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NO.	DATE	REVISIONS	BY	CHK	SUPV	PEM	CLIENT

SCALE: 1 IN = 75 FT. DESIGNED BY: G.M.WORTHY DRAWN BY: G.M.WORTHY

BECHTEL - GLOBAL ENERGY

US DEPARTMENT OF ENERGY

GASIFICATION PLANT COST AND PERFORMANCE OPTIMIZATION

OPTIMIZED COKE GASIFICATION POWER PLANT
WITH LIQUIDS COPRODUCTION
SUBTASK 2.2

SITE PLAN

BECHTEL

JOB NO:	DRAWING NO:	REV.
24355-104	SK - ØØ122	A

The HHV thermal efficiencies of these two plants are higher than the LHV thermal efficiencies because the F-T liquids have higher hydrogen contents than the coke feed. The catalyst In the F-T reactor promotes the water-gas shift reaction, which makes hydrogen from CO and water, and this hydrogen then becomes part of the F-T liquid product. When burned, the hydrogen is converted to water, and the difference between the LHV and HHV values is the latent heat of condensation of the product water. Because the Subtask 2.2 plant makes relatively more F-T liquids than the Subtask 2.1 plant, the difference between its HHV and LHV thermal efficiency is larger.

The LHV thermal efficiency for the Subtask 1.5B Coke IGCC Power Plant is greater than its HHV thermal efficiency because some of the hydrogen in the coke leaves the plant as water vapor in the stack gases, and very latent heat of vaporization is recovered as power.

Table III.2

Thermal Efficiencies of Three Coke Gasification Plants

	Subtask 2.2	Subtask 2.1	Subtask 1.5B Coke Power Plant
LHV Basis	55.14%	47.76%	43.35
HHV Basis	56.74%	47.87%	42.48

3.7 Emissions

Table III.3 shows the atmospheric emissions summary of the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant, the Subtask 2.1 [Non-optimized], and the Subtask 2.2 Optimized Coke Gasification Power Plants with Liquids Coproduction. All three plants process about the same amount of petroleum coke. The atmospheric emissions of the two Task 2 plants are lower than those of the Subtask 1.3 Next Plant with the Subtask 2.2 plant having the lowest emissions of all the three plants. The Subtask 2.2 plant emits about, 4,254,000 lb/hr of total exhaust gases having an average SOx concentration of 34 ppmv, an average NOx concentration of 16 ppmv, and an average CO concentration of 13 ppmv. Expressed another way, this is 276 lb/hr of SOx (as SO_2), 94 lb/hr of NOx (as NO_2), and 37 lb/hr of CO.

The sulfur emissions from the Subtask 2.2 gas turbine are very low because almost all the sulfur is removed from the syngas by adsorption on activated carbon (and recovered as sulfur) whether it goes to the F-T synthesis reactors or whether it goes directly to the turbine. In the Subtask 2.1 plant, sulfur is removed by adsorption on ZnO (and discarded as ZnS) only from that portion of the syngas going to the F-T reactor. In the Subtask 1.3 Next Plant, the sulfur content of the syngas going to the gas turbine is about 20 ppmv. The NOx and CO emissions from the gas turbines for all three cases are about the same when expressed on ppmv (at 15% oxygen, dry basis). However, the Subtask 2.2 NO_2 and CO rates are the highest on an weight basis because the Subtask 2.2 turbine exhaust is drier since the CO_2 produced in the F-T area is used for NOx control (instead of steam injection).

All three cases have about the same absolute sulfur emissions from the incinerator stack because the sulfur comes from the purge and blow down streams from the gasification block, and they are about the same in all three cases. However, the Subtask 1.3 Next Plant has a much larger flow because it contains the reject CO_2 product from the hydrogen production facilities. The slightly lower sulfur emissions from the Task 2 plants is the result of small process improvements made since the Subtask 1.6 design was developed. On a ppmv basis, all three cases produce about the same amount of NOx and CO, but because of the larger flow rate in the Subtask 1.3 plant, the absolute rates are higher.

In addition to the above atmospheric emissions, the Subtask 2.1 and 2.2 plants generate some solid wastes. The Subtask 2.1 non-optimized plant generates about 360,000 lb/year of used F-T catalyst, 500 ft^3/year of used COS hydrolysis catalyst, and 5,400 ft^3/year of used ZnO adsorbent as ZnS. The Subtask 2.2 optimized plant generates about 867,000 lb/year of used F-T catalyst. It also generates about 100,000 lb/year (3,000 ft^3/year) of used activated carbon adsorbent which is disposed of by mixing it with the coal feed and gasifing it to make syngas.

All three plants also generate about 195 tpd of slag which is a non-hazardous byproduct that is used in construction projects.

3.8 Plant Cost

Table III.4 shows the "overnight" EPC cost for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction and compares it with that of the Subtask [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Plant IGCC Coproduction Plant. These costs are on a mid-year 2000 basis; the same basis as those of the other Task 1 plant costs.[7]

The Subtask 2.2 EPC cost was developed from the Subtask 2.1 and Subtask 1.3 Next Plant EPC costs by subtracting the cost of the hydrogen production and compression facilities, and then adding the cost of the F-T hydrocarbon synthesis area and adjusting the cost of the power block. because it is now contains only one gas turbine and HRSG. No adjustments were made to the costs of the solids handling and ASU areas. The cost of the gasification block was adjusted to account for the cost of a smaller incinerator and the removal of one of the two syngas moisturizers. Adjustments also were made to the balance of plant area as appropriate.

The cost of the F-T area was estimated from the processing equipment sizes using an appropriate installation factor that was developed from previous cost estimates for similar facilities. The estimated cost of the large F-T slurry-bed hydrocarbon synthesis reactor is over 60% of the total equipment cost in the F-T area, and consequently, it dominates the cost of this area. Until wider experience is obtained with the construction of these large reactors, their estimated cost cannot have a high degree of accuracy.

[7] All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

Table III.3

Atmospheric Emissions Summary* of the Subtask 1.3 Next Plant, Subtask 2.1 and Subtask 2.2 Petroleum Coke IGCC Coproduction Plants

	Subtask 1.3 Next Optimixed Pet Coke IGCC Coproduction Plant	**Subtask 2.1** Coke Gasification Power Plant with Liquid Fuels	**Subtask 2.2** Optimized Coke Gasification Plant with Liquid Fuels
Total Gas Turbine Emissions			
Number of GT/HTSG Trains	2	1	1
GT/HTSG Stack Exhaust Flow Rate, lb/hr	7,967,000	7,966,800	4,231,000
GT/HTSG Stack Exhaust Temperature, ^{o}F	258	265	237
Emissions:			
SOx, ppmvd	3	2	<1
SOx as SO_2, lb/hr	50	41	<1
NOx, ppmvd (at 15% oxygen, dry basis)	10	10	10
NOx as NO_2, lb/hr	136	136	93
CO, ppmvd	10	10	10
CO, lb/hr	65	65	36
Incinerator Emissions			
Stack Exhaust Flow Rate, lb/hr	658,800+	21,670	23,450
Stack Exhaust Temperature, ^{o}F	500	500	500
Emissions:			
SOx, ppmvd	280	5,948	5,656
SOx as SO_2, lb/hr	300	280	276
NOx, ppmvd (at 15% oxygen, dry basis)	40	40	40
NOx as NO_2, lb/hr	31	1	1
CO, ppmvd	50	50	50
CO, lb/hr	24	1	1
Total Plant Emissions			
Exhaust Flow Rate, lb/hr	8,625,800+	7,988,470	4,254,450
Emissions:			
SOx, ppmvd	22	21	34
SOx as SO_2, lb/hr	350	321	276
NOx, ppmvd (at 15% oxygen, dry basis)	14	13	16
NOx as NO_2, lb/hr	166	136	94
CO, ppmvd	13	10	10
CO, lb/hr	89	66	37
VOC and Particulates, lb/hr	NIL	NIL	NIL
Opacity	0	0	0
Sulfur Removal, %	99.4	99.5	99.6

* Expected emissions performance
+ Includes PSA tail gas

The accuracy of the total installed cost for the Subtask 1.3 Next Optimized Coke IGCC Coproduction Plant was estimated to be on the order of ±11%. This level of accuracy reflects a high degree of confidence based on the large number of vendor quotes that were obtained and that the power block costs are based on a current similar Gulf Coast power project. This accuracy applies only to the total plant cost and does not apply to the individual areas or parts.

The accuracy of the total installed cost for the Subtask 2.1 Optimized Coke Gasification Power Plant with Liquids Coproduction is not as good. The estimated cost of the F-T area is only an order of magnitude cost estimate (nominally ±30%) because of the manner in which it was developed. Thus, the over estimate accuracy for the Subtask 2.1 plant probably is in the ±15% range. Because the cost of the F-T area of the Subtask 2.2 plant is a larger portion of the plant cost, the accuracy of the Subtask is less, and probably about ±20%.

Table III.4

Capital Cost Summary for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction, Subtask 2.2 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction, and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

Plant Area	Subtask 2.2 Optimized Power and Liquids Plant	Subtask 2.1 [Non-optimized] Power and Liquids Plant	Subtask 1.3 Next Plant (Coproduction Plant)
Solids Handling	8,012,000	8,012,000	8,012,000
Air Separation Unit	107,246,000	107,246,000	107,246,000
Gasification	300,288,000	312,591,000	312,591,000
Hydrogen Production	0	0	42,931,000
F-T Liquids Area	72,368,000	34,270,000	0
Power Block	178,631,000	276,414,000	237,045,000
Balance of Plant	68,748,613	79,420,000	79,420,000
Total	735,294,000	817,953,000	787,245,000

Notes:
1 Because of rounding, the columns may not add to the total that is shown.
2. All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

Section 4

Availability Analysis

In all the previous Task 1 cases, an availability analysis was used to determine the daily average production rates for calculating the annual production rates and cash flow. This analysis showed that the inclusion of a spare gasification train in the Subtask 1.3 Next Plant was a worthwhile addition that increased the return on investment although it increased the plant cost.

The common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period, all are dependent on the project cash flow. The net cash flow is the sum of all project revenues and expenses. Depending upon the detail of the financial analysis, the cash flow streams usually are computed on 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. Therefore, the annual production rate is a key parameter in determining the financial performance of a project.

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.[1] 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 three adjustments, this data was used to estimate the availability of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction under the two operating scenarios, with and without backup power purchases. The first adjustment increased the availability of the air separation plant from the observed availability of 96.32% to the industry average availability of 98%. The second adjusted the availability of the first gasification stage to remove a slag tap plugging problem caused by an unexpected change in the coal blend to the gasifier. This adjustment is justified since a dedicated petroleum coke plant would be very unlikely to experience this problem. The third eliminated a short outage that occurred in the water treatment facility because this plant will have sufficient treated water storage to handle this type of outage.

Using the EPRI recommended procedure, availability estimates were calculated for the two operating scenarios of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction.[8] Table IV.1 defines the plant configurations for the Subtask 2.2, Subtask 2.1, and Subtask 1.3 Next Plant designs. This table shows that the three plants have similar configurations except for byproduct production. All three plants contain a spare gasification train. Because the Low Temperature Heat Recovery/Acid Gas Removal (LTHR/AGR) area and Sulfur Recovery Unit (SRU) are highly reliable, these facilities are not spared. However, they only contain sufficient capacity so that only two of the three parallel gasification trains can be operated simultaneously. The only configuration difference between the Subtask 2.2 and the Subtask 2.1

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

plants is that the Subtask 2.2 plant contains only one combustion turbine whereas the Subtask 2.1 plant contains two turbines. However, the F-T area of the Subtask 2.2 optimized plant is over twice as large as that of the Subtask 2.1 non-optimized plant.

For each plant, the potential syngas availability from two of the three gasification trains at full design capacity is 86.5% of the time, and from only one of the two trains, it is 99.63%. The equivalent syngas availability is 93.24% of the design capacity.

Recent data presented at the 2002 Gasification Technologies Council conference by Clifton Keeler show further reliability improvements in the on-stream performance of the Wabash River Repowering Project.[9] However, the following availability and financial analyses will be based on the data reported in the final repowering project report for consistency with the Task 1 results. This will cause the following results to be somewhat conservative, but they will be consistent with the previously reported Task 1 results.

Because of the nature of the Subtask 2.2 plant with two parallel gasification trains, one F-T hydrocarbon synthesis train, and one combustion turbine power train, there are six possible operating modes for the plant as shown in Table IV.2. Cases A, B and C have both gasification trains available for operation. In Case A, both the F-T area and the power train are available and operating. In Case B, the gas turbine power train is unavailable, and in Case C, the F-T area is unavailable. Cases D, E and F are similar to Cases A, B and C, respectively, except that only one of the two gasification trains is available for operation.

Case A is normal operation with both gasification trains, the F-T hydrocarbon synthesis area and the power block all operating at capacity. Case D has only one gasification train operating causing the F-T hydrocarbon area and the power block to operate at reduced capacity.

Because the GE7FA+e combustion turbine required modifications to use the low Btu content fuel, it is not longer able to operate with pure syngas in the situation when the F-T area is not available for operation. Therefore, in Cases C and F, the entire plant is shut down when the F-T area is unavailable for processing syngas. During these periods of non-operation, it was assumed that the plant consumed 2 MW of import “hotel” power to maintain the facilities.

In Cases B and E, the power block is unavailable while the F-T area is available and at least one gasification train is available. Thus, there are two possible operating scenarios for these two cases. In the first scenario, the objective is to maximize F-T liquids production, and sufficient backup power is purchased to operate the plant. In the second scenario, the cost of purchasing backup power is expensive, and only the minimum amount of imported “hotel” power is purchased to maintain the facilities.

Table IV.2 shows the equipment status for of the Subtask 2.2 optimized plant for each of the two operating scenarios; the Maximum F-T Liquids Case (with backup power purchase) and the Minimum Power Purchase Case. Also shown in this table is the feed that each piece of equipment would be using in each situation when it is operating. In addition, below the case identification is shown the expected annual percentage of time the gasification and power blocks would be operating under these conditions. For example, in Case A under both scenarios, two gasification trains would be operating at design capacity on coke, the combustion turbine would

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

be operating at design capacity on syngas and F-T fuel gas (unconverted syngas and light hydrocarbons), and the F-T liquefaction would be operating at design capacity on syngas.

Table IV.3 shows the Subtask 2.2 design and annual average feed and product rates for both operating scenarios, the design and Base Case rates for the Subtask 2.1 non-optimum case, and the design and average rates for Subtask 1.3 Next Plant. At design conditions, the Subtask 2.2 Optimized Coke Gasification Plant with Liquids Coproduction consumes 5,417 tpd of dry petroleum coke and 110.6 tpd of flux, and produces 10,450 bpd of F-T liquids, 366.9 MW of export power, 373.4 tpd of sulfur, and 195.1 tpd of slag (containing 15% water). Compared to the Subtask 2.1 non-optimized design, the Subtask 2.2 case produces 6,325 more bpd of F-T liquids and 250 MW less of export power. The sulfur and slag product rates for the Subtask 2.2 case are slightly more than the Subtask 2.1 case reflecting the slightly increased coke and flux rates.

In the Maximum F-T Liquids Case with backup power purchase, the Subtask 2.2 plant has a daily average coke consumption of 4,984 tpd of dry coke and produces 9,702 bpd of F-T liquids and 316.4 MW of export power. The average F-T liquids production is 92.8% of the design capacity. The average export power production is only 86.2% of the design rate reflecting the amount of import power that is purchased when the power block is unavailable.

In the Minimum Power Purchase Case, the Subtask 2.2 plant has a daily average coke consumption of 4,776 tpd of dry coke and produces 9,267 bpd of F-T liquids and 320.4 MW of export power. The average F-T liquids production is 88.7% of the design capacity. The average export power production is 87.3% of the design rate reflecting the amount of lower amount of import power that is purchased when the power block is unavailable.

Table IV.1

Plant Configurations and Sections Capacities of the Subtasks 2.1 and 2.2 Coke Gasification Power Plants with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

Case Identification	**Subtask 2.2**	**Subtask 2.1**	**Subtask 1.3**
Case Description	**Optimized Coke Power & Liquids Plant**	**Non-Optimized Coke Power & Liquids Plant**	**Next Plant**
	Number of Trains and Section Capacity (Note 1)		
Gasification & Power Blocks			
Air Separation Unit (ASU)	2x50	2x50	2x50
Coke Handling	1x100	1x100	1x100
Slurry Preparation	2x60	2x60	2x60
Slurry Feed	3x50	3x50	3x50
Gasification (though HTHRU)	3x50	3x50	3x50
Slag Handling	1x100	1x100	1x100
Dry Particulate Removal			
Cyclone	3x50	3x50	3x50
Particulate Filters	3x50	3x50	3x50
Wet Scrubbing System	2x50	2x50	2x50
LTHR/AGR	2x50	2x50	2x50
SRU	2x50	2x50	2x50
Combustion Turbine	1x100	2x50	2x50
Steam Turbine	1x100	1x100	1x100
Byproduct Production			
Hydrogen	----	----	2x50
Fischer-Tropsch Process Area	1x100	1x100	----
Number of Fischer-Tropsch Reactors	1	1	0
Scheduled Outages per Gasification Train	15.34%	15.34%	15.34%
Spare Gasifier Vessels (1 per train)	No	No	No
Potential Syngas Availability, % (note 2)			
From 2 of 3 Gasification Trains (@100% rate)	86.85%	86.85%	86.85%
From 1 of 3 Gasification Trains (@50% rate)	99.63%	99.63%	99.63%
Equivalent Availability (note 3)	93.24%	93.24%	93.24%

Notes:
1. Capacity percentages are based on the total plant design capacity.
2. This is the clean syngas availability (including scheduled outages) without any downstream constraints on the consumption or use of the syngas; e.g., when exporting syngas to a pipline.
3. Equivalent availability is the annual average capacity expressed as a fraction of the design capacity.

Table IV.2

Equipment Status of the
Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction
for the Two Operating Scenarios

Case		**Maximum F-T Liquids (with Backup Power Purchase)**	**Minimum Power Purchase**
% of Time+	Equipment	Feed or Status	Feed or Status
A 82.13%	Gasifier 1 Gasifier 2 F-T Liquids Area Gas Turbine	Coke Coke Syngas Syngas & F-T Gas	Coke Coke Syngas Syngas & F-T Gas
B 3.85%	Gasifier 1 Gasifier 2 F-T Liquids Area Gas Turbine	Coke* Coke* Syngas Not Available	Coke* Coke* Available - Not Operating Not Available
C 0.83%	Gasifier 1 Gasifier 2 F-T Liquids Area Gas Turbine	Available - Not Operating Available - Not Operating Not Availavle Available - Not Operating	Available - Not Operating Available - Not Operating Not Available Available - Not Operating
D 12.09%	Gasifier 1 Gasifier 2 F-T Liquids Area Gas Turbine	Coke Not Available Syngas* Syngas & F-T Gas*	Coke Not Available Syngas* Syngas & F-T Gas*
E 0.57%	Gasifier 1 Gasifier 2 F-T Liquids Area Gas Turbine	Coke Not Available Syngas* Not Available	Available - Not Operating Not Available Available - Not Operating Not Available
F 0.12%	Gasifier 1 Gasifier 2 F-T Liquids Area Gas Turbine	Coke Not Available Not Available Available - Not Operating	Available - Not Operating Not Available Not Available Available - Not Operating

* Operations at reduced capacity

Table IV.3

Design and Daily Average Feed and Product Rates for the Subtask 2.2, Subtask 2.1 and Subtask 1.3 Next Plant Cases

Subtask 2.2 Optimum Design

		Daily Average Rates	
Case	Design	Maximum F-T Liquids (with backup power purchase)	Minimum Power Purchase
Feed Rates			
Coke, TPD dry	5,417	4,984	4,776
Flux, TPD	110.6	101.8	97.5
Natural Gas, MMBtu/d	0	0	0
Product Rates			
Export Power, MW	366.9	316.4	320.4
Sulfur, TPD	373.4	343.6	329.2
Slag, TPD	195.1	179.5	172.0
F-T Liquids, bpd	10,450	9,702	9,267

Subtask 2.1 Non-Optimum Design

Case	Design	Base Case Daily Average
Feed Rates		
Coke, TPD dry	5,375	4,805
Flux, TPD	109.7	98.1
Natural Gas, MMBtu/d	553	369
Product Rates		
Export Power, MW	617.0	572.5
Sulfur, TPD	370.6	331.3
Slag, TPD	193.6	173.1
F-T Liquids, bpd	4,125	3,983

Subtask 1.3 Next Plant

Case	Design	Daily Average
Feed Rates		
Coke, TPD dry	5,417	4,842
Flux, TPD	110.6	98.9
Natural Gas, MMBtu/d	0	9,059
Product Rates		
Export Power, MW	474.0	448.4
Steam, Mlb/hr	980.0	974.6
Hydrogen, MMscfd	80.0	78.8
Sulfur, TPD	373.4	333.8
Slag, TPD	195.1	174.4
F-T Liquids, bpd	0	0

Section 5

Financial Analysis

The following financial analysis was performed using the discounted cash flow (DCF) model that was developed by Bechtel Technology and Consulting (now Nexant Inc.) for the DOE as part of the Integrated Gasification Combined Cycle (IGCC) Economic and Capital Budgeting Practices Task.[10] This model calculates investment decision criteria used by industrial end-users and project developers to evaluate the economic feasibility of IGCC projects.

5.1 Financial Model Input Data

The required input information to the DCF financial model is organized into two distinct input areas that are called the Plant Input Sheet and the Scenario Input Sheet. The Plant Input Sheet contains data that are directly related to the specific plant as follows.

Data Contained on the Plant Input Sheet

- Project summary information
- Plant output and operating data
- Capital costs
- Operating costs and expenses

The Scenario Input Sheet contains data that are related to the general economic environment that is associated with the plant as well as some data that are plant related. The data on the Scenario Input Sheet are shown below.

Data Contained on the Scenario Input Sheet

- Financial and economic data
- Fuel data
- Tariff assumptions
- Construction schedule data
- Startup information

For all cases, the EPC spending pattern was adjusted to reflect forward escalation during the construction period since the EPC cost estimate is an "overnight" cost estimate based on mid-year 2000 costs.

Items that were excluded in the cost estimate, such as spares, owner's cost, contingency and risk are included in the financial analysis.

Table V.1 summarizes the basic input parameters to the financial model. The daily average plant input and output flow rates are given in Table IV.3

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

Table V.1

Basic Financial Model Input Parameters

Parameter	Value
Financial Parameters	
Owner's Contingency (% of EPC Cost)	5.0%
Development Fee (% of EPC Cost)	1.23%
Start-up Cost (% of EPC Cost)	1.50%
Additional Financing Cost, EPC Contingency, Risk and Fees, etc.	5.0%
Loan Amount (% of Cost)	80%
Loan Interest Rate	10% & 8%
Loan Financing Fee	3.0%
Loan Repayment Term, years	15
Income Tax Rate	40%
Construction Period, years	15
Start Up	
First Year's Average Capacity	60%
Prices	
Coke, $/dry ton	0.00
Flux, $/ton	5.00
Natural Gas, $/MMBtu HHV or $/Mscf *	2.60
Fischer-Tropsch Liquids, $/bbl	30.00
Electric Power, $/MW	27.00
Sulfur, $/ton	30.00
Slag, $/ton (15% water)	0.00
Annual Inflation Rates	
Coke $/dry ton	1.2%
Natural Gas, $/HHV MMBtu	3.9%
Fischer-Tropsch Liquids	3.1%
Electric Power, $/MW	1.7%
Sulfur, $/ton	0.0%
Slag, $/ton (15% water)	0.0%

* Natural gas is assumed to have a HHV Btu content of 1,000 Btu/scf.

5.2 Financial Model Results

Figure 5.1 shows the return on investment (ROI) as a function of the export power price for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction under both operating scenarios and compares them with the Base Case of the Subtask 2.1 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction. This figure shows that for all cases, the Subtask 2.2 Maximum F-T Liquids with backup power purchase operating scenario has a higher Return on Investment (ROI) than the Minimum Power Purchase operating scenario at all power prices with the F-T liquids at 30 $/bbl. At low F-T liquids prices and high power prices, the situation will reverse. However, such situations do not appear realistic, and therefore, all future financial analyses of Subtask 2.2 plant will use the Maximum F-T Liquids operating scenario.

Compared to the Subtask 2.1 [non-optimized] Base Case, both Subtask 2.2 cases have higher returns on investment when the power prices are below about 43 to 45 $/MW-hr and the F-T liquids selling price is 30 $/bbl. The relationship between the F-T liquids price and the power selling price will be explored in more detail later in this section.

Figure 5.1

Return on Investment vs. Power Price for the Subtask 2.2 [Optimized] and the Subtask 2.1 [Non-optimized] Coke Gasification Power Plants with Liquids Coproduction

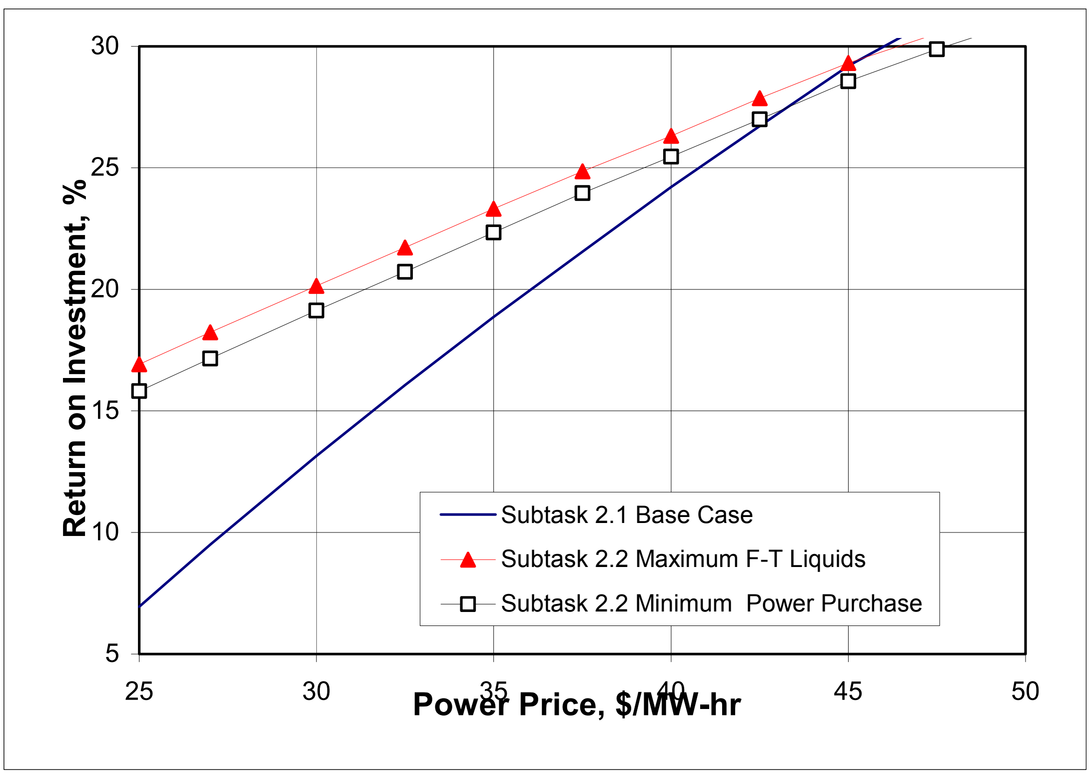


Table V.2 compares the two Subtask 2.2 operating scenarios with the Subtask 2.1 Base Case scenario and Subtask 1.3 Next Plant. At the basic economic conditions shown in Table V.1 (at a 10% loan interest rate), the Subtask 2.2 Maximum F-T Liquids Case with backup power purchase has a 18.24% return on investment. This is over 1% higher than the ROI of the Subtask 2.2 Minimum Power Purchase Case, and over 8% greater than the ROI for either the Subtask 2.1 Base Case or the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. The Subtask 2.2 Maximum F-T Liquids Case also requires a lower power selling price for a 12% ROI than the Minimum Power Purchase Case.

With an 8% loan interest rate, all four cases have higher ROIs by about 3.5%. However, their relative ranking remains the same. The Subtask 2.2 Maximum F-T Liquids Case with backup power purchase still has the best ROI followed by the Subtask 2.2 Minimum Power Purchases Case, the Subtask 2.1 Base Case, and the Subtask 1.3 Next Plant Case. The relative ranking of the required selling prices of the power and F-T liquids for the four cases also are the same.

Figure 5.2 shows the ROI as a function of the export power price for the Subtask 2.2 and 2.1 Optimized and [Non-optimized] Coke Gasification Power Plants with Liquids Coproduction and compares them with the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. The solid lines are with a 10% loan interest rate. At a 27.0 $/MW-hr export power price, the Subtask 2.2 Maximum F-T Liquids Case has a 18.24% ROI, the Subtask 2.1 Base Case has a 9.50% ROI, and the Subtask 1.3 Next Plant has a 9.05% ROI. Not shown on the figure, the Subtask 2.2 Maximum F-T Liquids Case has a 12.0% ROI at an export power price of 17.71 $/MW-hr which is significantly less than the Subtask 2.1 Base Case, which has a 12.0% ROI at an export power price of 29.04 $/MW-hr, and the Subtask 3.1 Next Plant, which has a 12.0% ROI at a power price of 30.02 $/MW-hr.

An 8% loan interest rate increases the return on investment by about 3.7 ROI percent as shown by the open symbols on the figure. At an 8% loan interest rate, and a 27 $/MW-hr power selling price, the Subtask 2.2 optimized plant has a 21.81% ROI, the Subtask 2.1 non-optimized plant has a 13.24% ROI, and the Subtask 1.3 Next Plant has a 12.70% ROI.

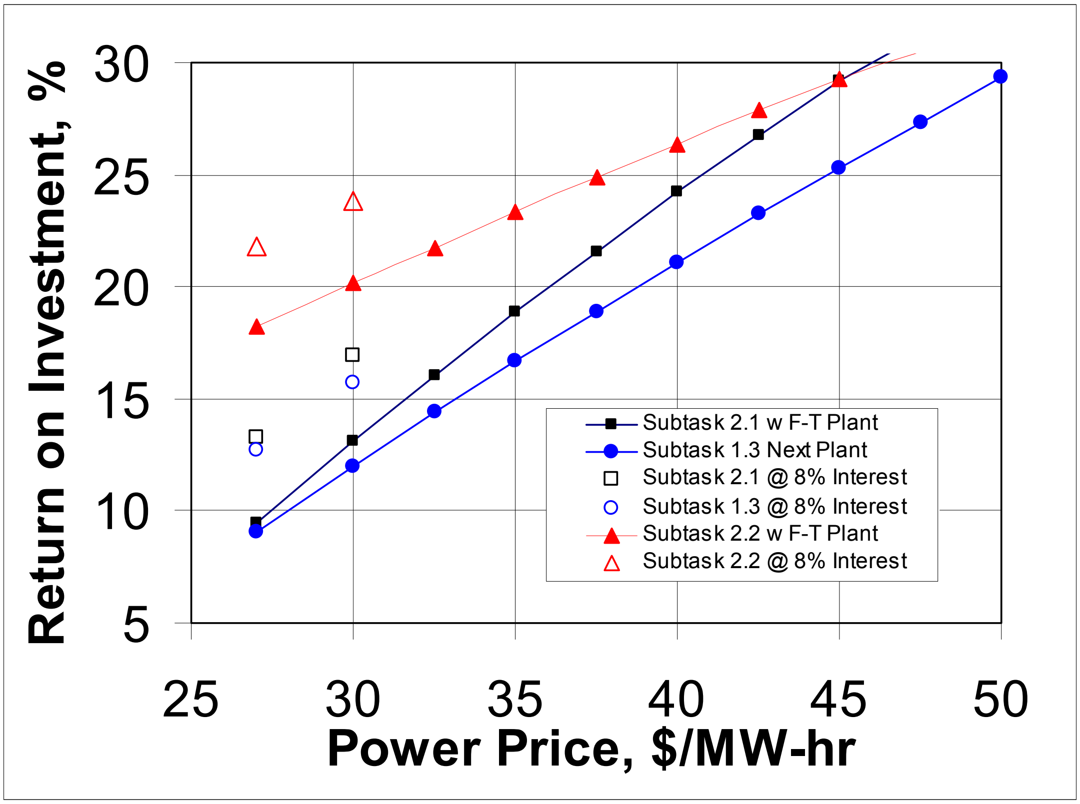
It is difficult to predict the future value of either power, natural gas and/or the F-T liquid fuel precursors. The liquid fuel precursors price is related to the crude oil price which also can be highly variable both because of market forces and the influence of international politics. Various studies have been made which attempt to relate the value of the F-T liquids to that of crude oil by replacing crude oil in the refinery feed stream with the F-T liquids. The resulting values for the F-T liquids generally are above the crude oil values, but the specific amount can range from 2 $/bbl up to 10 $/bbl depending upon the refinery configuration, the specific crude oils being replaced, and the required refinery product mix.[11]

Figure 5.3 shows the effect of the liquid fuel precursors selling price on the return on investment versus the power selling price for the Subtask 2.2 Maximum F-T Liquids Case with a 10% loan interest rate. The solid 30 $/bbl line is the same line as shown on the previous figure for the Subtask 2.2 plant. The dashed line is the corresponding 30 $/bbl line for the Subtask 2.1 Base Case.

[11] Marano, J. J., Rogers, S., Choi, G. N., and Kramer, S. J., "Product Valuation of Fischer-Tropsch Derived Fuels," ACS National Meeting, Washington, D. C., August 21-6, 1994.

Figure 5.2

Return on Investment vs. Power Price for the Subtask 2.2 and 2.1 Optimized and [Non-optimized] Coke Gasification Power Plants with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant

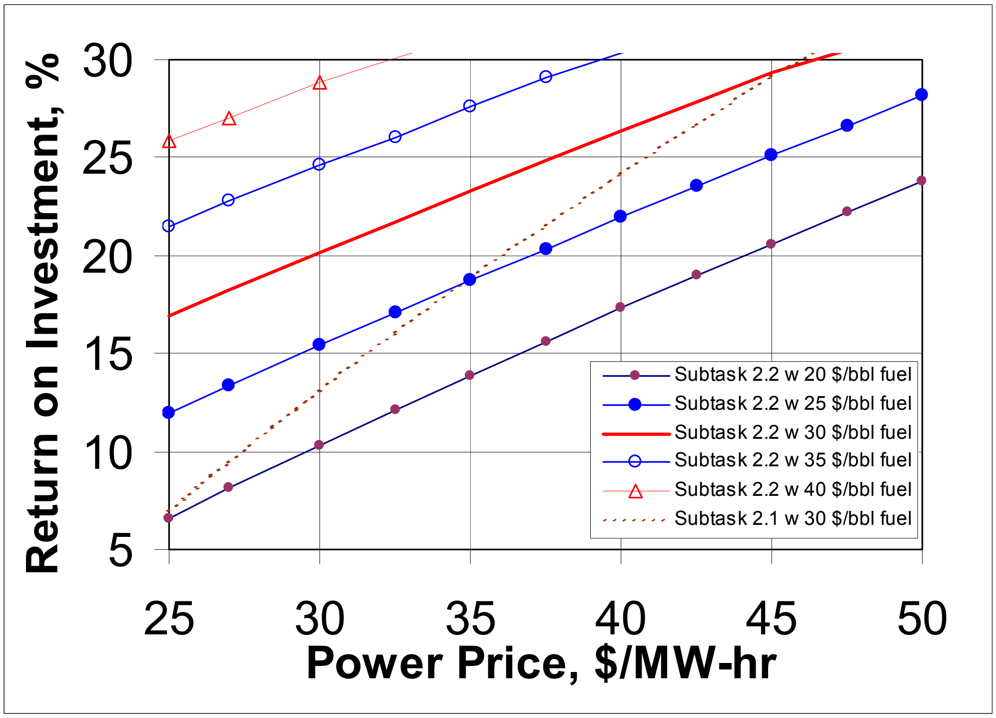


At a 27 $/MW-hr power selling price, for the Subtask 2.2 Maximum F-T Liquids operating scenario each 5 $/bbl change in the F-T liquids selling price changes the return on investment by 4.2 to 5.2 ROI percent. This compares with about a 2.0 ROI percent change for the Subtask 2.1 Base Case. As the power price increases, the effect of the liquids selling price decreases because that portion of the revenue generated from the liquid fuel sales becomes a smaller portion of the total plant revenue.

The ROI for the Subtask 2.1 plant has a greater slope versus power price than that of the Subtask 2.2 plant because the revenue generated from the power sales is a significantly larger portion of the total plant revenue. As such, any change in the power price will have a larger influence on the ROI.

Figure 5.3

Return on Investment vs. Power Price Showing the Effect of the Liquids Price for the Subtask 2.2 Optimized and Subtask 2.1 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction

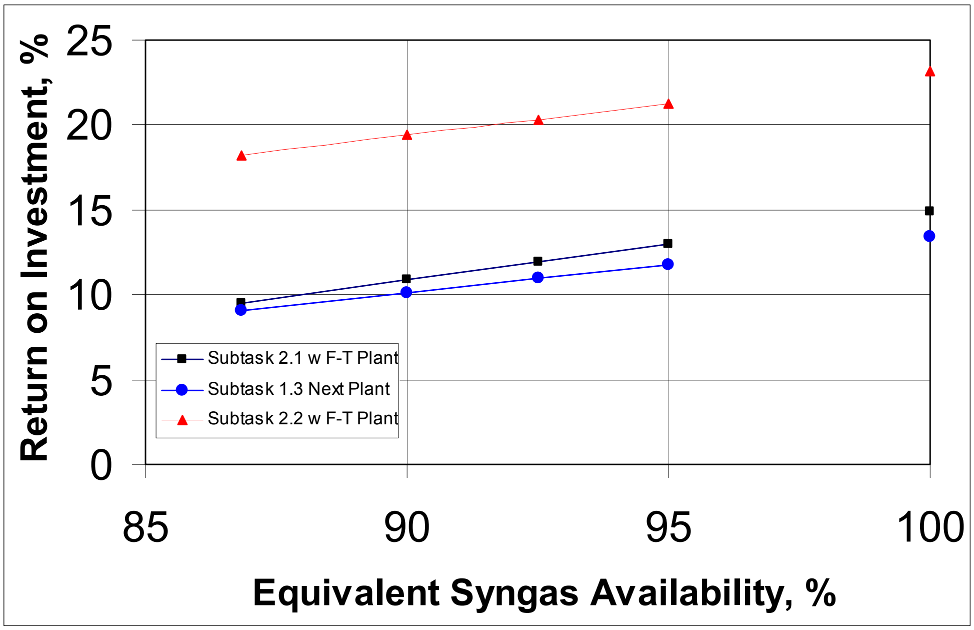


After commissioning all plants undergo a “learning curve” during which problem areas are corrected, inadequate equipment is modified or replaced, and adjustments are made. Consequently, performance improves as measured by increased capacity and/or improved on-stream factors. Figure 5.4 shows the effect of improved syngas availability on the return on investment. For the Subtask 2.2 plant, as the syngas availability improves, the amount of backup power that has to be purchased is reduced until it disappears at the unattainable 100% syngas availability. For the Subtask 2.1 plant, the amount of purchased natural gas decreases in a similar manner as the syngas availability improves. At the expected 86.85% syngas availability, the Subtask 2.2 Maximum F-T Liquids Case has an ROI of 18.24%, the Subtask 2.1 Base Case has a ROI of 9.50%, and the Subtask 1.3 Next Plant has a ROI of 9.05%. At 90% syngas availability, the ROI of the Subtask 2.2 plant increases to about 19.41%, that of the Subtask 2.1 plant increases to about 10.8%, and that of the Subtask 1.3 Next Plant increases to 10.1%. At the unattainable syngas availability of 100%, the Subtask 2.2 Maximum F-T Liquids Case will have an expected ROI of 23.13%, the Subtask 2.1 Base Case will have an expected ROI of 14.9%, and the Subtask 1.3 Next Plant will have an expected 13.4% ROI.

Table V.3 shows the sensitivity of some individual component prices and financial parameters for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction (Maximum F-T Liquids Case) starting from a 12% ROI (with a power price of 25.021 $/MW-hr and a 25.0 $/bbl liquids price). Each item was varied individually without affecting any other item. Most sensitivities are based on a ±10% change from the base value except when either a larger or smaller change is used because it either makes more sense or it is needed to show a meaningful result.

Figure 5.4

Return on Investment vs. Syngas Availability for the Subtask 2.2 Optimized and Subtask 2.1 [Non-optimized] Coke Gasification Power Plants with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant



The F-T liquids selling price is the most sensitive product price with a 10% increase to 27.5 $/bbl resulting in a 2.52% increase in the ROI to 14.52%, and a 10% decrease resulting in a 2.61% decrease in the ROI to 9.39%. A 10% increase in the electric power price to 27.52 $/MW-hr will cause a 1.75% increase in the ROI to 13.75%, and a 10% decrease in the power price to 22.52 $/MW-hr will result in a 1.76% decrease in the ROI to 10.24%. In the Subtask 2.1 Base Case, a 10% change in the power selling price had a greater sensitivity than a 10% change in the F-T Liquids price because most of the revenue came from power sales. However, in the Subtask 2.2 Maximum F-T Liquids Case, the situation is reversed and the F-T liquids price has a greater effect because most of the revenue now comes from the sale of the liquids.

Changes in the sulfur and slag prices only have a small influence on the ROI.

All the above economic cases were developed with a long-term coke netback price of zero at the refinery gate; i.e., the revenue obtained from the sale of the coke is the same as the expense of transporting it to a site where it is consumed. A decrease in the coke price of 5 $/ton to a negative 5.0 $/ton will increase the ROI by 1.96% to 13.96%, and a 5 $/ton increase in the coal price will lower the ROI by 1.95% to 10.05%.

A 5% decrease in the plant EPC cost from 735.3 MM$ to 689.5 MM$ will increase the ROI by 2.26% to 14.26%, and a 5% increase in the plant cost to 788 MM$ will decrease the ROI by 2.09% to 9.91%.

In today's unsettled financial situation, the loan interest rate and project financing conditions also can be uncertain. A 20% decrease in the loan interest rate to 8% from the base interest rate of 10% will increase the ROI to 15.57% from 12.00%, and a 20% increase in the interest rate to 12% will lower the ROI by 3.60% to 8.40%. A 20% decrease in the loan amount from 80% to 72% will lower the ROI by 0.55% to 11.45%, and a 20% increase in the loan amount to 88% will increase the ROI by 0.90% to 12.90%. Decreasing the income tax rate by 10% from 40% to 36% will increase the ROI by 0.48% to 12.48%, and a 10% increase in the tax rate to 44% will lower the ROI by 0.52% to 11.48%.

Influence of Product Prices on Design Selection

Figures 5.5, 5.5 and 5.7 show the return on investment for the Subtask 2.2 Maximum F-T Liquids Case, the Subtask 2.1 Base Case, and the Subtask 1.3 Next Plant as a function of the F-T liquids selling price at a 2.60 $/MMBtu natural gas price and at three different power selling prices. Figure 5.5 is with a 27 $/MW-hr power selling price; Figure 5.6 is with a 30 $/MW-hr power selling price; and Figure 5.7 is with a 40 $/MW-hr power selling price. Figure 5.5 shows that the Subtask 1.3 Next Plant has the highest ROI below a F-T liquids price of 20.9 $/bbl, and above that the Subtask 2.2 Maximum F-T Liquids Case has the highest ROI. Figure 5.6 shows similar behavior. With a 30 $/MW-hr power price, the Subtask 1.3 Next Plant has the highest ROI below a F-T liquids price of 21.5 $/bbl, and above this price, the Subtask 2.2 Maximum F-T Liquids Case has the highest ROI.

Figure 5.7 shows a different situation when the power selling price is 40 $/MW-hr. Below a F-T liquids price of 20.8 $/bbl, the Subtask 1.3 Next Plant has the highest ROI. Between a F-T liquids selling prices of 20.8 and 26.0 $/bbl, the Subtask 2.1 Base Case has the highest ROI, and above a F-T liquids price of 26.0 $/bbl, the Subtask 2.2 Maximum F-T Liquids Case has the highest ROI.

These cases show that the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction has robust economics when the F-T liquid products are worth 25 $/bb or above. At lower liquid values it may be more advantageous to install a Subtask 1.3 Next Plant type facility that co-produces hydrogen and steam in addition to power when there is a demand for the co-products.

Figure 5.5
Return on Investment vs. F-T Liquids Price
(With 27 $/MW-hr Power and 2.60 $/MMBtu Natural Gas)

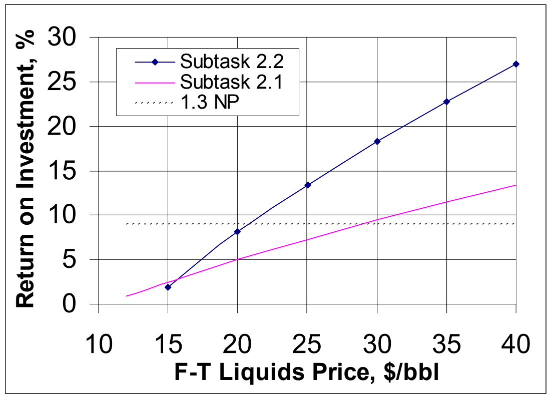


Figure 5.6
Return on Investment vs. F-T Liquids Price
(With 30 $/MW-hr Power and 2.60 $/MMBtu Natural Gas)

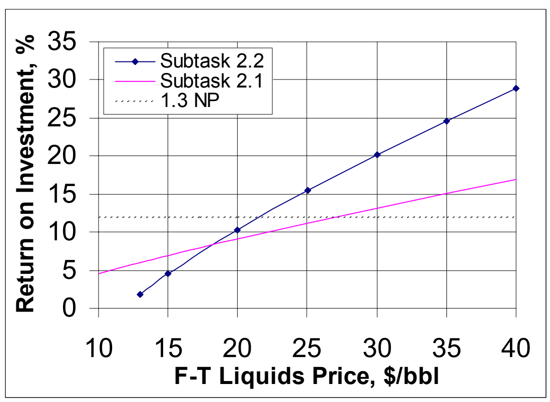


Figure 5.7
Return on Investment vs. F-T Liquids Price
(With 40 $/MW-hr Power and 2.60 $/MMBtu Natural Gas)

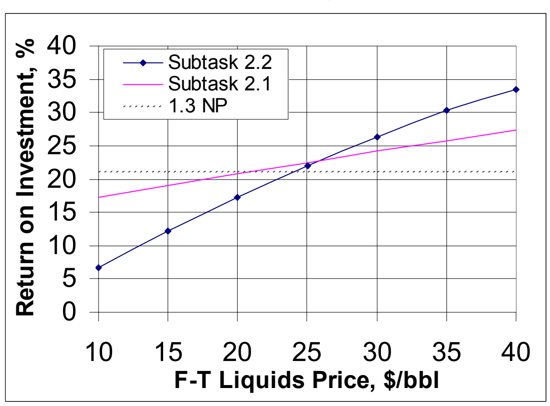


Table V.2

Return on Investments and Required Product Selling Prices for the Two Subtask 2.2 Operating Scenarios, Subtask 2.1 Base Case, and the Subtask 1.3 Next Plant (with a Natural Gas Price of 2.60 $/MMBtu)

	Subtask 2.1 Base Case	Subtask 2.2 Maximum F-T Liquids Case	Subtask 2.2 Minimum Power Case	Subtask 1.3 Next Plant
With a 10% Loan Interest Rate				
Return on Investment with 27 $/MW-hr Power and 30 $/bbl Liquids	9.50%	18.24%	17.16%	9.05%
Required Selling Price for a 12% ROI of				
Power with 30 $/bbl Liquids, $/MW-hr	29.04	17.71	19.47	30.02
Liquds with 27 $/MW-hr Power, $/bbl	36.22	23.65	24.54	---
With a 8% Loan Interest Rate				
Return on Investment with 27 $/MW-hr Power and 30 $/bbl Liquids	13.24%	21.81%	20.72%	12.70%
Required Selling Price for a 12% ROI of				
Power with 30 $/bbl Liquids, $/MW-hr	26.04	12.81	14.64	26.32
Liquds with 27 $/MW-hr Power, $/bbl	27.05	20.30	21.03	---

Table V.3

Sensitivity of Individual Component Prices on the Return on Investment For the Subtask 2.2 Optimized Coke to Liquids and Power Case from a 12% ROI (at a Liquids Price of 25 $/bbl and a Power Price of 25.021 $/MW-hr)

	Decrease			Base	Increase		
	ROI	Value	% Change	Value	% Change	Value	ROI
Products							
F-T Liquids	9.39%	22.5 $/bbl	-10%	**25.0 $/bbl**	+10%	27.5 $/bbl	14.52%
Power	10.24%	22.52 $/MW-hr	-10%	**25.021 $/MW-hr**	+10%	27.52 $/MW-hr	13.75%
Sulfur	11.92%	27.0 $/ton	-10%	**30.0 $/ton**	+10%	33.0 $/ton	12.08%
Slag	11.93%	-5.0 $/ton	---	**0 $/ton**	---	5.0 $/ton	12.07%
Feeds							
Coke	13.96%	-5.0 $/ton	---	**0 $/ton**	---	5.0 $/ton	10.05%
Flux	12.05%	0 $/ton	100%	**5.0 $/ton**	+100%	10.0 $/ton	11.95%
Financial							
EPC Cost	13.11%	716.9 MM$	-2.5%	**735.3 MM$**	+2.5%	769.4 MM$	10.94%
EPC Cost	14.26%	689.5 MM$	-5.0%	**735.3 MM$**	+5.0%	788.1 MM$	9.91%
Interest Rate	15.57%	8%	-20%	**10%**	+20%	12%	8.40%
Loan Amount	11.45%	72%	-20%	**80%**	+20%	88%	12.90%
Tax Rate	12.48%	36%	10%	**40%**	+10%	44%	11.48%

Note: Products and Feeds each are listed in decreasing sensitivity.

Section 6

Summary

A design for an optimized coke gasification power plant with liquid fuel precursors coproduction using Fischer-Tropsch technology has been developed. The plant consumes 5,417 tpd of coke (dry basis) and 111 tpd of flux to produce 367 MW of export power and 10,450 bpd of liquid fuel precursors. It also produces 373 tpd of elemental sulfur and 195 tpd of slag. The plant is located on the U.S. Gulf Coast adjacent to a petroleum refinery.

The design of this Subtask 2.2 optimized plant was developed from those of the Subtask 2.1 [Non-optimized] Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.3 Next Optimized Petroleum Coke IGCC Coproduction Plant. Each of these plants process about the same amount of coke and produce about the same amount of slag and byproduct sulfur. However, the Subtask 2.1 produces 4,125 bpd of liquid fuel precursors and 617 MW of export power whereas the Subtask 1.3 Next Plant produces 80 MMscfd of 99% hydrogen and 980,000 lb/hr of 750°F/700 psia steam for the adjacent petroleum refinery in addition to 474 MW of export power.

The Subtask 2.2 optimized design was developed from the Subtask 2.1 non-optimized design by enlarging and optimizing the F-T liquids production facilities. These changes increased the production of the F-T liquid fuel precursors from 4,125 bpd to 10,450 bpd at the expense of the production of export power, which was reduced from 617 MW to 367 MW. In contrast to the non-optimized Subtask 2.1 plant, the optimized Subtask 2.2 plant does not consume any natural gas during normal operations.

The Fischer-Tropsch hydrocarbon synthesis area basically consists of three sections; final sulfur removal, slurry-bed F-T reactor, and product recovery sections. The final sulfur removal section consists of three activated carbon beds in series, which adsorb the residual sulfur in the syngas. The activated carbon beds are regenerated with medium-pressure steam. This is a much simpler and less costly design then that of the non-optimized plant which contains a hydrolysis reactor followed by a non-regenerable ZnO adsorbent.

The sulfur-free syngas is fed to the slurry-bed F-T reactor which converts it to hydrocarbons over an iron-based catalyst. The heat of reaction is removed by generation of 440°F/375 psia steam inside tubes that are placed within the slurry-bed. The lighter hydrocarbon products and unconverted syngas leave the reactor as vapors and are cooled by refrigeration to condense and recover the hydrocarbons as liquids. The unconverted syngas and non-condensable light hydrocarbons (primarily C1 through C3s) are compressed, moisturized, and sent to the power block. The heavier products are removed from the reactor as liquids, separated from the entrained catalyst by filtration, cooled, mixed with the lighter hydrocarbons, and sent to the adjacent petroleum refinery for separation, upgrading and incorporation into liquid transportation fuels.

The F-T liquid fuel precursors essentially are a bottomless, sulfur-free crude oil. Basically they are straight-chain 1-olefins and paraffins without any aromatics. The diesel fraction has a very high cetane number (>70) and is a premium blending component for diesel fuel. The naphtha fraction is a low octane material that requires further upgrading for use as a gasoline blending component. However, it is an excellent feedstock for an ethylene cracker. Linear programming studies have shown that the F-T liquid fuel precursors may be worth up to 10 $/bbl more than crude oil depending upon the specific refinery configuration and product demands.

The combined-cycle power block includes one GE7FAe+ combustion turbine, one heat recovery steam generator (HRSG), and a non-reheat steam turbine. The combustion turbine fuel is a mixture of F-T off gas and about 8% of the available syngas, which bypasses the F-T synthesis reactor. The F-T off gas contains a significant quantity of CO_2, and the mixed fuel gas has a heat content of 147 Btu (LHV)/scf. This is less than General Electric's minimum specification of 200 Btu/scf before diluent addition, but General Electric believes this gas can be used in the turbine (after suitable burner testing). The HRSG is a three pressure boiler with most of the surface area in the economizer and the steam superheater. The F-T and gasification units generate over 2 million pounds per hour of high-pressure and medium-pressure steam. This, combined with limited amount of energy in the gas turbine exhaust, led to selection of a non-reheat steam cycle. The gross power output of the combined-cycle system is 474 MW (199 MW from the gas turbine and 275 MW from the steam turbine) resulting in 367 MW of net export power.

The Subtask 2.2 optimized plant has a LHV thermal efficiency of 55.1% and an HHV thermal efficiency of 56.7%, both of which are based on the heating value of the F-T liquids, the byproduct sulfur and the equivalent energy of the export power. These efficiencies are 7 to 9% greater than those of the Subtask 2.1 non-optimized plant, and about 12 to 14% greater than those of the Subtask 1.5B coke IGCC power plant.

The Subtask 2.2 Optimized Coke Power Plant with Liquids Coproduction produces about 2½ times as much liquids and about 60% as much export power at a lower EPC cost (35 MM versus 818 MM mid-year 2000$ dollars) than the Subtask 2.1 non-optimized plant. When power prices are low and the liquid fuel precursors are worth 25 $/bbl or more, the Subtask 2.2 optimized plant has a substantially better return on investment than the Subtask 2.1 plant. However, at low liquid fuel prices, the Subtask 1.3 Next Plant, which co-produces hydrogen and steam with power, may have a higher return where there is a demand for the co-products.

The improvements generated during the development of this Subtask 2.2 optimized design will be applied to the Subtask 2.3 design which will use coal rather than petroleum coke as the gasifier feedstock.

Appendix A - Attachment

Subtask 2.2

Major Equipment List

Appendix A

Major Equipment List

Table A1 lists the major pieces of equipment and systems by process area in the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. Detailed equipment lists for systems that would be purchased as complete units from a single vendor, such as the Air Separation Unit, are not available.

Table A1
Major Equipment of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction

Fuel Handling – 100
Coke Storage Dome
Reclaim Conveyors
Storage/Feed Bins
Coke Handling Electrical Equipment and Distribution
Electric Hoist
Metal Detector
Magnetic Separator
Flux Silo
Vibrating Feeder
Slurry Preparation – 150
Weigh Belt Feeder
Rod Charger
Rod Mill
Rod Mill Product Tank
Rod Mill Product Tank Agitator
Rod Mill Product Pumps
Recycle Water Storage Tank
Recycle Water Pumps
Slurry Storage Tank
Slurry Storage Tank Agitator
Slurry Recirculation Pumps
Solids Recycle Tank
Solids Recycle Tank Agitator
Solids Recycle Pumps
Rod Mill Lube Oil Pumps
Slurry Feed Pumps (1st Stage)
Slurry Feed Pumps (2nd Stage)
Fischer-Tropsch – 200 and 201
First Activated Carbon Adsorption Bed
Second Activated Carbon Adsorption Bed
Third Activated Carbon Adsorption Bed
F-T Reactor Steam Drum
3-Phase Overhead Flash Drum
Unconverted Syngas Wash Column
Sour Water Flash Drum
Wax L/V Separator
Wax Vapor L/V Separator
Liquid Fuel L/V Separator

Wax Mixer Surge Drum
Catalyst Slurry Mixing Tank
Catalyst Pretreater
Pretreated Catalyst Feed Tank
Catalyst Pretreater Overhead KO Drum
First F-T Reactor Vapor Overhead Flash Drum
Refrigeration KO Drum
Refrigeration Liquid Receiver
IP Steam Flash Drum
LP Steam Flash Drum
Reactor Vapor / BFW Exchanger
Reactor Vapor Water Cooler
HP Fuel Gas Steam Heater
Wax Vapor Air Cooler
Liquid Wax / BFW Exchanger
Liquid Fuel Water Cooler
Wax Cooler Heat Exchanger
Catalyst Pretreater Wax Heater
Cat Pretreater Feed/Effluent Exchanger
Catalyst Pretreater Overhead Cooler
Catalyst Pretreater Circulating Gas Heater
Refrigerant System Condenser
HP Fuel Gas Refrigeration Recovery Exchanger
Reactor Vapor Refrigeration Cooler
F-T Reactor Feed Preheater
Second Gas Turbine Fuel Heater (MP Steam)
Fuel Gas Compressor
Cat Pretreater Circulating Gas Compressor
Refrigeration Compressor
Liquid Fuel Recycle Pumps
BFW Circulation Pumps
Liquid & Catalyst Return Pumps
Wax Pumps to Filter
Wax Product Pumps
Clean Wax Pumps
Wax Recovery Pumps
Pretreated Cat to Reactor Pumps
Syngas Wash Tower Recirculation Pumps
Storage Tank Pumps
F-T Slurry Bed Reactor
Overhead Vapor Cyclone
Liquid Catalyst Hydroclone
Liquid Catalyst Cleanup Filter
Wax Catalyst Filters
Makeup Catalyst Feed Hopper Baghouse
Makeup Catalyst Feed Hopper
Catalyst Pretreater Baghouse
F-T Product Storage Tank
ASU & Gasifier Area Cooling Water - 250
Cooling Water Circulation Pumps
Cooling Tower (S/C)

Gasification - 300
Main Slurry Mixers
Second Stage Mixer
Gasifier Vessel
High Temperature Heat Recovery Unit (HTRU)
Cyclone Separators
Slag Pre-Crushers
Slag Crushers
Reactor Nozzle Cooling Pumps
Crusher Seal Water Pumps
Syngas Desuperheater
Nitrogen Heater
Pressure Reduction Units
Dry Char Filters
Cyclone Solids Pickup Vessel
Filter Solids Pickup Vessel
Slag Handling – 350
Slag Dewatering Bins
Slag Gravity Settler
Slag Water Tank
Slag Water Pumps
Gravity Settler Bottoms Pumps
Slag Recycle Water Tank
Slag Feedwater Quench Pumps
Slag Water Recirculation Pumps
Polymer Pumps
Slag Recycle Water Cooler
LTHR/AGR – 400
Syngas Scrubber Column
Syngas Scrubber Recycle Pumps
Syngas Recycle Compressor
Syngas Recycle Compressor Knock Out Drum
Syngas Heater
COS Hydrolysis Unit
Amine Reboiler
Sour Water Condenser
Sour Gas Condensate Condenser
Sour Gas CTW Condenser
Sour Water Level Control Drum
Sour Water Receiver
Sour Gas Knock Out Pot
Sour Water Carbon Filter
MDEA Storage Tank
Lean Amine Pumps
Acid Gas Absorber
MDEA Cross-Exchangers
MDEA CTW Coolers
MDEA Carbon Bed
MDEA Post-Filter
Acid Gas Stripper
Acid Gas Stripper Recirculation Cooler
Acid Gas Stripper Reflux Drum
Acid Gas Stripper Quench Pumps
Acid Gas Stripper Reboiler
Acid Gas Stripper Overhead Filter

Lean MDEA Transfer Pumps
Acid Gas Stripper Knock Out Drum
Acid Gas Stripper Preheater
Amine Reclaim Unit
Condensate Degassing Column
Degassing Column Bottoms Cooler
Sour Water Transfer Pumps
Ammonia Stripper
Ammonia Stripper Bottoms Cooler
Stripped Water Transfer Pumps
Quench Column
Quench Column Bottoms Cooler
Stripped Water Transfer Pumps
Degassing Column Reboiler
Ammonia Stripper Reboiler
Syngas Heater
Syngas Moisturizer
Moisturizer Recirculation Pumps
Sulfur Recovery – 420
Reaction Furnace/Waste Heat Boiler
Condensate Flash Drum
Sulfur Storage Tank
Storage Tank Heaters
Sulfur Pump
Claus First Stage Reactor
Claus First Stage Heater
Claus First Stage Condenser
Claus Second Stage Reactor
Claus Second Stage Heater
Claus Second Stage Condenser
Condensate Level Drum
Hydrogenation Gas Heater
Hydrogenation Reactor
Quench Column
Quench Column Pumps
Quench Column Cooler
Quench Strainer
Quench Filter
Tail Gas Recycle Compressor
Tail Gas Recycle Compressor Intercooler
Tank Vent Blower
Tank Vent Combustion Air Blower
Tank Vent Incinerator/Waste Heat Boiler
Tank Vent Incinerator Stack
Gas Turbine / HRSG – 500
Gas Turbine Generator (GTG), GE 7FA+e, Dual Fuel (Gas and Syngas) Industrial turbine set, Including: Lube Oil Console, Static Frequency Converter, Intake Air Filter, Compressor, Turbine Expander, Generator Exciter, Mark V Control System, Generator Control Panel and Fuel skids.
GTG Erection (S/C)
Heat Recovery Steam Generator (HRSG) - Dual Pressure, Unfired, with Integral Deaerator
HRSG Stack (S/C)
HRSG Continuous Emissions Monitoring Equipment

HRSG Feedwater Pumps
HRSG Blowdown Flash Tank
HRSG Atmospheric Flash Tank
HRSG Oxygen Scavenger Chemicals Injection Skid
HRSG pH Control Chemicals Injection Skid
GTG Iso-phase Bus Duct
GTG Synch Breaker
Power Block Auxilary Power XformerS
Steam Turbine Generator & Auxiliaries - 600
Steam Turbine Generator (STG), Non-reheat, complete with lube oil console
Steam Surface Condenser, 316L tubes
Condensate (hotwell) pumps
Circulating Water Pumps
Auxiliary Cooling Water Pumps
Cooling Tower
ASU – 800
Air Separation Unit Including:
Main Air Compressor
Air Scrubber
Oxygen Compressor
Cold Box (Main Exchanger)
Oxygen Compressor / Expander
Liquid Nitrogen Storage
Balance Of Plant - 900
High Voltage Electrical Switch Yard (S/C)
Common Onsite Electrical and I/C Distribution
Distributed Control System (DCS)
In-Plant Communication System
15KV, 5KV and 600V Switchgear
BOP Electrical Devices
Power Transformers
Motor Control Centers
Makeup Pumps
Substation & Motor Control Center (MCC)
Lighting, Heating & Ventilation
Makeup Water Treatment Storage and Distribution
Water Treatment Building Equipment
Carbon Filters
Cation Demineralizer Skids
Degasifiers
Anion Demineralizer Skids
Demineralizer Polishing Bed Skids
Bulk Acid Tank
Acid Transfer Pumps
Demineralizer - Acid Day Tank Skid
Bulk Caustic Tank Skid
Caustic Transfer Pumps
Demineralizer - Caustic Day Tank Skid
Firewater Pump Skids
Waste Water Collection and Treatment
Oily Waste - API Separator
Oily Waste - Dissolved Air Flotation
Oily Waste Storage Tank
Sanitary Sewage Treatment Plant
Wastewater Storage Tanks

Monitoring Equipment
Common Mechanical Systems
Shop Fabricated Tanks
Miscellaneous Horizontal Pumps
Auxiliary Boiler
Safety Shower System
Flare
Flare Knock Out Drum
Flare Knock Out Drum Pumps
Chemical Feed Pumps
Chemical Storage Tanks
Chemical Storage Equipment
Laboratory Equipment

The Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction facility is assumed to be located adjacent to a petroleum refinery, and thus, can share some infrastructure with the refinery. It is assumed that

1. The refinery delivers the coke to the coke storage dome.
2. The plant gets the river water from the refinery water intake system.
3. The refinery processes the process waste water from the plant through the refinery waste water treatment facilities.

Appendix C

Subtask 2.3

Optimized Coal Gasification

Power Plant with Liquids Coproduction

Subtask 2.3

Executive Summary

This report describes an Optimized Coal Gasification Power Plant with Liquids Coproduction using Fischer-Tropsch technology. The plant consumes 9,266 tpd of coal (dry basis) to produce 675.9 MW of export power and 12,377 bpd of liquid fuel precursors. It also produces 237 tpd of elemental sulfur and 1,423 tpd of slag. The plant is located in the U.S. Midwest adjacent to a suitable water source and in reasonable proximity to a petroleum refinery which can upgrade the liquid fuel precursors into transportation fuels.

The design of this Subtask 2.3 optimized plant was developed from those of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant. Starting from the Subtask 1.6 plant, two Fischer-Tropsch hydrocarbon production trains replaced once combined-cycle train (containing two gas turbines, two heat recovery steam generators (HRSGs), and one steam turbine). The Subtask 2.2 optimized F-T coproduction plant design provided a basis for integration of the Subtask 2.3 plant. In contrast to the Subtask 2.2 plant, the Subtask 2.3 plant purchases backup natural gas, instead of power, to increase availability.

Each of the two parallel Fischer-Tropsch hydrocarbon synthesis trains basically consists of three sections; final sulfur removal, slurry-bed F-T reactor, and product recovery sections. The final sulfur removal section contains three activated carbon beds in series, which adsorb the residual sulfur in the syngas. The activated carbon beds are regenerated with medium-pressure steam.

Sulfur-free syngas is fed to the slurry-bed F-T reactors which convert it to hydrocarbons over an iron-based catalyst. The heat of reaction is removed by generation of 440°F/375 psia steam inside tubes that are placed within the slurry-bed. The lighter hydrocarbon products and unconverted syngas leave the reactor as vapors and are cooled by refrigeration to condense and recover the hydrocarbons as liquids. The unconverted syngas and non-condensable light hydrocarbons (primarily C1 through C3s) are compressed, mixed with bypass syngas, moisturized, and sent to the power block. The heavier products are removed from the reactor as liquids, separated from the entrained catalyst by filtration, cooled, mixed with the lighter hydrocarbons, and sent to a petroleum refinery for separation, upgrading and incorporation into liquid transportation fuels.

The F-T liquid fuel precursors essentially are a bottomless, sulfur-free crude oil. Basically they are straight-chain 1-olefins and paraffins without any aromatics. The diesel fraction has a very high cetane number (>70) and is a premium blending component for diesel fuel. The naphtha fraction requires further upgrading for use as a gasoline blending component. However, it is an excellent feedstock for an ethylene cracker. Linear programming studies have shown that the F-T liquid fuel precursors may be worth up to 10 $/bbl more than crude oil depending upon the specific refinery configuration and product demands.

The combined cycle power block contains two GE7FAe+ combustion turbines, two HRSGs, and a reheat steam turbine. The combustion turbine fuel is a mixture of F-T off gas and syngas, which bypasses the F-T synthesis reactors. The net power output of the plant is 675.9 MW.

The Subtask 2.3 optimized coal plant has a LHV thermal efficiency of 53.2% and an HHV thermal efficiency of 53.4%, both of which are based on the heating value of the F-T liquids, the byproduct sulfur, and the equivalent energy of the export power. These efficiencies are about 11 to 12% greater than those of the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant. The thermal efficiencies of the optimized Subtask 2.2 coke coproduction plant are about 2 to 3% higher than those of the optimized coal plant.

The Subtask 2.3 Optimized Coal Power Plant with Liquids Coproduction produces 12,377 bpd of liquids at the expense of about 479 MW less power production compared to Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant. With 30 $/bbl liquids, the return on investment of the Subtask 2.3 plant exceeds 10% only when power prices are above 40 $/MW-hr. However, at these power prices, the Subtask 1.6 IGCC power plant has a higher return on investment. Therefore, the opportunity for a domestic coal based gasification power plant with liquid fuel precursors coproduction appears to be limited in today's economic environment unless there are special circumstances, such as the use of a low priced feedstock. However, future improvements in gasification area availability from future design enhancements, advanced gas turbines developments, and improved Fischer-Tropsch reactor performance could make a coal gasification power plant with liquid fuel precursors coproduction economically competitive in the future.

Subtask 2.3
Table of Contents

Attachment

Section 1

Introduction

The objective of this Gasification Plant Cost and Performance Optimization Project is to develop optimized engineering designs and costs for several Integrated Gasification Combined Cycle (IGCC) plant configurations. These optimized IGCC plant systems build on the commercial demonstration cost data and operational experience from the Wabash River Coal Gasification Repowering Project.[1] The Wabash River Repowering Project consists of a nominal 2,500 TPD E-GAS™ gasifier producing clean syngas for a GE 7A gas turbine and steam for repowering an existing steam turbine.

Task 1 of this IGCC Plant Cost and Performance Optimization study consists of the following nine subtasks:

- Subtask 1.1 – Expand the Wabash River Project facility design to a greenfield unit
- Subtask 1.2 – Petroleum Coke based IGCC plant with the coproduction of steam and hydrogen
- Subtask 1.3 – Optimized petroleum coke based IGCC plant with the coproduction of steam and hydrogen
- Subtask 1.4 – Optimized coal to power IGCC plant
- Subtask 1.5 – Comparison between single-train coal and coke fueled IGCC power plants
- Subtask 1.6 – Optimized coal fueled 1,000 MW IGCC power plant
- Subtask 1.7 – Optimized single-train coal to hydrogen plant
- Subtask 1.8 – Review the status of warm gas clean-up technology applicable to IGCC plants
- Subtask 1.9 – Discuss the Value Improving Practices availability and reliability optimization program

Task 1 has been completed. The Task 1 Topical Report was issued to the Department of Energy on May 30, 2002.[2]

Task 2 has the objectives of developing optimum plant configurations for IGCC power plants with the coproduction of liquid fuel precursors. Task 2 is divided into the three subtasks.

Subtask 2.1 – [Non-Optimum] Petroleum Coke Gasification Power Plant with Liquids Coproduction Starting with the same petroleum coke and gasification plant designs generated in the Subtask 1.3 Next Optimized Coke IGCC Coproduction Plant, a design shall be developed for a coke gasification power plant co-producing liquid transportation fuel precursors containing a single-train, once-through Fischer-Tropsch (F-T) gas-to-liquids (GTL) plant. The liquid hydrocarbons from the F-T hydrocarbon synthesis section will be recovered and sent to the adjacent petroleum refinery for upgrading and blending into premium liquid transportation fuels. The unconverted syngas and non-condensable hydrocarbons from the F-T hydrocarbon synthesis section will be used for power production in the combined-cycle power block.

[1] "Wabash River Coal Gasification Repowering Project, Final Technical Report", U. S. Department of Energy, Contract Agreement DE-FC21-92MC29310, August 2000.
[2] "Task 1 Topical Report – IGCC Plant Cost Optimization", Gasification Plant Cost and Performance Optimization, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, May 30, 2002.

Subtask 2.2 – Optimum Petroleum Coke Gasification Power Plant with Liquids Coproduction The Subtask 2.1 plant shall be optimized to develop an optimized coke gasification plant co-producing liquid transportation fuel precursors. Optimization activities primarily will be concerned with the F-T area and overall plant integration. Since the Subtask 2.1 gasification area was developed from an optimized IGCC petroleum coke gasification coproduction plant, a review of the plant is appropriate at this time to ensure that the previous modifications are still applicable to this case.

Subtask 2.3 - An Optimized Coal Gasification Power Plant with Liquids Coproduction The Subtask 2.2 plant shall be converted to a coal-fueled gasification unit using Illinois No. 6 coal. This will involve combining the optimized coal gasification plant developed in Subtask 1.6 with the Subtask 2.2 plant, to develop an optimized coal gasification power plant co-producing liquid transportation fuel precursors. Because of differences in the syngas generation area and resulting syngas composition, the F-T hydrocarbon synthesis area, and overall plant integration required some modification.

This report describes the design, performance and economics of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. Section 2 provides background information on two previous studies that are the basis for this current study. Section 2.1 briefly describes the Subtask 2.2 Optimized Petroleum Coke Gasification Power Plant with Liquids Coproduction. Section 2.2 describes previous work on Fischer-Tropsch Synthesis. Section 2.3 briefly describes the Subtask 1.6 Optimized Nominal 1,000 MW Coal IGCC Power Plant.

Section 3 described the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction that was developed in this study and compares it with the Subtask 1.6 optimized nominal 1,000 MW coal plant. Section 4 provides an availability analysis. Section 5 contains a financial analysis of plant performance and compares it with the previously developed Subtask 2.2 plant and the Subtask 1.6 plant.

An attachment contains a listing of the major equipment in the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction.

Section 2

Background

During Task 1, several designs were developed for coal IGCC power plants. Subtask 1.6 developed a design and cost estimate for a nominal 1,000 MW coal IGCC power plant located in the Midwest based on the Subtask 1.1 Wabash River Greenfield Plant and the Subtask 1.3 optimized petroleum coke IGCC coproduction plants. Section 2.1 describes the Subtask 1.6 IGCC Plant.

In 1991, Bechtel with Amoco as the main subcontractor was awarded DOE contract DE-AC22-91PC90027 to develop designs and computer process simulation models for indirect coal liquefaction plants using advanced Fischer-Tropsch Technology.[3] This is discussed further in Section 2.2. Subsequently, the simulation model was improved by adding additional components.[4] The Fischer-Tropsch (F-T) hydrocarbon synthesis section of this ASPEN process simulation model was used to develop the design of the F-T hydrocarbon synthesis section of the Subtask 2.1 Coke Gasification Power Plant with Liquids Coproduction. Subtask 2.2 used the optimization techniques developed in Task 1 to optimize the Subtask 2.1 design. Section 2.3 briefly describes the Subtask 2.2 Optimized Petroleum Coke Gasification Power Plant with Liquids Coproduction

2.1 Subtask 1.6 Nominal 1,000 MW IGCC Power Plant

The Subtask 1.6 plant design was optimized by applying nine Value Improving Practices (VIPs) to the Subtask 1.1 plant and by expanding the plant to four trains to reduce costs and improve operability.[5] As a result of this effort, plant performance was improved, the plant cost was reduced, and the return on investment was significantly improved. The results of this VIP and optimization study included:

- Simplified solids handling system
- Removal of the feed heaters and spare pumps
- Maximum use of slurry quench
- Maximum syngas moisturization
- Use of a cyclone and a dry particulate removal system to clean the syngas
- Smaller T-120 post reactor residence vessel
- Simplified Claus plant, amine and sour water stripper
- Use of state-of-the-art GE 7FA+e gas turbines with 210 MW output and lower NOx
- Use of steam diluent in the gas turbines

[3] "Topical Report – Volume I, Process Design – Illinois No. 6 Coal Case with Conventional Refining", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.
"Topical Report – Volume IV, Process Flowsheet (PFS) Models", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, October, 1994.
[4] "Topical Report VI – Natural Gas Fischer-Tropsch Case, Volume II, Plant Design and Aspen Process Simulation Model", Baseline Design/Economics for Advance Fischer-Tropsch Technology, U. S. Department of Energy, Contract Number DE-AC22-91PC90027, August, 1996.
[5] "Task 1 Topical Report – IGCC Plant Cost Optimization", Gasification Plant Cost and Performance Optimization, Chapter II, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, May 30, 2002

- Development of a compact plant layout to minimize the use of large bore piping
- Used Bechtel's advanced construction techniques to reduce costs
- Added design features to reduce O&M costs and increase syngas availability

Table II.1 shows the design input and output streams for the Subtask 1.6 IGCC Plant. The plant consumes 9,266 tpd of dry coal and produces 1,154.6 MW of export power. It also produces 237 tpd of sulfur and 1,423 tpd of slag. No natural gas is consumed during design operations. However, the plant does use natural gas during startup and as a supplementary fuel to fire the combustion turbines when insufficient syngas is available.

The resulting design configuration for the Subtask 1.6 IGCC Plant is shown in Figure 2.1. The plant basically is a four-train gasification facility, without a spare gasification train or spare gasification capacity. There are also four gas turbines in two combined cycle trains.

The Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant has an overall LHV thermal efficiency of 42.4% and a HHV thermal efficiency of 40.8%, both of which include the heat content of the byproduct sulfur.

Table II.1

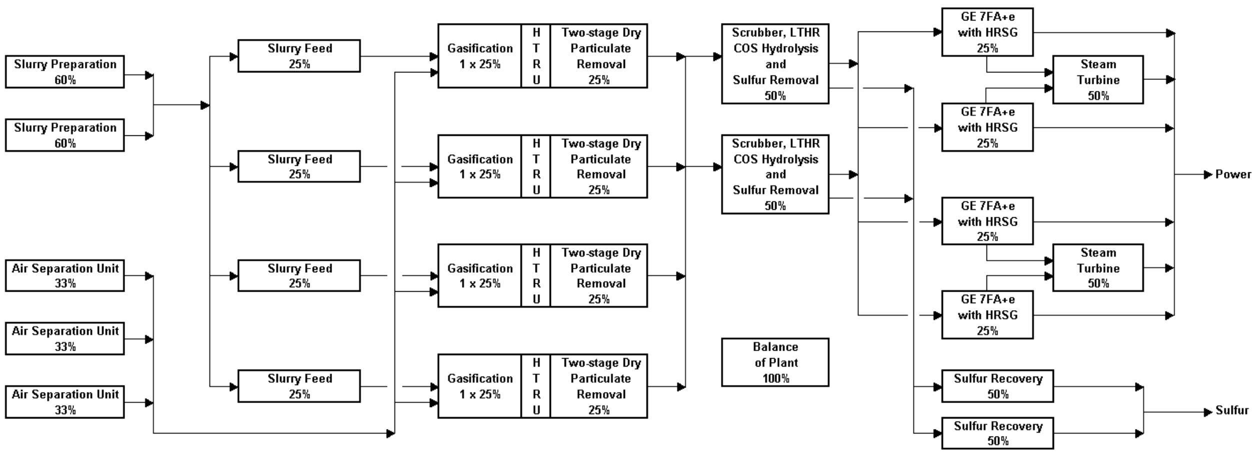
Design Input and Output Streams for the Subtask 1.6 Nominal 1,000 MW Optimized Coal IGCC Plant

	Subtask 1.6 1,000 MW Plant
Plant Inputs	
Coal Feed, as received TPD	10,837
Coal Feed to Gasifiers, TPD	9,266
Oxygen Production, TPD of 95% O_2	8,009
Total Fresh Water Consumption, gpm	9,752
Plant Outputs	
Net Power Output, MW	1,154.6
Sulfur, TPD	237
Slag, TPD (15% moisture)	1,423

Figure 2.1

Subtask 1.6 - Train Diagram

Nominal 1,000 MW Coal IGCC Power Plant



Note: Capacity percentages are based on total plant capacity.

2.2 Fischer Tropsch Hydrocarbon Synthesis Process

The Fischer-Tropsch hydrocarbon synthesis process is an old process in which synthesis gas or syngas (carbon monoxide and hydrogen) react over a catalyst to produce aliphatic hydrocarbons (principally normal paraffins and straight chain 1-olefins). It was used by Germany during the Second World War to make liquid fuels for military use. Subsequent cost reductions may have made F-T processes competitive in certain situations. Currently, there is a lot of interest in using the F-T process to monetize remote natural gas by converting it into an easily transportable synthetic crude oil that can be upgraded to liquid transportation fuels.

In general, the F-T hydrocarbon synthesis reactions for olefins and normal paraffins can be written as

$$n\,CO + 2n\,H_2 \rightarrow C_nH_{2n} + n\,H_2O$$

$$n\,CO + (2n+2)\,H_2 \rightarrow H\text{-}(CH_2)_n\text{-}H + n\,H_2O$$

As seen from the above reaction stoichiometry, the ideal syngas composition is just over 2 moles of hydrogen for each mole of carbon monoxide.

The reaction is very exothermic. Traditionally, at a large scale the reaction has been performed over solid catalyst that is placed in small diameter tubes immersed in a cooling medium (such as boiling water) to remove the heat of reaction. The hydrocarbon product yield distribution can be characterized by a Schultz-Flory distribution in which the molar ratio of a component containing n carbon atoms to one with n+1 carbon atoms is a constant called alpha (α). As the reaction temperature increases, the yield distribution shifts to lighter hydrocarbons; i.e., the α parameter gets smaller. As time has progressed, more sophisticated mathematical yield models using multiple α parameters have been developed to represent the F-T reaction yields.

In the 1950s, the slurry-bed reactor was developed in which fine catalyst particles are suspended in a liquid, and the reactant syngas is bubbled up through the catalyst/liquid mixture. Steam is generated within cooling coils immersed in the slurry-bed to remove the heat of reaction. This system has a high heat transfer rate resulting in a cheaper reactor with a higher productivity rate than catalyst particles packed in tubes. The lighter hydrocarbon products and unconverted syngas are withdrawn as vapor from the top of the reactor. Slurry is withdrawn from the reactor and pumped through a hydroclone and filter system which separates the clarified liquid products from the catalyst. The concentrated catalyst/slurry stream is returned to the reactor. A constant (steady-state) catalyst activity is maintained by continually withdrawing a small portion of catalyst from the reactor and replacing it with fresh catalyst.

Iron-based and promoted cobalt-based catalysts are the two primary catalysts currently used for F-T synthesis. Iron-based catalysts promote the water gas shift reaction which produces hydrogen from carbon monoxide and water; whereas cobalt catalysts generally do not. Therefore, for a syngas with a low hydrogen to carbon molar ratio, an iron based catalyst is preferred because it will produce hydrogen within the slurry-bed F-T synthesis reactor; whereas with a cobalt based catalyst, additional hydrogen has to be produced externally to the F-T synthesis reactor.

In the early 1990s, Bechtel developed several designs for indirect coal liquefaction plants using Fischer-Tropsch technology (references 3 and 4). Table II.2 shows the major input and output streams for the Baseline plant. The plant consumes 20,323 tpd of ROM Illinois No. 6 coal (8.6 wt% water) and 3,119 bpsd of normal butane to produce at total of 50,491 bpsd of petroleum products (1,921 bpsd of C3 LPG, 23,915 bpsd of gasoline, and 24,655 bpsd of distillate fuels). The plant is divided into three processing areas.

Area 100	Clean Syngas Production Area
Area 200	Fischer-Tropsch Synthesis Loop
Area 300	Product Upgrading and Refining Area

The Area 100 Clean Syngas Production Area grinds and dries the coal, gasifies the coal in six Shell gasifiers (five operating and one spare), scrubs and cleans the syngas, and recovers 46.69 Mlb/hr of sulfur for sale.

The Area 200 Fischer-Tropsch Synthesis Loop obtains a high CO conversion by recycling the unconverted syngas after recovering hydrogen and removing CO_2. The F-T hydrocarbon synthesis section contains 25 slurry-bed reactors (24 operating and one spare) arranged in eight parallel trains with each train having three reactors in parallel.

The Area 300 Product Upgrading and Refining Area essentially is a small refinery that upgrades the F-T products into liquid transportation fuels. It contains a saturated gas plant, C3/C4/C5 alkylation unit, C4 isomerization unit, C5/C6 isomerization unit, catalytic reformer, naphtha hydrotreater, distillate hydrotreater, and a wax hydrocracker. To increase the gasoline yield, normal butanes are purchased, isomerized to isobutene, and used to alkylate the C3, C4 and C5 olefins to make a high-octane gasoline blending component.

Table II.2

Design Input and Output Streams for the Baseline Indirect Coal Liquefaction Plant

Plant Inputs				
Illinois No. 6 ROM Coal*	1,693.6	Mlb/hr	20,323	TPD
Electric Power	54.36	MW		
Normal Butane	26.50	Mlb/hr	3,119	bpsd
Raw Water	10,042	gpm		
Plant Outputs				
C3 LPG	14.22	Mlb/hr	1,921	bpsd
Gasoline	251.44	Mlb/hr	23,915	bpsd
Diesel	278.21	Mlb/hr	24,655	bpsd
Sulfur	46.69	Mlb/hr		
Slag	187.03	Mlb/hr		

* As received coal containing 8.6 wt% water

2.3 Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction

The Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction was developed from the Subtask 2.1 non-optimized IGCC and liquids coproduction plant and the Subtask 1.3 optimized Next Petroleum Coke IGCC Coproduction Plant. These plants have been described previously.[6] Therefore, this section only provides a brief overview of these facilities.

The Subtask 1.3 Next Petroleum Coke IGCC Coproduction Plant consumed 5,417 tpd of dry petroleum coke and produced 474 MW of export power, 980 Mlb/hr of 750^{o}F/700 psig steam, 80 MMscfd of hydrogen, 373 tpd of sulfur, and 195 tpd of slag. The steam and hydrogen were sold to an adjacent petroleum refinery. This plant essentially is a two train facility with a spare 50% gasification block to increase syngas availability.

The Subtask 2.1 [non-optimized] Coke Gasification Power Plant with Liquids Coproduction replaced the export steam and hydrogen production facilities of the Subtask 1.3 plant with a single-train, once-through Fischer-Tropsch hydrocarbon synthesis area. It still contains two GE7FA+e combustion turbines and HRSGs for power production. As shown in Table II.3, this plant consumes 5,376 tpd of dry petroleum coke and produces 617 MW of export power, 4,124 bpd of hydrocarbon liquids, 371 tpd of sulfur, and 194 tpd of slag. Since this plant does not export any steam, all the steam production (from both the HRSG and F-T area) now is used for power production. The Subtask 2.1 plant also contains a spare 50% gasification block to increase syngas availability.

[6] Task 2 Progress Report – Subtask 2.2, An Optimized Coke Gasification Power Plant with Liquid Fuels Coproduction," Gasification Plant Costs and Performance Optimization, U. S. Department of Energy, Contract Number DE-AC26-99FT40342, Draft Report of May, 2003.

Table II.3

Design Input and Output Streams for the Subtask 2.1 and 2.2 Coke IGCC Power Plants with Liquids Coproduction

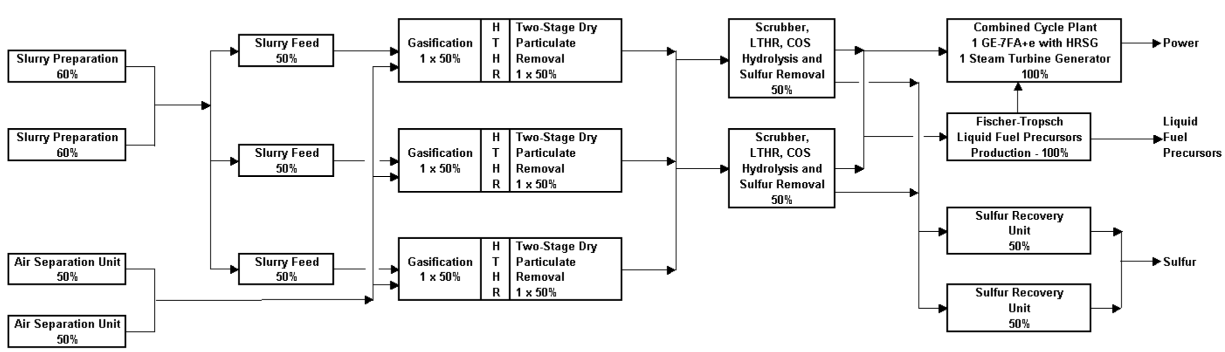
	Subtask 2.1 [Non-optimized] Coke IGCC Coproduction Plant	Subtask 2.2 Optimized Coke IGCC Coproduction Plant
Plant Inputs		
Coke Feed, as received TPD	5,649	5,684
Coke Feed to Gasifiers, TPD	5,376	5,417
Oxygen Production, TPD of 95% O_2	5,919	5,877
Total Fresh Water Consumption, gpm	6,472	5,693
Flux, TPD	109.7	110.6
Plant Outputs		
Net Power Output, MW	617.0	366.9
Liquid Fuel Precursors, bpd	4,125	10,450
Sulfur, TPD	371	373
Slag, TPD (15% moisture)	194	195

Starting from the Subtask 1.3 Next Plant, (and Subtask 2.1), the Subtask 2.2 plant was developed by eliminating one gas turbine along with the export steam and hydrogen production facilities and replacing them with a large single-train, once-through Fischer-Tropsch hydrocarbon synthesis area as shown schematically in Figure 2.2. The energy that was used to produce the export steam now is used to generate additional power. Even with almost the same coke feed rate to the gasifiers, the process required adjustments to the steam and water flows both in and between the gasification block and the power generation block, which was switched to a less efficient, non-reheat steam cycle in order to effectively use all the steam.

Figure 2.2

Subtask 2.2 - Train Diagram

Optimized Coke Gasification Power Plant with Liquid Fuels Coproduction



Notes: Capacity percentages are based on total plant capacity.

Table II.3 shows the design input and output streams for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. The plant consumes 5,417 tpd of dry petroleum coke, and produces 366.9 MW of export power and 10,450 bpd of liquid hydrocarbons. It also produces 373 tpd of sulfur and 195 tpd of slag. During periods when the plant produces insufficient power to satisfy its own internal demands, power is purchased to maintain the liquid hydrocarbon production. No natural gas is consumed during design operations. However, the plant does use natural gas during startup.

The Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction has an overall LHV thermal efficiency of 55.1% and a HHV thermal efficiency of 56.7%, both of which include the heat content of the byproduct sulfur. These efficiencies are significantly higher than those of the Subtask 2.1 non-optimized plant, which has a LHV efficiency of 47.8% and a HHV efficiency of 47.9%. This is because the liquid hydrocarbon product is a larger portion of the useable energy output of the optimized Subtask 2.2 plant than it is in the non-optimized Subtask 2.1 plant. The thermal efficiencies of both of these plants are significantly higher than those of the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant.

Section 3

Description of the Subtask 2.3 Optimized Coal Gasification Power Plant With Liquids Coproduction

The Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction was developed from the Subtask 1.6 Nominal 1,000 MW IGCC plant using the design approach adapted for the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction. The coal gasification capacity of the plant was kept the same as Subtask 1.6; i.e., that amount which could be processed in four gasification trains, to allow direct comparison of these two cases. However, the F-T liquids production was maximized, and power production was reduced to only one power block train consisting of two combustion turbines, two HRSGs, and a single steam turbine. The unconverted syngas and light hydrocarbons from the F-T synthesis section is compressed and combined with the small amount of syngas bypassing the F-T area to provide fuel for the two combustion turbines. Figure 3.1 shows the train configuration of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. As shown, Subtask 2.3 includes two parallel 50% F-T hydrocarbon synthesis trains.

Table III.1 compares the design input and output stream flows for the Subtask 2.3 and Subtask 1.6 Nominal 1,000 MW IGCC plant. From 9,266 tpd of dry coal, the plant produces 12,377 bpd of liquid fuel precursors, 675.9 MW of export power, and 237 tpd of sulfur. The LHV heating value of the liquid fuel product is 2,661 MMBtu/hr, or 780 MW_t. The export power production is reduced to 675.9 MW from 1154.7 MW for the Subtask 1.6 IGCC plant. Overall, the combined energy in the liquid fuel precursors and the electric power products (1,455 MW) is increased compared to Subtask 1.6 (1,154.7 MW) with the export power being only 46.5% of the plant output.

The F-T processing area of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction includes the following design improvements that were applied to Subtask 2.1:

- Use of regenerable activated carbon for final/trace sulfur removal
- Addition of refrigeration to increase the light oil recovery from the F-T area
- Replacement of the fired heater in the F-T catalyst preparation area with steam heating using high-pressure steam from the gasification block to eliminate the use of natural gas during normal operation
- Effective utilization of steam in the F-T area

Except for size, this design is the same as that used in Subtask 2.2.

On a lower heating value (LHV) basis, the plant has a thermal efficiency 53.2% when the heating value of the byproduct sulfur is included and 52.4% when the byproduct sulfur is not included. On a higher heating value (HHV) basis, the plant has a thermal efficiency 53.4% when the heating value of the byproduct sulfur is included and 52.6% when the byproduct sulfur is not included. These thermal efficiencies are higher than those that would be obtained from a coal IGCC power plant of a similar design because it includes the heating value of the liquid fuel that is produced. Since the second law of thermodynamics states this liquid fuel cannot be used at a 100% thermal efficiency, the thermal efficiency of the plant will be somewhat lower when the final disposition of the liquid fuel is considered.

The Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction has a LHV thermal efficiency of 55.1%. This is lower than that of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction (57.2% thermal efficiency) because the Subtask 2.3 plant produces less F-T fuel liquids (52.7% of the total energy in the products compared to 62.0% liquids for Subtask 2.2). Figure 3.2 shows the relationship between the LHV thermal efficiency for the three Task 2 plants versus the total LHV energy in the export products (liquids, power and sulfur). The three points with no liquids production represent Subtasks 1.5A (coal), 1.5B (coke), and Subtask 1.6 in order of increasing efficiency.

Liquids production for Subtask 2.3 was limited for two reasons:

- Gasification of coal in the E-GAS™ system produces significant by-product methane compared to coke gasification. This limits the quantity of CO and H_2 in the syngas available for conversion to F-T liquids and leads to increased F-T offgas which must be used in a gas turbine.
- Increased offgas requires two gas turbines, which in turn requires that more syngas be bypassed around the F-T plant to fully load the gas turbine.

However, in Subtask 2.3, the lower efficiency from the reduced liquids production is somewhat offset by the use of a more efficient reheat steam cycle.

Several options were considered to optimize the plant configuration and to improve the project economics. These options focused on maximizing the production of F-T liquids at the expense of reduced power output to take advantage of the high value of the F-T liquids and the higher efficiency of F-T liquids production compared to the conversion to electricity. If all the available syngas were sent to F-T hydrocarbon synthesis (in a once-through configuration), the plant would produce 15,016 bpd (compared to 12,377 bpd for the current design). In this configuration the F-T offgas available for gas turbine fuel would be 2,729 MMBtu/hr LHV at 185 Btu/scf (compared to about 1,750 MMBtu/hr LHV at a minimum energy content of 200 Btu/scf as specified by General Electric for the 7FA+e combustion turbine). The current design, which bypasses some syngas around the F-T reactors, sends 3,532 MMBtu/hr at 210 Btu/scf to fire two GE 7FA+e turbines (1,766 MMBtu/hr per turbine). A better approach would be to use a General Electric 9F or 7G/H class gas turbine, but the 9F model turbine is a 50-Hertz machine, and General Electric is not currently offering the advanced machines for syngas service.

Reducing the amount of fuel gas was also considered. One approach is to increase F-T reactor conversion to offset the high methane production during gasification with the coal feedstock. However this approach does not appear to be realistic because of the requirements for adequate reactor sizing (residence time and height), mixing, and heat transfer. Hydrogen recovery from the F-T offgas also was considered, but was ruled out because of the low concentration of hydrogen. The last option considered was to provide fuel to one GE 7FAe+ gas turbine and use the excess offgas for supplemental duct firing of the HRSG. However, this option was dropped from consideration because of the large quantity of energy (970 MMBtu/hr), which would cause a high HRSG inlet temperature. Also, the offgas would only be converted to electricity at the moderate steam cycle efficiency. In any case, the overriding considerations are to minimize capital cost, and to maximize availability and use of all process equipment to produce high value products and maximize revenue.

Figure 3.4 is a simplified block flow diagram of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. This figure also contains the flow rates of the major plant input and output streams. This plant can be considered to consist of three distinct main processing areas:

- The gasification island and air separation unit (Areas 100, 150, 250, 300, 350, 400, 420, and 800)
- The F-T hydrocarbon synthesis area (Areas 200 and 201)
- The power block (Areas 500 and 600)

In addition there is a balance of plant area (Area 900). The remainder of this section describes the three main processing areas, the balance of plant area, and discusses the plant EPC cost.

Table III.1

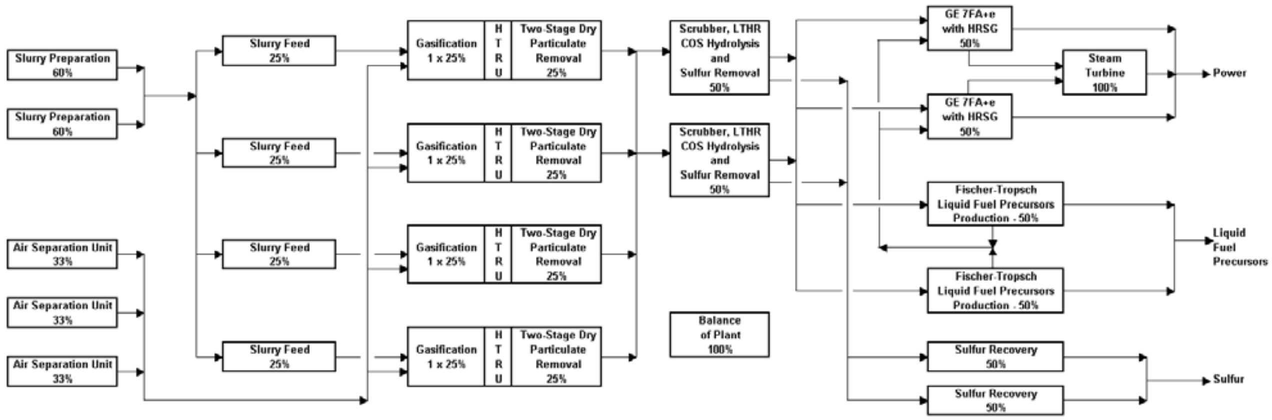
Design Input and Output Streams for the Subtask 2.3 Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 MW IGCC plant

	Subtask 1.6 Nominal 1,000 MW IGCC Plant	Subtask 2.3 Optimized Power and Liquids Plant
Plant Inputs		
Coal Feed, as received TPD	10,837	10,837
Dry Coke Feed to Gasifiers, TPD	9,266	9,266
Oxygen Production, TPD of 95% O_2	8,009	7,919
Total Fresh Water Consumption, gpm	9,752	7,403
Plant Outputs		
Net Power Output, MW	1,154.6	675.9
Sulfur, TPD	273	237
Slag, TPD (15% moisture)	1,423	1,423
Liquid Fuel Precursors, bpd	0	12,377

Figure 3.1

Subtask 2.3 - Train Diagram

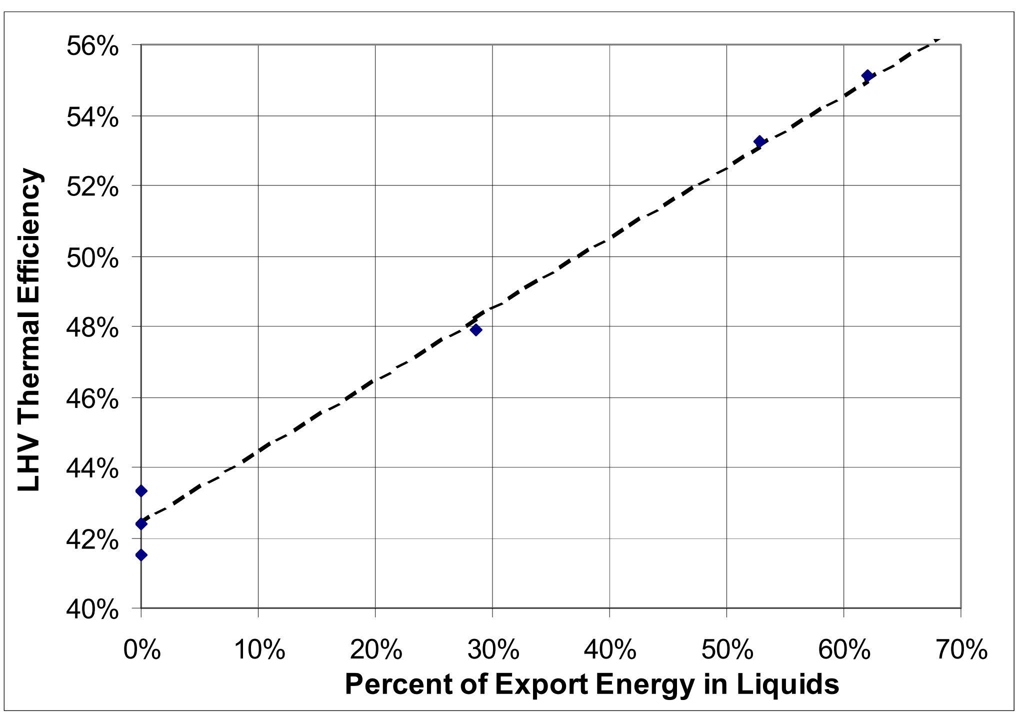
Optimized Coal Gasification Power Plant with Liquid Fuels Coproduction

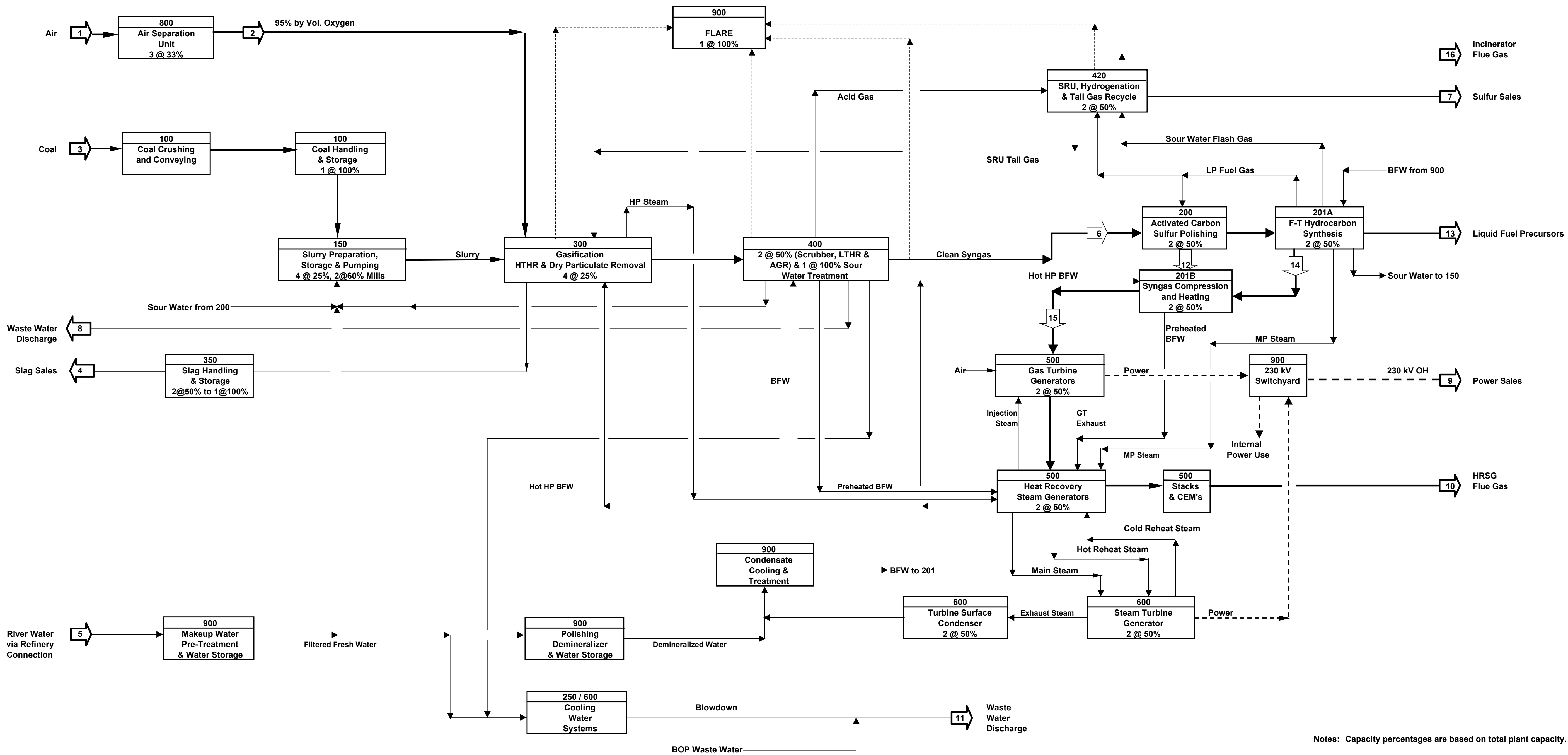


Note: Capacity percentages are based on total plant capacity.

Figure 3.2

Effect of Liquids Production on the LHV Thermal Efficiency





Notes: Capacity percentages are based on total plant capacity.

	1	2	3	4	5	6	7	Check 8	9	10	11	12	13	14	15	16					
Flow	Air 34,528 Tons/Day	Oxygen 7,919 Tons/Day	Coal 9,266 Tons/Day	Slag 1,423 Tons/Day	Water 3,702,000 Lb/Hr	Syngas 1,477,400 Lb/Hr	Sulfur 236.5 Tons/Day	Water 55,563 Lb/Hr	Power 675,900 kWe	Flue Gas 7,970,000 Lb/Hr	Water 695,500 Lb/Hr	Syngas 263,425 Lb/Hr	Liq Fuel 146,018 Lb/hr	Fuel Gas 1,041,200 Lb/hr	GT Fuel 1,304,625 Lb/hr	Flue Gas 24,758 Lb/Hr					
Nominal Pressure - psig	Atmos.	610	NA	NA	50	365	NA	62	NA	Atmos.	Atmos.	360	50	314	350	Atmos.					
Temperature - F	59	240	Ambient	180	70	100	332	80	NA	237	71	100	100	80	400	500					
HHV Btu/lb	NA	NA	12,749	NA	NA	5,255	3,983	NA	NA	NA	NA	5,255	19,701	2,376	2,957	NA					
LHV Btu/lb	NA	NA	12,275	NA	NA	4,894	3,983	NA	NA	NA	NA	4,894	18,225	2,154	2,707	NA					
Energy - MM HHV/hr	NA	NA	9,844	NA	NA	7,764	78	NA	NA	NA	NA	1,384	2,877	2,474	3,858	NA					
Energy - MM LHV/hr	NA	NA	9,478	NA	NA	7,231	78	NA	NA	NA	NA	1,289	2,661	2,243	3,532	NA					
Notes	Dry Basis	7,468 O2	Dry Basis	15%Wtr.	7404 GPM		Sales	111 GPM	230 kV		1390 GPM	No S	12377 bpd								

DOE Gasification Plant Cost and Performance Optimization

Figure 3.3

Subtask 2.3

OPTIMIZED COAL GASIFICATION POWER PLANT WITH LIQUIDS COPRODUCTION

BLOCK FLOW DIAGRAM

File: Fig 3.2.xls

July 25, 2003

3.1 Air Separation Unit and Gasification Island

The gasification island and the air separation unit basically are the same as those in the Subtask 1.6 Nominal 1,000 MW IGCC plant, as described in Appendix G of reference 2. Therefore, detailed descriptions of these areas will not be repeated here. The Gasification Island deals with coke handling, gasification, syngas cooling and cleanup, sulfur production, and slag handling. These are Areas 100, 150, 250, 300, 350, 400, and 420. The Air Separation Unit has been renumbered to Area 800 for this case in order to allow the F-T hydrocarbon synthesis area to be Areas 200 and 201 to be more consistent with the nomenclature used in the indirect baseline study.

The fuel handling system (Area 100) provides the means to receive, unload, store, and convey the coal to the storage facility. The coal is passed over the weigh belt feeders and transferred by the feed conveyors to the day storage bins above the rod mills in the slurry preparation area.

The slurry preparation area also contains the rod mills for grinding the coke to the desired slurry solids concentration. Coal and raw water are fed to each rod mill, along with water that is recycled from other areas of the gasification plant. Prepared slurry is stored in agitated tanks before being pumped into the gasifier.

The gasification, high temperature heat recovery, and particulate removal system (Area 300) is the heart of the Gasification Island. Global Energy's E-GAS™ gasification process consists of two stages, a slagging first-stage and an entrained flow non-slagging second-stage. The slagging section, or first-stage, is a horizontal refractory lined vessel into which the coal slurry, recycle solids, and oxygen are atomized via opposing mixer nozzles. The coal slurry, recycle solids, and oxygen are fed sub-stoichiometrically to the gasifier vessel and react at elevated temperature and pressure to produce a high temperature syngas. The oxygen feed rate to the mixers is carefully controlled to maintain the gasification temperature above the ash fusion point; thereby ensuring good slag removal while producing high quality syngas.

The raw synthesis gas generated in the first-stage flows up from the horizontal section into the second-stage of the gasifier. The second-stage is a vertical refractory-lined vessel into which additional coke slurry is injected via an atomizing nozzle to mix with the hot syngas stream exiting the first-stage. No oxygen is introduced into the second-stage. This additional slurry lowers the temperature of the gas exiting the first-stage by vaporizing the water in the slurry feed and by the endothermic nature of the steam and CO_2 reactions with carbon, thereby generating syngas and increasing cold gas efficiency.

The coal is almost totally gasified to form a synthetic fuel gas consisting primarily of hydrogen, carbon monoxide, carbon dioxide, and water. Sulfur in the coke is primarily converted to hydrogen sulfide (H_2S) with a small portion converted to carbonyl sulfide (COS); both of which are easily removed by downstream processing. Residual tars are removed by passing the hot syngas through the post reactor residence vessel.

The gas and entrained particulate matter exiting the gasifier is further cooled in a firetube heat recovery boiler system to produce saturated steam at 1,650 psia which is superheated in the HRSG and used for power generation. The syngas leaving the high temperature heat recovery unit passes through a two-step cyclone/dry char filter particulate removal system to remove solids from the syngas. The recovered particulates are recycled to the gasifier. Water-soluble impurities are removed from the syngas in a wet scrubbing column following the dry char filters.

Mineral matter in the coal, form a molten slag which flows continuously through the tap hole into a water quench bath located below the first-stage. The slag then is crushed and removed through a continuous pressure letdown system as a slag/water slurry. This continuous slag removal technique eliminates high-maintenance, problem-prone lock hoppers and completely prevents the escape of raw gasification products to the atmosphere during slag removal.

The Area 350 slag handling and storage system processes and stores the slag. The slag slurry leaving the slag crushers at the outlet of the quench section of the gasifier flows continuously through the pressure let down system into a dewatering bin. After passing through a settling tank to remove fine particles, the clear water is cooled in heat exchangers before it is returned to the gasifier quench section. The dewatered slag is loaded into trucks or rail cars for transport to market or to storage. The fines from the bottom of the settling tank are recycled to the slurry preparation area.

Area 400 contains the COS hydrolysis unit, low temperature heat recovery system, sour water treatment system, and the acid gas removal system.

Since COS is not removed efficiently by the downstream Acid Gas Removal (AGR) system, the COS must be converted to H_2S in order to obtain the desired high sulfur removal level. This is accomplished by the catalytic reaction of the COS with water vapor in the COS hydrolysis unit to create H_2S and CO_2. The H_2S is removed in the downstream AGR section, and most of the CO_2 remains in the syngas.

Upon exiting the COS hydrolysis unit, the syngas is cooled in a series of shell and tube exchangers which condense water, ammonia, some carbon dioxide, and hydrogen sulfide in an aqueous solution. This water goes to the sour water treatment unit. Some of the cooled syngas goes to the syngas recycle compressor for use in various areas of the plant; such as for quench gas in the second-stage of the gasifier, particulate recycle, and for back pulsing the dry char filters.

The heat removed prior to the AGR unit provides energy to heat F-T feed and gas turbine fuel gas, steam for the AGR stripper, and condensate heating. Cooling water provides trim cooling to ensure that the syngas enters the AGR at a sufficiently low temperature.

The sour water treatment system removes the small amounts of dissolved gases (i.e., carbon dioxide, hydrogen sulfide, ammonia, and other trace contaminants) from the condensed water and any other process water. The gases are stripped out of the sour water in a two-step process. First, the acid gases are removed in the acid gas stripper column by steam stripping. The stripped gases go to the Sulfur Recovery Unit (SRU). The water from the acid gas stripper column, is cooled, and a major portion is recycled to slurry preparation. The remainder is treated in the ammonia stripper column to remove the ammonia, filtered to remove trace organics and solids, and then sent to the waste water management system. The stripped ammonia is combined with water that is recycled back to the slurry mix tank after being cooled with cooling tower water.

The acid gas removal (AGR) system removes the H_2S from the syngas to produce a low sulfur syngas. The H_2S is removed from the sour syngas in an absorber column at high pressure and low temperature using a solvent, methydiethanolamine (MDEA). After the hydrogen sulfide removal, all of the un-moisturized syngas is sent to Area 200 for sulfur polishing before F-T synthesis.

The H_2S rich MDEA solution leaving the absorber goes to a stripper column where the H_2S is removed by steam stripping at a lower pressure. The concentrated H_2S exits the top of the stripper column and goes to the Sulfur Recovery Unit. The lean amine exits the bottom of the stripper, is cooled, and then recycled to the absorber. An online MDEA reclaim unit continuously removes impurities from the lean amine to improve system efficiency.

The Area 420 sulfur recovery unit (SRU) processes the concentrated H_2S from the AGR unit and the CO_2 and H_2S stripped from the sour water in a reaction furnace, a waste heat recovery boiler, and then in a series of Claus catalytic reaction stages where the H_2S is converted to elemental sulfur. The sulfur is recovered as a molten liquid and sold as a by-product.

The tail gas stream, composed of mostly carbon dioxide and nitrogen with trace amounts of sulfur dioxide, exits the last catalytic stage and goes to tail gas recycling. It is hydrogenated to convert all the remaining sulfur species to H_2S, cooled to condense the bulk of the water, compressed, and then injected into the gasifier. This allows for very high sulfur removal efficiency with low recycle rates.

Area 800 contains three 33% capacity Air Separation Units (ASUs) to deliver the required oxygen for the gasification process. Each ASU consists of several subsystems and major pieces of equipment, including an air compressor, air cooling system, air purification system, cold box, and product handling and backup systems.

Gaseous oxygen leaves the cold boxes at moderate pressure and is compressed in centrifugal compressors and delivered to the gasifiers. Nitrogen tanks with steam vaporizers provide gaseous nitrogen for various in-plant uses such as purging vessels.

The Area 250 cooling water system provides cooling water to the Gasification Island and ASU. A second system provides the cooling duty for the power block.

The major components of the cooling water system consist of a cooling tower, circulating water pumps, and appropriate piping for distribution of the cooling water around the facility. Both cooling towers are multi-cell mechanically induced draft towers, sized to provide the design heat rejection at the ambient conditions corresponding to the maximum summer temperature. Chemical treatment systems, including metering pumps, storage tanks and unloading facilities provide the necessary biocide, pH treatment and corrosion inhibiting chemicals for the circulating water system. Cooling tower blowdown discharges to the wastewater management system.

3.2 Fischer-Tropsch Hydrocarbon Synthesis Area

The design for the Fischer-Tropsch Hydrocarbon Synthesis Area was developed based on the ASPEN Plus process flowsheet reactor model that was developed for the Baseline Design/Economics for Advanced Fischer-Tropsch Technology study.[4] The ASPEN Plus process flowsheet model of the Fischer-Tropsch (F-T) hydrocarbon synthesis area that was developed for this study does not include the following systems:

- Filter system (and associated hydrocarbon circulation loop) which removes the catalyst from the liquid product leaving the slurry-bed F-T reactor
- Used catalyst removal and disposal system
- Fresh catalyst handling and pretreatment systems

The designs for these systems were developed based on the previous Baseline F-T design study.[3]

The Fischer-Tropsch hydrocarbon synthesis area consists of two parallel trains with each train containing two sub areas, Area 200 and Area 201. The configuration of each train is the same as that used in the Subtask 2.2 coke plant. Area 200 is the Final Syngas Cleanup Area, which removes the final traces of sulfur from the syngas, before it is converted to hydrocarbons in Area 201, the Hydrocarbon Synthesis and Product Recovery Area.

3.2.1 Final Syngas Cleanup Area

The purpose of the Final Syngas Cleanup Area, Area 200, is to reduce the sulfur concentration of the cleaned syngas from the acid gas removal area of the gasification block to less than 0.5 ppm of sulfur. This is accomplished by absorbing the small amounts hydrogen sulfide, carbonyl sulfide (COS) and trace amounts of other light organic sulfur compounds (such as CS_2) on metal impregnated activated carbon. The active bed is regenerated weekly with IP steam and a small quantity of air, and the off gas is sent to the sour water stripper (SWS) overhead cooling system to condense the steam prior to going to Claus sulfur recovery. After its useful life, the deactivated carbon is sent to the gasifier for destruction and conversion to syngas and slag. The metal activator is entrained in the slag, which is a non-hazardous waste.

In order to provide continuous H_2S removal, the process design uses a three bed reactor configuration with two beds in series to remove sulfur (the second bed is a guard bed). The third bed is in regeneration. Necessary piping is provided so that these beds can be switched into any position, and when necessary, the spent adsorbent can be replaced without any interruption of service. When H_2S breakthrough occurs in the first bed (lead bed), it is taken out of service for regeneration (or adsorbent replacement, when necessary), and the other bed (lag bed) is placed in the first position. The freshly regenerated bed now becomes the second bed. This two bed in series operation continues until H_2S breakthrough occurs in the first bed, and it is removed from service for regeneration causing the operating cycle to repeat. Each carbon bed is sized for a one week cycle. Each activated carbon bed has an expected life of about three years so that, on average, one bed should be replaced each year.

Figure 3.4 contains a schematic flow diagram of the Final Syngas Cleanup Area, Area 200 and the F-T slurry-bed reactor and product recovery area, Area 201.

The Area 200 Final syngas cleanup area is sized to treat all of the syngas produced by the gasification block. After treatment, about 82% of the cleaned syngas is set to the Area 201 Fischer-Tropsch slurry-bed reactor. The remaining cleaned syngas is mixed with the unconverted syngas and light hydrocarbons from the F-T area, compressed, moisturized, heated, and mixed with steam to become fuel gas for the combustion turbine.

3.2.2 Fischer-Tropsch Slurry-bed Reactor Area

The Fischer-Tropsch slurry-bed reactor converts the sulfur-free syngas primarily into olefinic hydrocarbons by the reaction

$$n\,CO + 2n\,H_2 \rightarrow C_nH_{2n} + n\,H_2O$$

The reaction is promoted by an iron-based catalyst, which also promotes the water-gas shift reaction

$$CO + H_2O \rightarrow CO_2 + H_2$$

The lighter hydrocarbon products leave the slurry-bed reactor in the vapor phase, are cooled and the condensed liquid collected. The heavier hydrocarbons are removed as liquids from the reactor, separated from the suspended catalyst, cooled, and combined with the lighter products to form the liquid fuel precursors product.

In order to maintain a constant catalyst activity, there is a continual addition of fresh catalyst and a continual withdrawal of used catalyst from the slurry-bed. The fresh catalyst must be pretreated in a reducing atmosphere at an elevated temperature to activate it. The catalyst pretreating system consists of a similar vessel to the slurry-bed reactor, but without the internal cooling facilities.

Figures 3.4 and 3.5 contain flow diagrams of the syngas and liquids processing areas of Plants 200 and 201. Figure 3.6 contains flow diagrams of the catalyst preparation and removal systems in the F-T processing area.

The cleaned syngas from the gasification block is preheated to 244ºF with low-level heat from the gasification area. The preheated syngas is mixed with 440ºF/375 psia stream that was generated in the slurry-bed F-T reactor and fed to the 200R-1 slurry-bed F-T hydrocarbon synthesis reactor.

The slurry-bed F-T hydrocarbon synthesis reactor, 201R-1, converts the hydrogen and carbon monoxide to straight chain aliphatic hydrocarbons, carbon dioxide and water. The heat of reaction is removed from the slurry-bed F-T reactor by the generation of 440°F/375 psia steam inside tubes located within the slurry-bed reactor. Pump 201P-2 circulates boiler feed water (BFW) between the steam drum 201C-1 and the F-T reactor 201R-1 to ensure that sufficient BFW always is flowing through the cooling tubes.

Cyclone 201T-1 removes entrained catalyst particles from the vapor stream leaving the top of the F-T reactor. The vapor stream then is cooled to 40°F in four exchangers. The first exchanger (201E-1) cools the syngas to 130ºF by heating BFW. The cooled syngas leaving the first exchanger enters the 201C-22 reactor overhead flash drum to separate condensate. The next exchanger (201E-2) cools the syngas to 100ºF with cooling water. The next two exchangers (210E-19 and 201E-20) chill the syngas to 40ºF. The chilled syngas enters the 201C-2 reactor overhead flash drum. The sour water from the boot of 201C-2 goes to the 201C-4 sour water flash drum. The vapor stream leaving the sour water flash drum goes to the incinerator, and the sour water is recycled to the gasifier.

The vapor stream from the reactor overhead vapor flash drum is routed through 201E-19 to recover refrigeration and is washed in 201C-3 to remove any residual catalyst particles prior to compression. A propane refrigeration system provides the refrigerant used in 201E-20. The washed vapor stream is compressed to 390 psia in 201K-1. This compressed stream is mixed with clean syngas from final gas cleanup and is sent to the power block where it is heated to 250ºF by low-level energy from the gasification area and then to 400ºF with intermediate pressure (400 psia) steam in 201E-23 and is used for combustion turbine fuel. This stream consists of unconverted syngas (carbon monoxide and hydrogen) and light hydrocarbons (primarily C1 through C5s) and carbon dioxide that are produced in the F-T reactor.

The liquid hydrocarbon stream leaving 201C-2 is mixed with the cooled liquid hydrocarbons from the slurry-bed F-T reactor and sent for upgrading into liquid transportation fuels.

The liquid stream leaving the slurry-bed F-T reactor passes through hydroclone 201T-2 to remove a majority of the entrained catalyst particles. The catalyst-rich hydroclone bottoms goes to mixing tank 201C-10 from which most of it is returned to the slurry-bed reactor by pump 201P-3. A portion of the hydroclone bottoms is withdrawn and sent to the catalyst withdrawal system shown in Figure 3.4. Residual catalyst particles are removed from the hydroclone overhead stream in the filter system 201T-3.

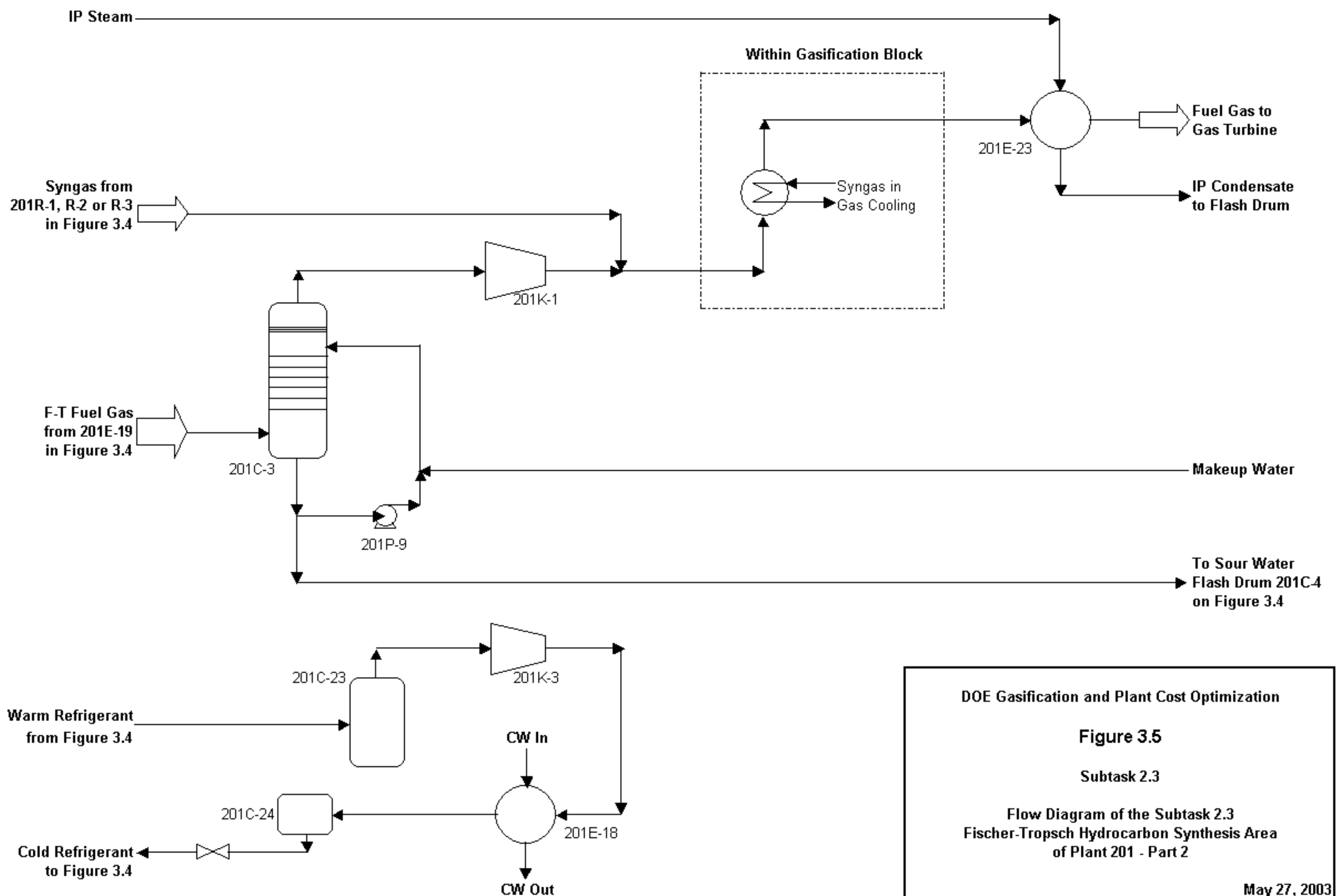
The catalyst-free liquid leaving the filter system is reduced in pressure and flashed in drum 201C-5. The vapor stream is further cooled to 100°F in exchanger 201E-4 with cooling water and flashed in drum 201C-6. The vapor stream from drum 201C-6 is a low-pressure fuel gas, which is used as fuel in the 200F-1 furnace.

The liquid leaving the flash drum 201C-5 is cooled to 200ºF in 201E-5 by preheating boiler feed water. The cooled liquid from 201C-5 is mixed with the liquid stream from the flash drum 201C-6 in drum 201C-9 and a cooled liquid recycle stream from 201C-8. This mixture now is cooled to 110ºF by cooling water in exchanger 201E-6 and sent to the 201C-8 liquid fuel flash drum along with the liquid from the reactor overhead vapor flash drum 201C-2. The vapor leaving the 201C-8 liquid fuel flash drum is mixed with the vapor from the flash drum 201C-6 and is used as low-pressure fuel gas in the gasification area.

The liquid from the flash drum 201C-8 is split into two streams. One of the liquid streams is recycled back to flash drum 201C-9 via pump 201P-1 to dilute the heavier hydrocarbons in order to control their viscosity as they are cooled in exchanger 201E-6. The other liquid stream is the liquid fuel precursors product, which is sent to the adjacent petroleum refinery for upgrading into liquid transportation fuels (gasoline, diesel, etc.).

Figure 3.6 shows the catalyst withdrawal system. The hot catalyst-rich stream from the 201C-10 drum is cooled in exchanger 201E-7 and pumped by pump 201P-4 through the 201T-4 filters to remove the used catalyst which is collected and discarded. The catalyst free liquid is mixed with the liquid fuel precursors product stream from drum 201C-6 and sent to the adjacent petroleum refinery for upgrading.

The catalyst pretreatment system also is shown in Figure 3.6. The makeup catalyst is fed into the 201C-11 catalyst pretreater where it is combined with heated liquid product from storage. Recycle gas is circulated through the pretreater vessel via compressor 201K-2, exchanger 201E-9, and exchanger 201E-11, which uses high-pressure steam from the gasification plant as the heating media. Vapors leaving the pretreater vessel are cooled in exchangers 201E-9 and 201E-10 before being flashed in drum 201C-13. A portion of the vapor from 201C-13 is withdrawn and sent to the incinerator to remove inerts from the system. However, most of the vapors from 201C-13 are recycled to the pretreater after addition of some fresh syngas or hydrogen via the 201K-2 compressor, exchanger 201E-9, and exchanger 201E-11. Pretreated catalyst is withdrawn from the pretreater vessel and stored in the heated 201C-12 mixing tank until it is injected into the slurry-bed F-T reactors via pump 201P-8.



DOE Gasification and Plant Cost Optimization

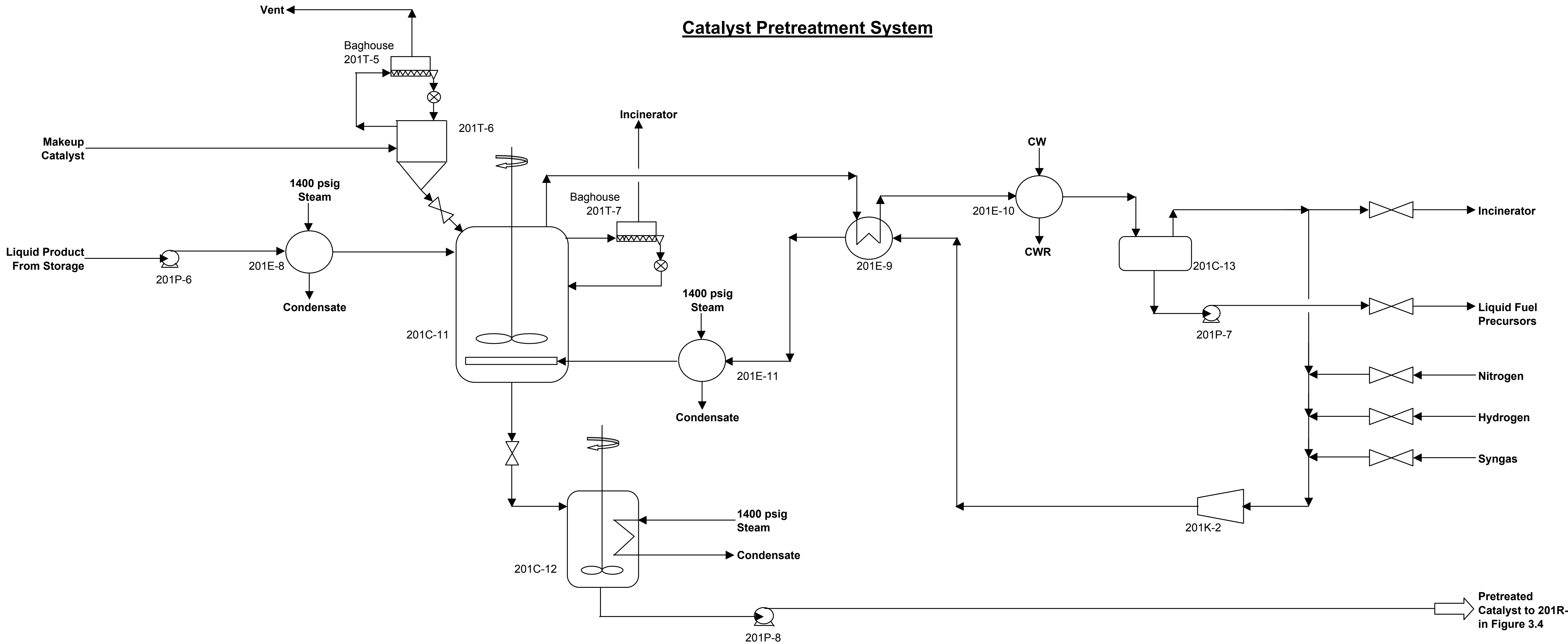
Figure 3.5

Subtask 2.3

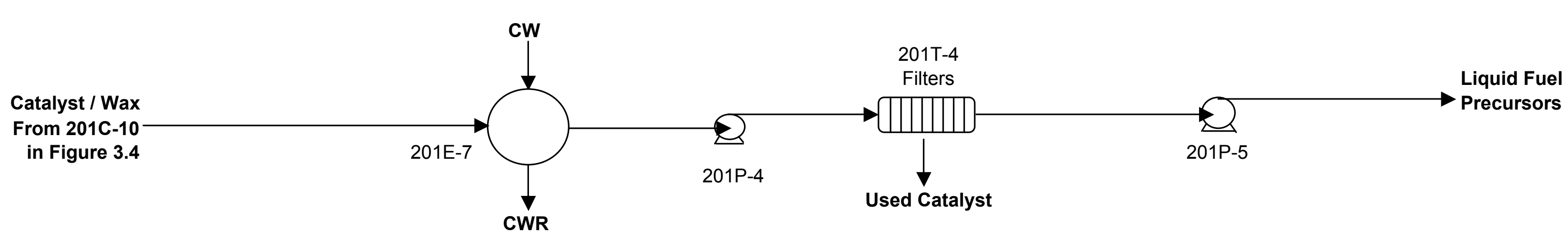
Flow Diagram of the Subtask 2.3
Fischer-Tropsch Hydrocarbon Synthesis Area
of Plant 201 - Part 2

May 27, 2003

Catalyst Pretreatment System



Catalyst Withdrawal System



DOE Gasification and Plant Cost Optimization

Figure 3.6

Subtask 2.3

Flow Diagram of the Subtask 2.3
Catalyst Pretreating and Catalyst Removal
Areas of Plant 201

May 19, 2003

3.3 Power Block

The major components of the combined cycle power block include two gas turbine generators (GTG), two heat recovery steam generators (HRSG), a steam turbine generator (STG), and numerous supporting facilities.

Area 500 contains two trains, each train consists of a gas turbine (GT), heat recovery steam generator (HRSG), and stack. The combustion turbine generators are a General Electric 7FA+e machine with a nominal output of 208 MW. The GTs utilize syngas at 210 Btu/scf with additional steam diluent injection for NO_x control. Combustion exhaust gases from each GT are routed to the HRSG and stack. Natural gas is used as back-up fuel for the gas turbines during startup, shutdown, and short duration transients in syngas supply. Optionally, the gas turbines can be fully fired on natural gas to generate power when syngas is unavailable.

Each HRSG receives the GT exhaust gases and generates steam at the main steam, medium-pressure-steam, and low-pressure steam pressure levels. The HRSG's provide superheating for all three levels of steam generation, including superheating the high-pressure steam generated in the gasification process and the medium-pressure steam from F-T liquids production. In addition the HRSG's include steam reheater sections producing hot reheat steam for the steam turbine. The HRSG's also provide condensate and feed water heating for both the combined cycle and the gasification facilities. The HRSG is a fully integrated system consisting of all required ductwork and boiler components. Each component is designed for pressurized operation.

The HRSG boilers include steam drums for proper steam purity and to reduce surge during cold start. Large unheated down comers assure proper circulation in each of the banks.

Each stack includes stack emissions sampling equipment. The power block HRSG's share a continuous emission monitoring (CEM) system.

The Area 600 steam turbine (ST) is a reheat, condensing turbine that includes an integrated HP/IP opposed flow section and four-flow, down-exhaust, LP section. Turbine exhaust steam is condensed in a surface condenser. The steam turbine produces 403.6 MW of electric power.

The power delivery system includes the GT generator output at 18 kilovolts (kV) with each connected through a generator breaker to its associated main power step-up transformer. A separate main step-up transformer and generator breaker is included for the ST generator. The HV switchyard receives the energy from the three generator step-up transformers at 230 kV.

Two auxiliary transformers are connected between the GTG breakers and the step-up transformers. Due to the large auxiliary load associated with the plant, internal power is distributed at 33 kV from the two auxiliary power transformers. 33/13.8 kV transformers service the major motor loads in the air separation units. Several substations serve the other internal loads with 33/4.16 kV transformers supplying a double-ended electrical bus.

Area 600 also includes a cooling water system similar to that in Area 250 and an emergency shutdown transformer to connect the 230 kV switchyard with essential safe shutdown loads.

3.4 Balance of Plant

The Area 900 balance of plant contains nine subsystems.

The fresh water supply system filters river water from an industrial water supply network for the fresh makeup water supply. A demineralizer supplies water for boiler water makeup. The demineralizer regeneration wastewater is sent to a process waste collection tank, where it is neutralized before discharge.

The fire and service water system includes a loop around the principal facilities with fire hydrants located for easy access. It also includes an onsite water storage tank. A jockey pump maintains line pressure in the loop during stand-by periods. During periods of high water usage, motor and diesel driven pumps are available.

The wastewater management system processes both clear wastewater and storm water from a clean water collection pond. Clear wastewater includes water treatment effluent, cooling water blowdown, flushes and purges from equipment maintenance, filtered water from the ammonia stripper column (in Area 400), clarifier overflow, and sewage treatment overflow. Storm water is collected in a storm-water pond before going to the clean water collection pond. The water in the clean water collection pond is analyzed and treated, as required, until it meets permitted outfall specifications for discharge.

The service and instrument air system provides compressed air and dried instrument air to users throughout the plant. The system consists of air compressors, air receivers, hose stations, and piping distribution for each unit.

The incineration system destroys the tank vent streams from various in-process storage tanks and drums that may contain small amounts of hydrocarbons and other gases such as ammonia and acid gas. During process upsets of SRU, tail gas streams also can be processed in the high temperature incinerator. The high temperature produced in the incinerator thermally destroys any residual hydrogen sulfide before the gas is vented to the atmosphere. The incinerator exhaust feeds into a heat recovery boiler to produce process steam.

The flare system provides for safe disposal of syngas during startup or short term upsets. The flare includes a natural gas fired pilot flame to ensure that the flare is continually operating.

The instrumentation and control system provides data acquisition, monitoring, alarming and control by a digital distributed control system (DCS). The DCS allows the plant to be operated from the central control room using the DCS as the control platforms. The gas and steam turbines, and the coke handling programmable logic controllers will continue to execute all permissive, protective, and sequence control related to their respective equipment. They will be controlled either locally using the turbine vendor man machine interface system, or from the DCS.

Other balance of plant equipment such as air compressors, condenser vacuum pumps, and water treatment facilities can be controlled by either local PLCs, or from contact and relay control cabinets. All remaining plant components are exclusively controlled by the DCS including the HRSG, the gasifier, ASU, hydrogen plant, electrical distribution, and other power block and gasification support systems.

The plant has a central building housing the main control room, office, training, other administration areas, and a warehouse/maintenance area. Other buildings are provided for water

treatment equipment and the motor control centers. The buildings, with the exception of water treatment, are heated and air-conditioned to provide a climate-controlled environment for personnel and electrical control equipment.

A series of strategically placed safety showers are located throughout the facility.

3.5 Plant Layout

Figure 3.7 is a site plan of the Subtask 2.3 Optimized Coal Gasification Power Plant with F-T Liquids Coproduction. The plant occupies about 61 acres and is slightly smaller than the site of the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant.

The site plan is very similar to that of the Subtask 1.6 IGCC Plant. However, in the Subtask 2.3 plant, the two F-T hydrocarbon synthesis trains replace one of the two combined cycle power blocks of the Subtask 1.6 IGCC Plant. Also, Subtask 2.3 requires a slightly larger cooling tower containing 20 cells rather than the 18 cell tower provided for Subtask 1.6. The larger cooling tower is needed to condense the additional medium-pressure (440ºF/360 psig) steam generated in the F-T area.

3.6 Thermal Efficiency

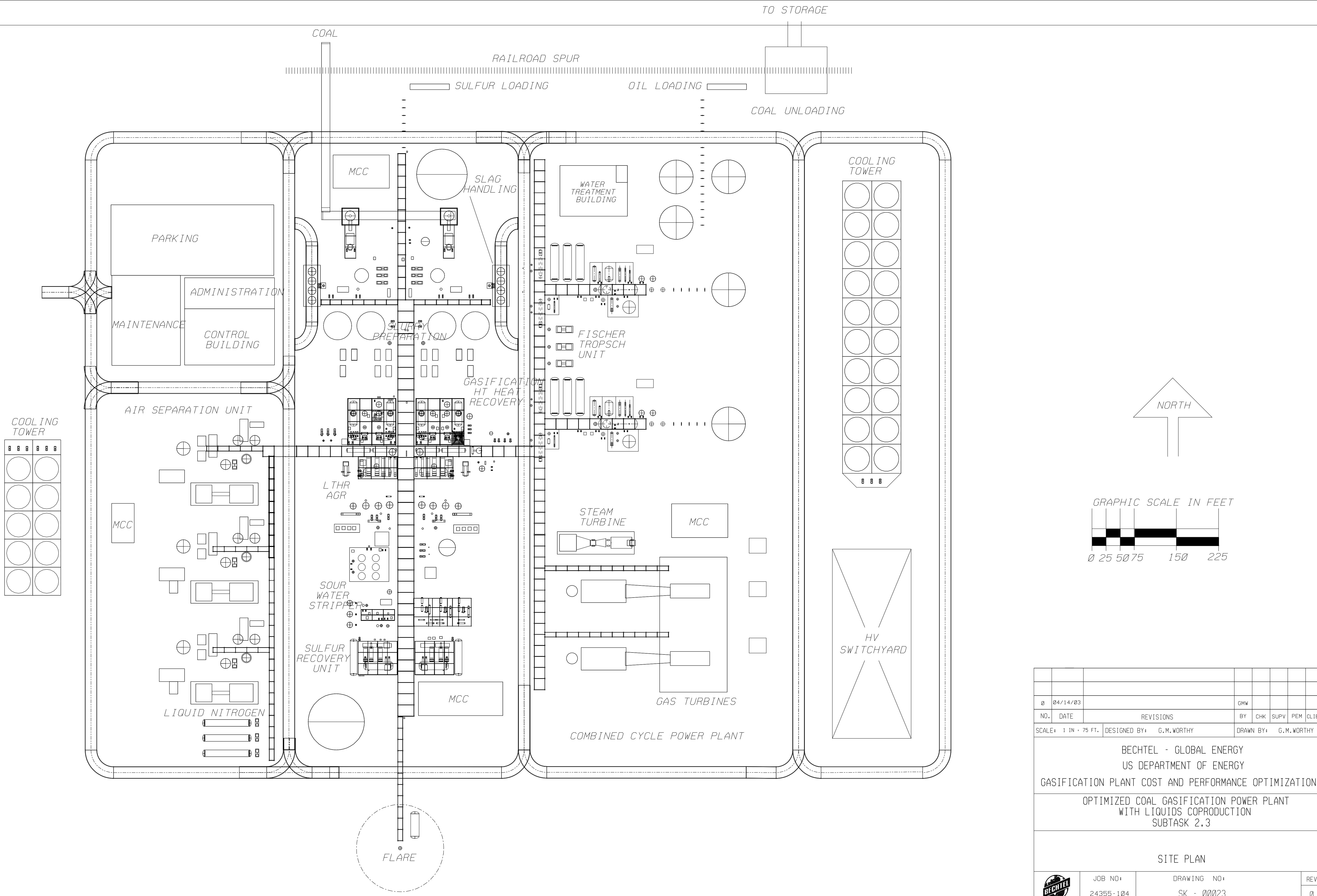
Table III.2 shows the thermal efficiencies of the Subtask 2.2 Optimized Coke and Subtask 2.3 Optimized Coal Gasification Power Plants with Liquids Coproduction based on the energy content of the F-T liquid fuel, sulfur products, and the equivalent energy in the export power. Also shown are the thermal efficiencies of the Subtask 1.5A and 1.6 Coal IGCC Power Plants. The thermal efficiencies of the Subtask 2.2 and 2.3 optimized plants are significantly higher than those of Subtask 1.5A and 1.6 coal IGCC plants because of the energy contained in the liquid F-T product. The efficiency of syngas conversion to liquids (excluding the syngas that ends up as fuel gas) is about 75%. Most of the remaining 25% is recovered as steam for power generation. In comparison, the efficiency of a combined-cycle power plant is less than 60%, and the steam cycle efficiency is less than 30%. Therefore, increasing the production of F-T liquids increases the overall plant efficiency

Table III.2

Thermal Efficiencies of Four Gasification Plants

	Subtask 2.2 Coke FT/IGCC	Subtask 2.3 Coal FT/IGCC	Subtask 1.5A 300 MW Coal IGCC	Subtask 1.6 1,000 MW Coal IGCC
LHV Basis	55.14%	53.24%	41.48%	42.39%
HHV Basis	56.74%	53.45%	39.94%	40.82%

On an LHV basis, the thermal efficiency of the Subtask 2.3 optimized plant is 53.24%, which is 10.85% higher than the thermal efficiency of the Subtask 1.6 IGCC plant. On an HHV basis, the thermal efficiency of the Subtask 2.3 optimized plant is 53.45%, which is 12.63% higher than that of the IGCC plant.



NO.	DATE	REVISIONS	BY	CHK	SUPV	PEM	CLIENT
Ø	Ø4/14/Ø3		GMW				

SCALE: 1 IN - 75 FT. DESIGNED BY: G.M.WORTHY DRAWN BY: G.M.WORTHY

BECHTEL - GLOBAL ENERGY
US DEPARTMENT OF ENERGY
GASIFICATION PLANT COST AND PERFORMANCE OPTIMIZATION

OPTIMIZED COAL GASIFICATION POWER PLANT
WITH LIQUIDS COPRODUCTION
SUBTASK 2.3

SITE PLAN

JOB NO: 24355-104 DRAWING NO: SK - ØØØ23 REV. Ø

3.7 Emissions

Table III.3 shows the atmospheric emissions summary of the Subtask 1.6 Nominal 1,000 MW IGCC Plant, and the Subtask 2.3 Optimized Coal Gasification Power Plants with Liquids Coproduction. The atmospheric emissions of the Subtask 2.3 plant are lower than those of the Subtask 1.6 IGCC Plant. The Subtask 2.3 plant emits about 7,995,000 lb/hr of total exhaust gases having an average SOx concentration of 22 ppmv, an average NOx concentration of 16 ppmv, and an average CO concentration of 16 ppmv. Expressed another way, this is 329 lb/hr of SOx (as SO_2), 166 lb/hr of NOx (as NO_2), and 65 lb/hr of CO.

The sulfur emissions from the Subtask 2.3 gas turbine are very low because almost all the sulfur is removed from the syngas by adsorption on activated carbon (and recovered as sulfur) whether it goes to the F-T synthesis reactors or whether it goes directly to the turbine. In the Subtask 1.6 Nominal 1,000 MW IGCC Plant, the sulfur content of the syngas going to the gas turbine is about 20 ppmv. The NOx and CO emissions from the gas turbines for both cases are about the same when expressed on ppmv (at 15% oxygen, dry basis). However, the Subtask 2.3 NO_2 and CO rates are the highest on an weight basis because the Subtask 2.3 turbine exhaust is drier since the CO_2 produced in the F-T area is used for NOx control (instead of moisturization).

Both cases have about the same absolute sulfur emissions from the incinerator stack because the sulfur comes from the purge and blow down streams from the gasification block. The slightly lower sulfur emissions from the Task 2.3 plant are the result of small process improvements made since the Subtask 1.6 design was developed. On a ppmv basis, both cases produce about the same amount of NOx and CO, but because of the larger flow rate in the Subtask 2.3 incinerator stack, the absolute rates are higher.

In addition to the above atmospheric emissions, the Subtask 2.3 and 1.6 plants generate some solid wastes. The Subtask 2.3 optimized plant generates about 1,133,000 lb/year of used F-T catalyst. It also generates about 200,000 lb/year (6,000 ft^3/year) of used activated carbon adsorbent which is disposed of by mixing it with the coal feed and gasifying it to make syngas.

Both plants also generate about 1,423 tpd of slag which is a non-hazardous byproduct that is used in construction projects.

3.8 Plant Cost

Table III.4 shows the "overnight" EPC cost for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and compares it with the Subtask 1.6 Nominal 1,000 MW IGCC. These costs are on a mid-year 2000 basis; the same basis as those of the other Task 1 plant costs.[7]

The Subtask 2.3 EPC cost was developed from the Subtask 1.6 IGCC plant and Subtask 2.2 liquid fuel precursors coproduction plant EPC costs by adding the cost of the F-T hydrocarbon synthesis area to the subtask 1.6 cost estimate and adjusting the cost of the power block, to reflect only one 2x1x1 combined cycle train. No adjustments were made to the costs of the solids

[7] All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

handling area. The ASU cost was reduced to reflect a slight reduction in oxygen usage. The cost of the gasification block was adjusted to account for the removal of two syngas moisturizers. Adjustments also were made to the balance of plant area, as appropriate.

The cost of the F-T area was estimated from the processing equipment sizes using an appropriate installation factor that was developed from previous cost estimates for similar facilities. The estimated cost of the large F-T slurry-bed hydrocarbon synthesis reactor is over 60% of the total equipment cost in the F-T area, and consequently, it dominates the cost of this area. Until wider experience is obtained with the construction of these large reactors, their estimated cost cannot have a high degree of accuracy.

The accuracy of the total installed cost for the Subtask 1.6 Nominal 1,000 MW IGCC Plant was estimated to be on the order of ±15%. This level of accuracy reflects a high degree of confidence based on the large number of vendor quotes that were obtained and that the power block costs are based on a current similar Gulf Coast power project. This accuracy applies only to the total plant cost and does not apply to the individual areas or parts.

The accuracy of the total installed cost for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction is not as good. The estimated cost of the F-T area is only an order of magnitude cost estimate (nominally ±30%) because of the manner in which it was developed. Thus, the over estimate accuracy for the Subtask 2.2 plant probably is in the ±20% range. Because the cost of the F-T area of the Subtask 2.3 plant also is a large portion of the plant cost, the accuracy of the Subtask 2.3 is approximately the same, ±20%.

Table III.3

Atmospheric Emissions Summary* of the Subtask 1.6 and Subtask 2.3 Gasification Plants

	Subtask 1.6 1,000 MW Coal IGCC Power Plant	**Subtask 2.3** Coal Gasification Power Plant with Liquid Fuels
Total Gas Turbine Emissions		
Number of GT/HTSG Trains	4	2
GT/HTSG Stack Exhaust Flow Rate, lb/hr	15,928,800	7,970,230
GT/HTSG Stack Exhaust Temperature, ^{o}F	238	237
Emissions:		
SOx, ppmvd	2.6	<1
SOx as SO_2, lb/hr	95	<1
NOx, ppmvd (at 15% oxygen, dry basis)	10	10
NOx as NO_2, lb/hr	274	165
CO, ppmvd	10	10
CO, lb/hr	130	64
Incinerator Emissions		
Stack Exhaust Flow Rate, lb/hr	21,360	24,760
Stack Exhaust Temperature, ^{o}F	610	500
Emissions:		
SOx, ppmvd	7,365	6,351
SOx as SO_2, lb/hr	343.4	329
NOx, ppmvd (at 15% oxygen, dry basis)	40	40
NOx as NO_2, lb/hr	0.7	1
CO, ppmvd	50	50
CO, lb/hr	1	1
Total Plant Emissions		
Exhaust Flow Rate, lb/hr	15,950,100	7,995,000
Emissions:		
SOx, ppmvd	15	22
SOx as SO_2, lb/hr	438	329
NOx, ppmvd (at 15% oxygen, dry basis)	13	16
NOx as NO_2, lb/hr	275	166
CO, ppmvd	10	10
CO, lb/hr	131	65
VOC and Particulates, lb/hr	NIL	NIL
Opacity	0	0
Sulfur Removal, %	98.9	99.5

* Expected emissions performance

Table III.4

Capital Cost Summary for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 Coal IGCC Power Plant

Plant Area	Subtask 2.3 Optimized Power and Liquids Plant	Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant
Solids Handling	28,317,000	28,317,000
Air Separation Unit	149,791,000	151,496,000
Gasification	434,094,000	443,301,000
F-T Liquids Area	94,283,000	0
Power Block	348,788,000	493,795,000
Balance of Plant	103,785,000	114,419,000
Total	1,159,058,000	1,231,328,000

Notes:
1 Because of rounding, the columns may not add to the total that is shown.
2. All plant EPC costs mentioned in this report are mid-year 2000 order of magnitude cost estimates which exclude contingency, taxes, licensing fees, and owners costs (such as land, operating and maintenance equipment, capital spares, operator training, and commercial test runs).

Section 4

Availability Analysis

In all the previous Task 1 and Task 2 cases, an availability analysis was used to determine the daily average production rates for calculating the annual production rates and cash flow. The common measures of financial performance, such as return on investment (ROI), net present value (NPV), and payback period, all are dependent on the project cash flow. The net cash flow is the sum of all project revenues and expenses. Depending upon the detail of the financial analysis, the cash flow streams usually are computed on 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. Therefore, the annual production rate is a key parameter in determining the financial performance of a project.

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.[1] 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 three adjustments, this data was used to estimate the availability of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction under two operating scenarios, with and without backup natural gas. The first adjustment increased the availability of the air separation plant from the observed availability of 96.32% to the industry average availability of 98%. The second adjusted the availability of the first gasification stage to remove a slag tap plugging problem caused by an unexpected change in the coal blend to the gasifier. This adjustment is justified since a dedicated petroleum coke plant would be very unlikely to experience this problem. The third eliminated a short outage that occurred in the water treatment facility because this plant will have sufficient treated water storage to handle this type of outage.

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.[8] However, the following availability and financial analyses will be based on the data reported in the final repowering project report for consistency with the Task 1 results. This will cause the following results to be somewhat conservative, but they will be consistent with the previously reported Task 1 results.

Using the EPRI recommended procedure, availability estimates were calculated for the two operating scenarios of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction.[9]

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

[9] Research Report AP-4216, *Availability Analysis Handbook for Coal Gasification and Combustion Turbine-based Power Systems,* Research Project 1800-1, Electric Power Research Institute, 3412 Hillview Avenue, Palo Alto, CA 94304, August 1985.

The syngas production and cleanup area of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction are configured identical to that of the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant in that they both contain the same number of units used to generate the syngas. These areas contain two 60% slurry preparation areas, three 33.3% Air Separation Units, and four 25% gasification blocks, each with associated slurry feed, syngas cooling and cleanup sections. The final syngas cleanup and conditioning section consisting of a wet scrubber, low temperature heat recovery (LTHR), COS hydrolysis, sulfur removal, and sulfur recovery is in two 50% trains; the same as the Subtask 1.6 design. In the Subtask 1.6 design, the syngas is sent to four 25% GE7FA+e combustion turbines and HRSGs with one 50% steam turbine associated with two combustion turbines. In the Subtask 2.3 design, most of the syngas is sent to two Fischer-Tropsch hydrocarbon synthesis trains with the remainder along with the unconverted syngas from the F-T area going to two GE7FA+e combustion turbines and HRSGs. One steam turbine converts the steam from the other areas of the plant to power.

An availability analysis of both these facilities showed that the syngas availability from one gasification train including scheduled downtime is about 76%. Therefore, two modes of operation were considered; one without the use of backup natural gas and one that uses backup natural gas when sufficient syngas is not available to fire the combustion turbines. In the Subtask 2.3 case, backup natural gas only is used when insufficient syngas and F-T off gas are available to fully load a turbine. No turbine is operated only on backup natural gas.

Table IV.1 shows the design and annual average feed and product rates for both operating scenarios, with and without backup natural gas for the Subtask 1.6 and Subtask 2.3 plants. At design conditions, both plants process the same amount of dry coal, 9,266 tpd. However, because of some efficiency improvements developed since the Subtask 1.6 design was completed, the Subtask 2.3 plant uses slightly less oxygen. The Subtask 2.3 average rates were developed based on the premise that it was best to maximize power production from all available gas turbines by using backup natural gas even if it were necessary to fully fire the turbine on natural gas.

On a daily average basis without backup natural gas, the Subtask 2.3 plant processes slightly less coal that the Subtask 1.6 plants because there are some situations where it was necessary to slightly reduce the coal rate in order not to overload a combustion turbine. It processes 6,899 tpd of dry coal (74.4% of design) to produce 474.4 MW of power (70.2% of design) and 9,889 bpd of liquid hydrocarbons (80.7% of design).

On a daily average basis with backup natural gas, the Subtask 2.3 plant processes 6,929 tpd of dry coal (74.8% of design) and consumes 26.5 MMscfd of natural gas to produce 613.7 MW of power (90.8% of design) and 10,397 bpd of liquid hydrocarbons (84.0% of design).

Table IV.1

Design and Daily Average Feed and Product Rates for the Subtask 1.6 Coal IGCC Power Plant and the Subtask 2.3 Coal IGCC Coproduction Plant

	Subtask 1.6 1,000 MW Coal IGCC Power Plant			Subtask 2.3 Coal IGCC Coproduction Plant		
		Daily Average			Daily Average	
	Design	Without Backup Gas	With Backup Gas	Design	Without Backup Gas	With Backup Gas
Feeds						
Coal, TPD dry	9,266	7,018	7,018	9,266	6,899	6,929
Natural Gas, Mscfd	0	0	34,961	0	0	26,466
River Water, gpm	9,752	7,386	NC	7,404	5,513	NC
Products						
Export Power, MW	1,154.6	874.5	1,081.0	675.9	474.4	613.7
Liquid Hydrocarbons, bpd	---	---	---	12,377	9,989	10,397
Sulfur, TPD	236.6	179.2	179.2	236.6	176.1	176.9
Slag, TPD	1,423	1,078	1,078	1,423	1,059	1,064
Performance						
Oxygen Consumption,						
TPD of 95% O_2	8,009	6,066	6,066	7,919	5,896	5,922
TPD O_2/TPD dry coal	0.86	0.86	0.86	0.85	0.85	0.85
Water Discharge, gpm						
Process Water	59	45	45	111	83	83
Clear Water	1,248	945	NC	1,390	1,035	NC
Total Discharge	1,307	990	NC	1,501	1,118	NC
Heat Rate, Btu/kW	8,526	8,526	8,245	NC	NC	NC
Thermal Efficiency, %HHV*	40.0%	40.0%	41.4%	52.6%	52.6%	56.7%
Emissions						
SO_2, lb/MW-hr	0.38	0.38	0.31	0.49	0.49	0.40
CO, lb/M-hr	0.11	0.11	NC	0.10	0.10	NC
NOx, lb/MW-hr	0.24	0.24	NC	0.25	0.25	NC
Sulfur Removal, %	98.9	98.9	98.9	99.5	99.5	99.5
Plant Area, acres	62			61		
Installed Cost, MM$[2]	1,231			1,159		
Installed Cost, $/kW	1,066			NC		

NC = Not Calculated
* = Without including the byproduct sulfur

Section 5

Financial Analysis

The following financial analysis was performed using the discounted cash flow (DCF) model that was developed by Bechtel Technology and Consulting (now Nexant Inc.) for the DOE as part of the Integrated Gasification Combined Cycle (IGCC) Economic and Capital Budgeting Practices Task.[10] This model calculates investment decision criteria used by industrial end-users and project developers to evaluate the economic feasibility of IGCC projects.

5.1 Financial Model Input Data

The required input information to the DCF financial model is organized into two distinct input areas that are called the Plant Input Sheet and the Scenario Input Sheet. The Plant Input Sheet contains data that are directly related to the specific plant as follows.

Data Contained on the Plant Input Sheet

- Project summary information
- Plant output and operating data
- Capital costs
- Operating costs and expenses

The Scenario Input Sheet contains data that are related to the general economic environment that is associated with the plant as well as some data that are plant related. The data on the Scenario Input Sheet are shown below.

Data Contained on the Scenario Input Sheet

- Financial and economic data
- Fuel data
- Tariff assumptions
- Construction schedule data
- Startup information

For all cases, the EPC spending pattern was adjusted to reflect forward escalation during the construction period since the EPC cost estimate is an "overnight" cost estimate based on mid-year 2000 costs.

Items that were excluded in the cost estimate, such as spares, owner's cost, contingency and risk are included in the financial analysis.

Table V.1 summarizes the basic input parameters to the financial model. The daily average plant input and output flow rates are given in Table IV.1

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

Table V.1

Basic Financial Model Input Parameters

Parameter	Value
Financial Parameters	
Owner's Contingency (% of EPC Cost)	5.0%
Development Fee (% of EPC Cost)	1.23%
Start-up Cost (% of EPC Cost)	1.50%
Additional Financing Cost, EPC Contingency, Risk and Fees, etc.	5.0%
Loan Amount (% of Cost)	80%
Loan Interest Rate	10% & 8%
Loan Financing Fee	3.0%
Loan Repayment Term, years	15
Income Tax Rate	40%
Construction Period, years	15
Start Up	
First Year's Average Capacity	60%
Prices	
Coal, $/dry ton	22.00
Natural Gas, $/MMBtu HHV or $/Mscf *	2.60
Fischer-Tropsch Liquids, $/bbl	30.00
Electric Power, $/MW	27.00
Sulfur, $/ton	30.00
Slag, $/ton (15% water)	0.00
Annual Inflation Rates	
Coal, $/dry ton	1.2%
Natural Gas, $/HHV MMBtu	3.9%
Fischer-Tropsch Liquids	3.1%
Electric Power, $/MW	1.7%
Sulfur, $/ton	0.0%
Slag, $/ton (15% water)	0.0%

* Natural gas is assumed to have a HHV Btu content of 1,000 Btu/scf.

5.2 Financial Model Results

Figure 5.1 shows the return on investment (ROI) as a function of the export power price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction under both operating scenarios and compares them with the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant with 2.60 $/MMBtu natural gas at a 10% loan interest rate. This figure shows that generally the operating scenarios that use backup natural gas have higher Return on Investments (ROIs) than the cases without backup natural gas. Below a power selling price of about 38 $/MW-hr, the Subtask 2.3 plant with the F-T liquids at 30 $/bbl has a higher ROI than the Subtask 1.6 power plant. Above this power selling price, the Subtask 1.6 plant has a higher ROI. The same situation is true for the two operating scenarios without backup natural gas except that with these cases, the breakeven power selling price is slightly higher, about 40 $/MW-hr.

Figure 5.1

Return on Investment vs. Power Price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant (10% Loan Interest Rate)

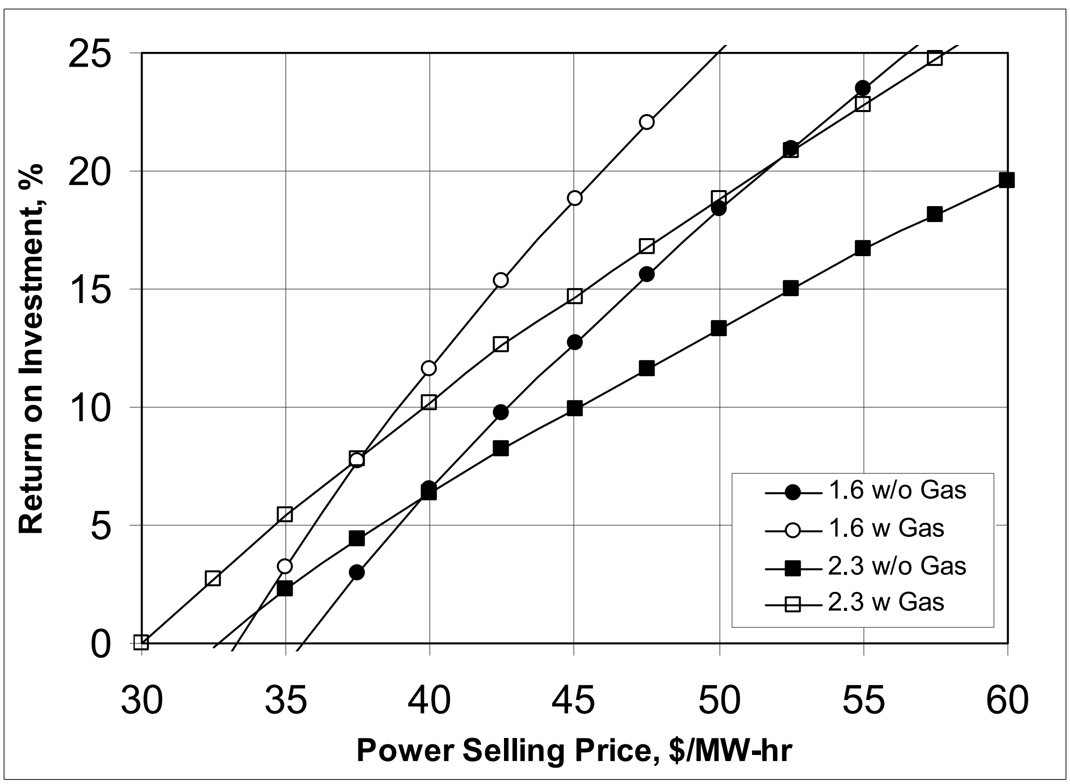


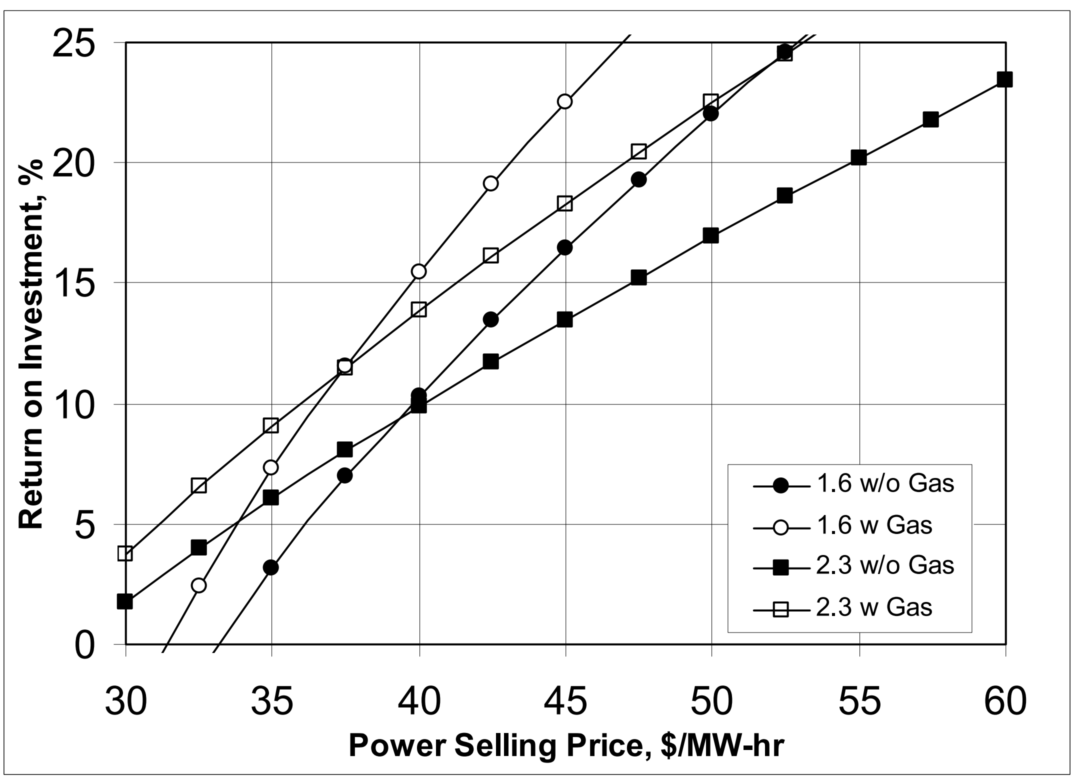
Table V.2 compares the power and F-T liquids selling prices required by the Subtask 1.6 and Subtask 2.3 plants to generate a 12% ROI for the two operating scenarios. At the basic economic conditions shown in Table V.1 (at a 10% loan interest rate), the Subtask 2.3 Coproduction Plant with backup power purchase requires a 42.02 $/MW-hr power selling price for a 12% ROI, and without backup power purchase, the required power selling price is 48.06 $/MW-hr. These required power selling prices are higher than those for the corresponding Subtask 1.6 cases. With a fixed 27 $/MW-hr power selling price, the required selling prices of the F-T liquids to produce a 12 ROI are 48.59 and 50.97 $/bbl for the cases with and without backup natural gas cases, respectively.

With an 8% loan interest rate the relative ranking of the cases remains almost the same except that the required selling prices are lower. However, the Subtask 2.3 case with backup natural gas now has a slightly lower power selling price than the Subtask 1.6 case. This is a result of the Subtask 2.3 case having a lower EPC cost than the Subtask 1.6 Case.

Figure 5.2 shows the return on investment (ROI) as a function of the export power price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction under both operating scenarios and compares them with the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant with 2.60 $/MMBtu natural gas at an 8% loan interest rate. This figure is very similar to Figure V.1 at a 10% loan interest rate, but in this case, the ROIs are higher.

Figure 5.2

Return on Investment vs. Power Price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant (8% Loan Interest Rate)

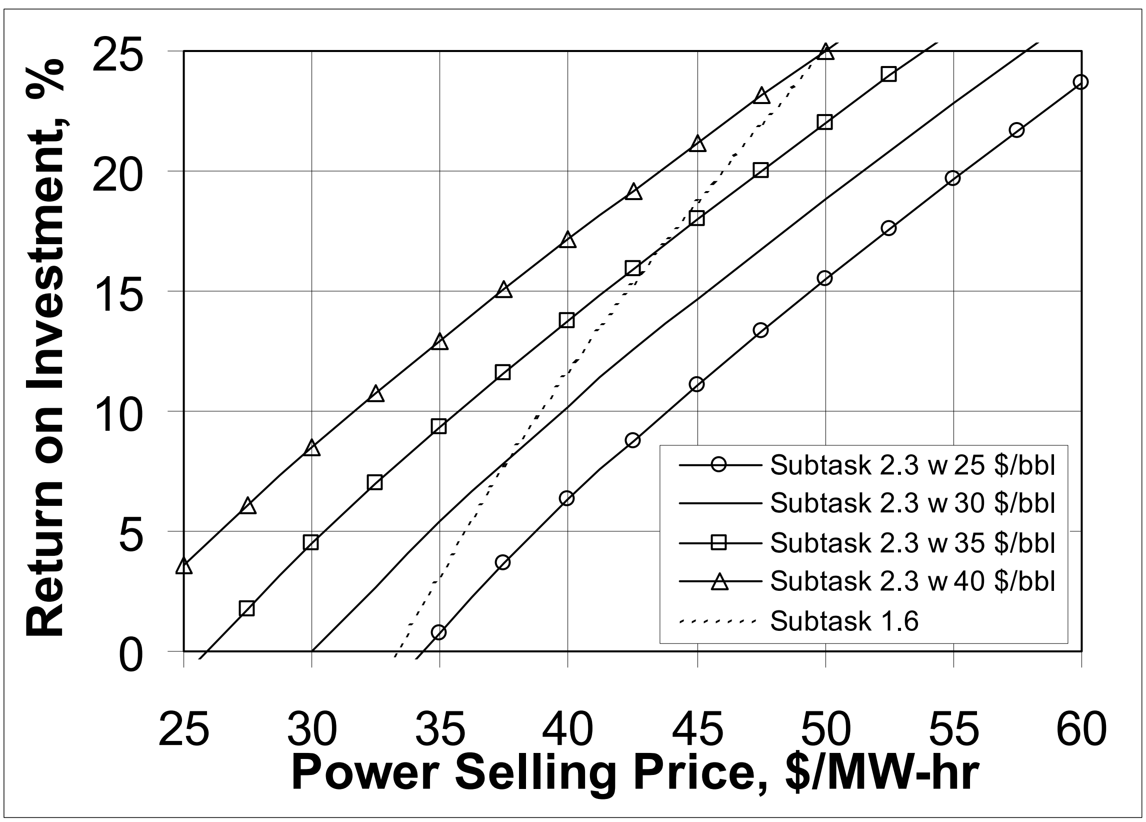


It is difficult to predict the future value of either power, natural gas and/or the F-T liquid fuel precursors. The liquid fuel precursors price is related to the crude oil price which also can be highly variable both because of market forces and the influence of international politics. Various studies have been made which attempt to relate the value of the F-T liquids to that of crude oil by replacing crude oil in the refinery feed stream with the F-T liquids. The resulting values for the F-T liquids generally are above crude oil values, but the specific amount can range from 2 $/bbl up to 10 $/bbl depending upon the refinery configuration, the specific crude oils being replaced, and the required refinery product mix.[11]

Figure 5.3 shows the effect of the liquid fuel precursors selling price on the return on investment versus the power selling price for the Subtask 2.2 Maximum F-T Liquids Case with a 10% loan interest rate and 2.60 $/MMBtu natural gas. The solid 30 $/bbl line is the same line as shown on the previous figure for the Subtask 2.3 coproduction plant with backup natural gas. The dashed line represents the Subtask 1.6 power plant with backup natural gas. The ROI for the Subtask 1.6 plant has a greater slope versus the power price than that of the Subtask 2.3 plant because the revenue generated from the power sales is a significantly larger portion of the total plant revenue. As such, any change in the power price will have a larger influence on the ROI.

Figure 5.3

Return on Investment vs. Power Price Showing the Effect of the Liquids Price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Power Plant



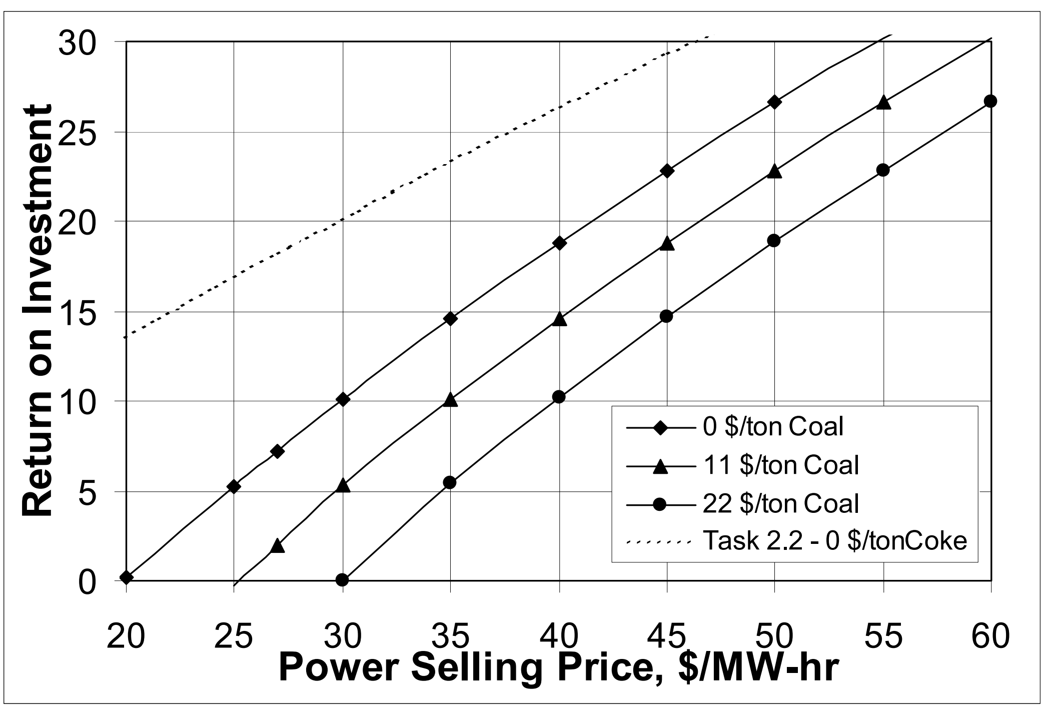
[11] Marano, J. J., Rogers, S., Choi, G. N., and Kramer, S. J., "Product Valuation of Fischer-Tropsch Derived Fuels," ACS National Meeting, Washington, D. C., August 21-6, 1994.

This figure shows that the Subtask 2.3 coproduction plant requires F-T liquids selling prices above 30 $/bbl to generate ROIs greater than 10% with power prices below 40 $/MW-hr. With a 38 $/MW-hr power Selling price, the Subtask 2.3 coproduction plant will have higher ROIs that the Subtask 1.6 power plant only when the F-T liquids are selling for 30$/bbl or greater. As the power selling price increases, the Subtask 2.3 coproduction plant requires higher F-T liquids prices to be competitive with the Subtask 1.6 plant. At a 50 $/MW-hr power price, the F-T liquids should be about 40 $/bbl or greater for the Subtask 2.3 plant to have a higher ROI.

Figure 5.4 shows the effect of the coal price on the return on investment on the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. The base coal price is 22.0 $/ton. Also shown on the figure are two other coal prices, 11.0 $/ton (50% of the base price) and 0 $/ton. As expected, as the coal price decreases, the ROI increases. For comparison, the return on investment of the Subtask 2.2 optimized coke is shown as the dotted line on the figure. This return is based on zero net coke price. The higher returns of the Subtask 2.2 plant shows that the cost of the coal alone does not account for the entire difference in returns between the two plants. Part of this difference is attributable to the higher fraction of high value liquid yields and higher thermodynamic efficiency of the Subtask 2.2 plant as discussed at the beginning of this chapter. The other part appears to be the higher availability of the gasification area of the Subtask 2.2 coke plant, which contains a spare gasification train (two operating and one spare), compared to the Subtask 2.3, which does not contain a spare train. Thus, on a daily average basis, the Subtask 2.3 plant uses a significant amount of higher priced natural gas (compared to coal) to increase export power production. Finally, Subtask 2.2 uses CO_2 instead of steam as diluent for NOx control in the combustion turbine which further increases the export power production from Subtask 2.2.

Figure 5.4

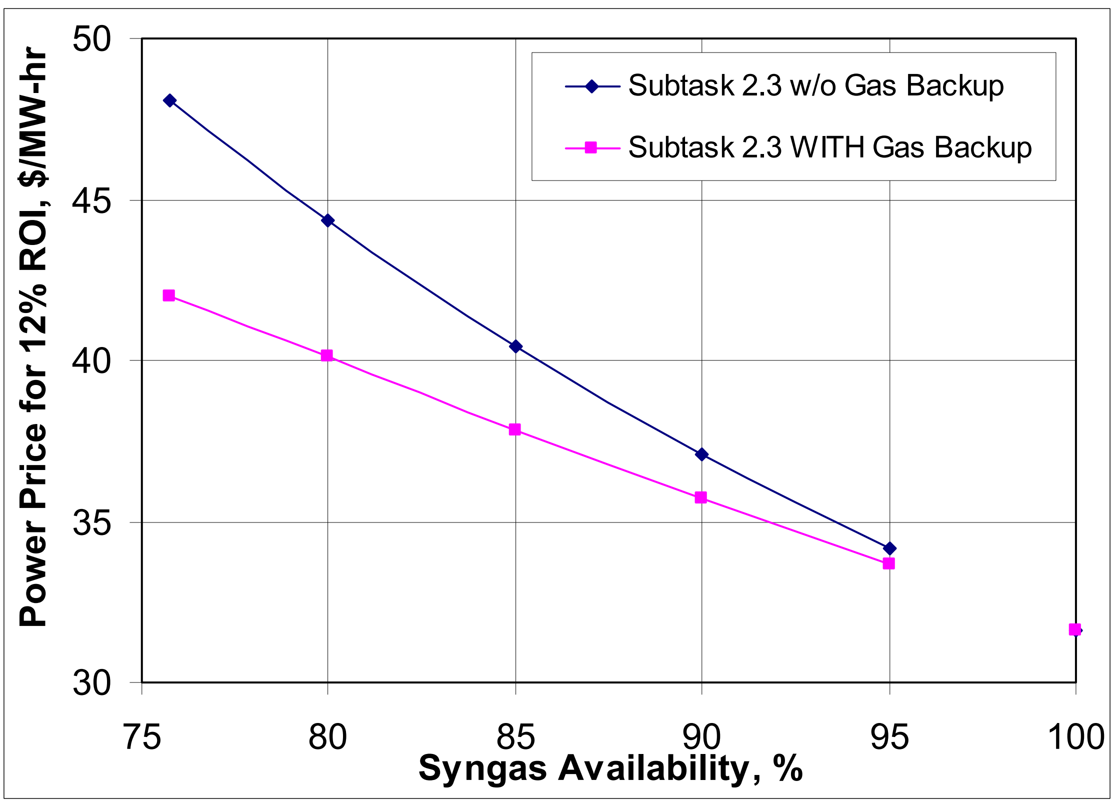
Return on Investment vs. Power Price Showing the Effect of the Coal Price for the Subtask 2.2 Optimized Coal Gasification Power Plants with Liquids Coproduction



After commissioning all plants undergo a "learning curve" during which problem areas are corrected, inadequate equipment is modified or replaced, and adjustments are made. Consequently, performance improves as measured by increased capacity and/or improved on-stream factors. Figure 5.5 shows the effect of improved syngas availability on the return on investment. This improved availability can be the result of "learning curve" improvements or design changes that are yet to be developed. For the Subtask 2.3 plant, as the syngas availability improves, the amount of backup natural gas is reduced until it disappears at the unattainable 100% syngas availability. At the expected 75.7% single train syngas availability, the Subtask 2.3 Case with backup natural gas requires power selling price of 42.02 $/MW-hr with 30 $/bb F-T liquids selling price to generate a 12% ROI. At an 80% syngas availability, the required power selling price drops by almost 2 $/MW-hr to 40.1 $/MW-hr. At the unattainable 100% syngas availability, no backup natural gas is required, and the required power selling price for a 12% ROI is 31.6 $/MW-hr.

Figure 5.5

Effect of Improved Syngas Availability for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction on the Required Power Selling Price for a 12% Return on Investment



Without backup natural gas, at the expected 75.7% single train syngas availability, the required power selling price for a 12% ROI with 30 $/bb F-T liquids selling price is 48.1 $/MW-hr. At an 80% syngas availability, the required power selling price drops by almost 4 $MW-hr to 44.3 $/MW-hr. At the unattainable 100% syngas availability, it is the same as the case with backup natural gas, 31.6 $/MW-hr, since at this point, no backup natural gas is required.

Table V.3 shows the sensitivity of some individual component prices and financial parameters for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction (Maximum F-T Liquids Case) starting from a 12% ROI (with a power price of 42.02 $/MW-hr and a 30.0 $/bbl liquids price). Each item was varied individually without affecting any other item. Most sensitivities are based on a ±10% change from the base value except when either a larger or smaller change is used because it either makes more sense or it is needed to show a meaningful result.

The power selling price is the most sensitive product price with a 10% increase to 46.22 $/MW-hr resulting in a 3.67% increase in the ROI to 15.67%, and a 10% decrease to 37.81 $/MW-hr resulting in a 3.82% decrease in the ROI to 8.18%. A 10% increase in the F-T Liquids price to 33.0 $/bbl will cause a 2.13% increase in the ROI to 14.13%, and a 10% decrease in the liquids price to 27 $/bbl will result in a 2.20% decrease in the ROI to 9.80%.

Changes in the sulfur and slag prices only have a small influence on the ROI.

A decrease in the coal price of 5 $/ton from 22 $/dry ton to a 17 $/ton will increase the ROI by 1.99% to 13.99%, and a 5 $/ton increase in the coal price will lower the ROI by 2.02% to 9.98%.

A 5% decrease in the plant EPC cost from 1159.1 MM$ to 1130.1 MM$ will increase the ROI by 2.34% to 14.34%, and a 5% increase in the plant cost to 1217.0 MM$ will decrease the ROI by 2.17% to 9.83%.

In today's unsettled financial situation, the loan interest rate and project financing conditions also can be uncertain. A 20% decrease in the loan interest rate to 8% from the base interest rate of 10% will increase the ROI to 15.66% from 12.00%, and a 20% increase in the interest rate to 12% will lower the ROI to 8.30%. A 20% decrease in the loan amount from 80% to 72% will lower the ROI by 0.55% to 11.44%, and a 20% increase in the loan amount to 88% will increase the ROI by 0.93% to 12.93%. Decreasing the income tax rate by 10% from 40% to 36% will increase the ROI by 0.48% to 12.48%, and a 10% increase in the tax rate to 44% will lower the ROI by 0.51% to 11.49%.

The final two lines of Table V.3 show the effect of hypothetically individually increasing and/or reducing the power and F-T liquids production rates without any other changes to the input and output stream flow rates or the plant cost. For example, suppose higher efficiency combustion and steam turbines are developed so that the net power output is increased by 2.5% to 602.2 MW from 613.7 MW, then the ROI would increase by 0.93% to 12.93%.

Effect of a Spare Gasification Train on Plant Performance

One way to increase availability and to improve the daily average production from the plant at minimal extra cost is to enlarge the gasification capacity of each train by one third so that each train now is 33.3% of the total design capacity of the plant. If this is done, then the gasification section of the plant now is a three train facility with a spare train. With this redesign, now there are three air separation units supplying oxygen to three operating gasification trains. The following table shows the effect of this redesign on the daily average feed and product rates for the coal gasification plant with liquid fuel precursors coproduction with backup natural gas.

Table V.4

Design and Daily Average Feed and Product Rates for Two Train Configurations of the Subtask 2.2 Optimized Coal Gasification Plant with Liquids Coproduction

	Design	4 x 25% Gasification Trains		3 x 33.3% Gasification Trains	
		Daily Avg. Rate	% of Design	Daily Avg. Rate	% of Design
Feeds					
Coal, dry tpd	9,266	6,929	74.8%	8,097	87.4%
Natural Gas, Mscf/hr	0	1,103	---	345	---
Products					
Export Power, MW	675.9	613.7	90.8%	618.9	91.6%
F-T Liquids, bpd	12,377	10,397	84.0%	11,260	91.0%
Sulfur, tpd	236.5	176.9	74.8%	206.7	87.4%
Slag, tpd (15% water)	1,423	1,064	74.8%	1,244	87.4%

Increasing the capacity of each gasification train by 33.3% to create a spare train configuration increases the daily average coal consumption rate by over 1,000 tpd of dry coal to 8,097 tpd (87.4% of design capacity) from 6,929 tpd (74.8% of design capacity). The increased coal consumption results in increased product rates and lower backup natural gas consumption rates. The daily average F-T liquids production increases to 11,260 bpd (91.6% of design) from 10,395 bpd (84% of design). The daily average power production does not increase as much, only to 618.9 MW (91.0% of design) because syngas is more available, and it, rather than backup natural gas, is used to generate power. The increase in the byproduct sulfur and slag production rates is directly proportional to the increase in the coal consumption rate.

Increasing the size of the four gasification trains from 25% to 33.3% of design capacity was estimated to increase the plant cost by about 43 MM$ to 1,202.06 MM$.

Figure 5.6 shows the effect of the size and number size of gasification trains on the return on Investment versus power price for the 4 x 25% and 4 x 33.3% cases, both with and without backup natural gas. The two 4 x 33.3% cases each have higher returns than the corresponding 4 x 25% case. Without backup natural gas at a power selling price of 40 $/MW-hr, the 4 x 33.3% case has an ROI of 12.1%, which is about 5.7 ROI percent higher than the 4 x 25% case. With backup natural gas, the ROI increase for the 4 x 33.3% case is not as great, only about 2.9 ROI percent, from 10.2% to 13.1% ROI.

Figure 5.6

Return on Investment vs. Power Price for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction Showing the Effect of the Number of Gasification Trains (10% Loan Interest Rate)

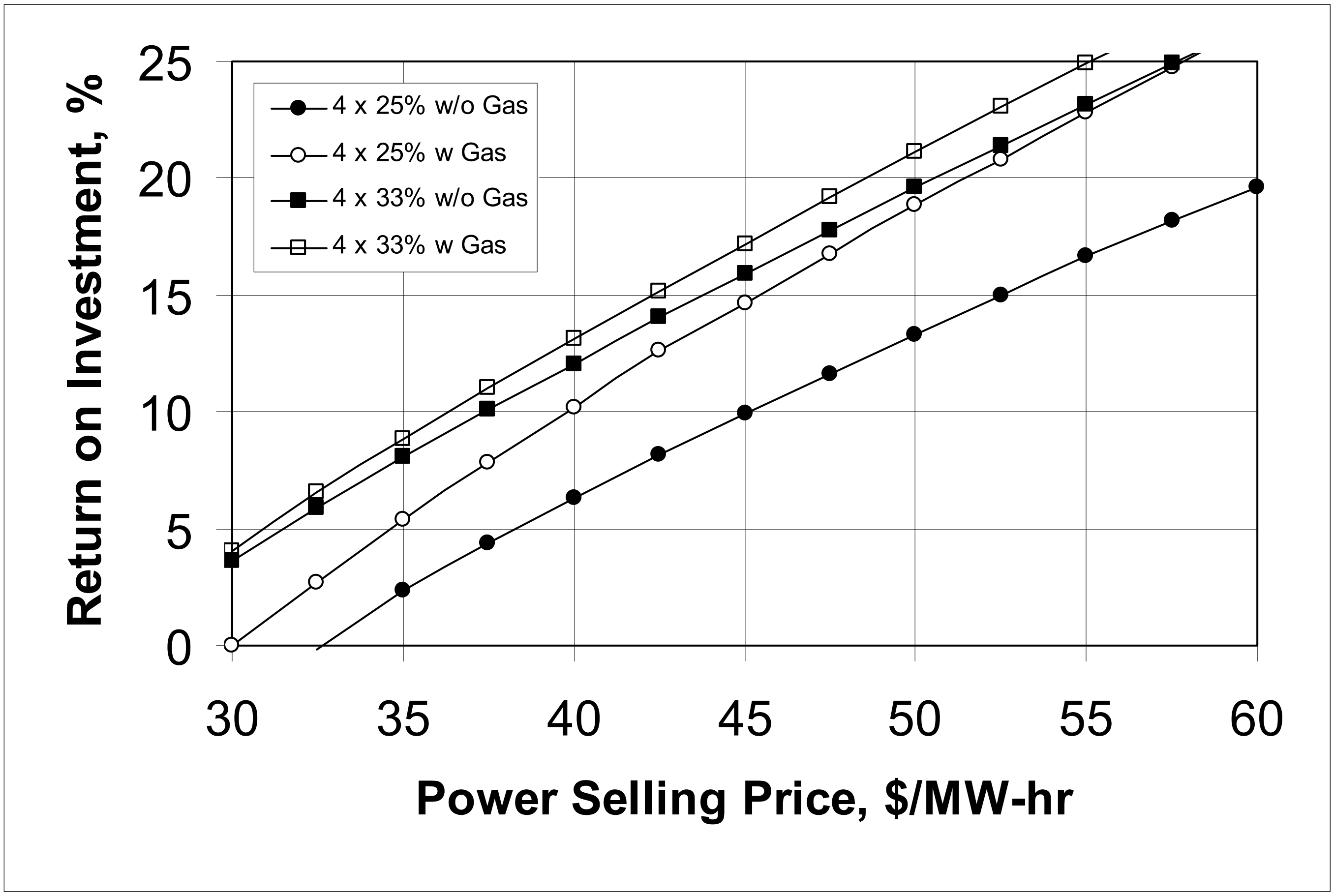


Table V.2

Required Power Selling Prices for the for the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Power Plant With an Without Backup Natural Gas

	Subtask 1.6		Subtask 2.3	
	Without Backup Gas	With Backup Gas	Without Backup Gas	With Backup Gas
With a 10% Loan Interest Rate				
Required Selling Price for a 12% ROI of Power with 30 $/bbl Liquids, $/MW-hr	44.37	40.23	48.06	42.02
Liquds with 27 $/MW-hr Power, $/bbl	---	---	50.97	48.59
With a 8% Loan Interest Rate				
Required Selling Price for a 12% ROI of Power with 30 $/bbl Liquids, $/MW-hr	41.34	37.77	42.93	38.06
Liquds with 27 $/MW-hr Power, $/bbl	---	---	45.87	43.69

Table V.3

Sensitivity of Individual Component Prices on the Return on Investment for the Subtask 2.3 Case with Backup Gas from a 12% ROI (at a Power Price of 42.02 $/MW-hr)

	Decrease				Increase		
	ROI	Value	% Change	Base Value	% Change	Value	ROI
Products							
Power	8.18%	37.81 $/MW-hr	-10%	**42.02 $MW-hr**	+10%	46.22 $/MW-hr	15.67%
Liquids	9.80%	27 $/bbl	-10%	**30 $/bbl**	+10%	33 $/bbl	14.13%
Slag	11.73%	-5 $/t	---	**0 $/t**	---	5 $/t	12.27%
Sulfur	11.97%	27 $/t	-10%	**30 $/t**	+10%	33 $/t	12.03%
Feeds							
Coal	13.99%	17 $/t	-23%	**22.0 $/t**	23%	27 $/t	9.98%
Natural Gas	12.51%	2.34 $/MMBtu	-10%	**2.60 $/MMBtu**	+10%	2.86 $/MMBtu	11.48%
Financial							
Plant Cost	13.15%	1130.1 MM$	-2.5%	**1159.1 MM$**	+2.5%	1188.0 MM$	10.90%
Plant Cost	14.34%	1101.1 MM$	-5.0%	**1159.1 MM$**	+5.0%	1217.0 MM$	9.83%
Interest Rate	15.66%	8%	-20%	**10%**	+20%	12%	8.30%
Loan Amount	11.44%	72%	-20%	**80%**	+20%	88%	12.93%
Tax Rate	12.48%	36%	10%	**40%**	+10%	44%	11.49%
Performance							
Average Power	11.06%	598.4 MW	-2.5%	**613.7 MW**	+2.5%	629.0 MW	12.93%
Liquids	11.46%	10,137 bpd	-2.5%	**10,397 bpd**	+2.5%	10,657 bpd	12.54%

Note: Products and Feeds each are listed in decreasing sensitivity.

Section 6

Summary

A design for an optimized coal gasification power plant with liquids coproduction using Fischer-Tropsch technology has been developed. The plant consumes 9,266 tpd of coal (dry basis) to produce 675.9 MW of export power and 12,377 bpd of liquid fuel precursors. It also produces 237 tpd of elemental sulfur and 1,423 tpd of slag. The plant is located in the U.S. Midwest adjacent to a suitable water source and in reasonable proximity to a petroleum refinery which can upgrade the liquid fuel precursors into transportation fuels.

The design of this Subtask 2.3 optimized plant was developed from those of the Subtask 2.2 Optimized Coke Gasification Power Plant with Liquids Coproduction and the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant. Starting from the Subtask 1.6 plant, two Fischer-Tropsch hydrocarbon production trains replaced once combined-cycle train (containing two gas turbines, two heat recovery steam generators (HRSGs), and one steam turbine). The Subtask 2.2 optimized F-T coproduction plant design provided a basis for integration of the Subtask 2.3 plant. In contrast to the Subtask 2.2 plant, the Subtask 2.3 plant purchases backup natural gas, instead of power, to increase availability.

Each of the two parallel Fischer-Tropsch hydrocarbon synthesis trains basically consists of three sections; final sulfur removal, slurry-bed F-T reactor, and product recovery sections. The final sulfur removal section contains three activated carbon beds in series, which adsorb the residual sulfur in the syngas. The activated carbon beds are regenerated with medium-pressure steam. This is a much simpler and less costly design than that of the Subtask 2.1 non-optimized plant which contains a hydrolysis reactor followed by a non-regenerable ZnO adsorbent.

Sulfur-free syngas is fed to the slurry-bed F-T reactors which convert it to hydrocarbons over an iron-based catalyst. The heat of reaction is removed by generation of 440°F/375 psia steam inside tubes that are placed within the slurry-bed. The lighter hydrocarbon products and unconverted syngas leave the reactor as vapors and are cooled by refrigeration to condense and recover the hydrocarbons as liquids. The unconverted syngas and non-condensable light hydrocarbons (primarily C1 through C3s) are compressed, mixed with bypass syngas, moisturized, and sent to the power block. The heavier products are removed from the reactor as liquids, separated from the entrained catalyst by filtration, cooled, mixed with the lighter hydrocarbons, and sent to a petroleum refinery for separation, upgrading and incorporation into liquid transportation fuels.

The F-T liquid fuel precursors essentially are a bottomless, sulfur-free crude oil. Basically they are straight-chain 1-olefins and paraffins without any aromatics. The diesel fraction has a very high cetane number (>70) and is a premium blending component for diesel fuel. The naphtha fraction is a low octane material that requires further upgrading for use as a gasoline blending component. However, it is an excellent feedstock for an ethylene cracker. Linear programming studies have shown that the F-T liquid fuel precursors may be worth up to 10 $/bbl more than crude oil depending upon the specific refinery configuration and product demands.

The combined cycle power block contains two GE7FAe+ combustion turbines, two HRSGs, and a reheat steam turbine. The combustion turbine fuel is a mixture of F-T off gas and about 18% of the available syngas, which bypasses the F-T synthesis reactors. The F-T off gas contains a significant amount of CO_2, and the mixed fuel gas has a heat content of 210 Btu (LHV)/scf. The gross power output of the combined cycle system is 819.6 MW (416 MW from the two gas turbines and 403.6 MW from the steam turbine) resulting in 675.9 MW of net export power.

The Subtask 2.3 optimized coal plant has a LHV thermal efficiency of 53.2% and an HHV thermal efficiency of 53.4%, both of which are based on the heating value of the F-T liquids, the byproduct sulfur, and the equivalent energy of the export power. These efficiencies are about 11 to 12% greater than those of the Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant. The thermal efficiencies of the optimized Subtask 2.2 coke coproduction plant are about 2 to 3% higher than those of the optimized coal plant.

The Subtask 2.3 Optimized Coal Power Plant with Liquids Coproduction produces 12,377 bpd of liquids at the expense of about 479 MW less power production compared to Subtask 1.6 Nominal 1,000 MW Coal IGCC Power Plant. With 30 $/bbl liquids the return on investment of the Subtask 2.3 exceeds 10% only when power prices are above 40 $/MW-hr. However, at these power prices, the Subtask 1.6 IGCC power plant has a higher return on investment. Therefore, the opportunity for a domestic coal based gasification power plant with liquid fuel precursors coproduction appears to be limited in today's economic environment unless there are special circumstances, such as the use of a low priced feedstock. However, future improvements in gasification area availability from future design enhancements, advanced gas turbines developments, and improved Fischer-Tropsch reactor performance could make a coal gasification power plant with liquids coproduction economically competitive in the future.

Appendix A

Subtask 2.3

Major Equipment List

Appendix A

Major Equipment List

Table A1 lists the major pieces of equipment and systems by process area in the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction. Detailed equipment lists for systems that would be purchased as complete units from a single vendor, such as the Air Separation Unit, are not available.

Table A1
Major Equipment of the Subtask 2.3 Optimized Coal Gasification Power Plant with Liquids Coproduction

Fuel Handling – 100
Unit Train Rail Loop
Rotary Coal Car Dumper
Rotary Car Dumper Coal Pit
Rotary Dumper Vibratory Feeders
Rotary Dumper Building & Coal Handling Control/Electrical Rooms
Rotary Car Dumper Dust Collector
Rotary Car Dumper Sump Pumps
Coal Car Unloading Conveyor
Coal Crusher
Reclaim Coal Grizzly
Coal Storage Dome
Reclaim Conveyors
Storage/Feed Bins
Reclaim Pit Sump Pumps
Coal Dust Suppression System
Coal Handling Electrical Equipment and Distribution
Electric Hoist
Metal Detector
Magnetic Separator
Vibrating Feeder
Slurry Preparation – 150
Weigh Belt Feeder
Rod Charger
Rod Mill
Rod Mill Product Tank
Rod Mill Product Tank Agitator
Rod Mill Product Pumps
Recycle Water Storage Tank
Recycle Water Pumps
Slurry Storage Tank
Slurry Storage Tank Agitator
Slurry Recirculation Pumps
Solids Recycle Tank
Solids Recycle Tank Agitator
Solids Recycle Pumps
Rod Mill Lube Oil Pumps
Slurry Feed Pumps (1^{st} Stage)
Slurry Feed Pumps (2^{nd} Stage)

Fischer-Tropsch – 200 and 201
First Activated Carbon Adsorption Bed
Second Activated Carbon Adsorption Bed
Third Activated Carbon Adsorption Bed
F-T Reactor Steam Drum
3-Phase Overhead Flash Drum
Unconverted Syngas Wash Column
Sour Water Flash Drum
Wax L/V Separator
Wax Vapor L/V Separator
Liquid Fuel L/V Separator
Wax Mixer Surge Drum
Catalyst Slurry Mixing Tank
Catalyst Pretreater
Pretreated Catalyst Feed Tank
Catalyst Pretreater Overhead KO Drum
First F-T Reactor Vapor Overhead Flash Drum
Refrigeration KO Drum
Refrigeration Liquid Receiver
IP Steam Flash Drum
LP Steam Flash Drum
Reactor Vapor / BFW Exchanger
Reactor Vapor Water Cooler
HP Fuel Gas Steam Heater
Wax Vapor Air Cooler
Liquid Wax / BFW Exchanger
Liquid Fuel Water Cooler
Wax Cooler Heat Exchanger
Catalyst Pretreater Wax Heater
Cat Pretreater Feed/Effluent Exchanger
Catalyst Pretreater Overhead Cooler
Catalyst Pretreater Circulating Gas Heater
Refrigerant System Condenser
HP Fuel Gas Refrigeration Recovery Exchanger
Reactor Vapor Refrigeration Cooler
F-T Reactor Feed Preheater
Second Gas Turbine Fuel Heater (IP Steam)
Fuel Gas Compressor
Cat Pretreater Circulating Gas Compressor
Refrigeration Compressor
Liquid Fuel Recycle Pumps
BFW Circulation Pumps
Liquid & Catalyst Return Pumps
Wax Pumps to Filter
Wax Product Pumps
Clean Wax Pumps
Wax Recovery Pumps
Pretreated Catalyst to Reactor Pumps
Syngas Wash Tower Recirculation Pumps
Storage Tank Pumps
F-T Slurry Bed Reactor

Overhead Vapor Cyclone
Liquid Catalyst Hydroclone
Liquid Catalyst Cleanup Filter
Wax Catalyst Filters
Makeup Catalyst Feed Hopper Baghouse
Makeup Catalyst Feed Hopper
Catalyst Pretreater Baghouse
F-T Product Storage Tank
ASU & Gasifier Area Cooling Water - 250
Cooling Water Circulation Pumps
Cooling Tower (S/C)
Gasification - 300
Main Slurry Mixers
Second Stage Mixer
Gasifier Vessel
Post Reactor Residence Vessel
High Temperature Heat Recovery Unit (HTRU)
Cyclone Separators
Slag Pre-Crushers
Slag Crushers
Reactor Nozzle Cooling Pumps
Crusher Seal Water Pumps
Syngas Desuperheater
Nitrogen Heater
Pressure Reduction Units
Dry Char Filters
Cyclone Solids Pickup Vessel
Filter Solids Pickup Vessel
Syngas Scrubber Column
Syngas Scrubber Recycle Pumps
Slag Handling – 350
Slag Dewatering Bins
Slag Gravity Settler
Slag Water Tank
Slag Water Pumps
Gravity Settler Bottoms Pumps
Slag Recycle Water Tank
Slag Feedwater Quench Pumps
Slag Water Recirculation Pumps
Polymer Pumps
Slag Recycle Water Cooler
LTHR/AGR – 400
Syngas Recycle Compressor
Syngas Recycle Compressor Knock Out Drum
Syngas Heater
COS Hydrolysis Unit
Amine Reboiler
Sour Water Condenser
Sour Gas Condensate Condenser
Sour Gas CTW Condenser
Sour Water Level Control Drum
Sour Water Receiver
Sour Gas Knock Out Pot
Sour Water Carbon Filter

MDEA Storage Tank
Lean Amine Pumps
Acid Gas Absorber
MDEA Cross-Exchangers
MDEA CTW Coolers
MDEA Carbon Bed
MDEA Post-Filter
Acid Gas Stripper
Acid Gas Stripper Recirculation Cooler
Acid Gas Stripper Reflux Drum
Acid Gas Stripper Quench Pumps
Acid Gas Stripper Reboiler
Acid Gas Stripper Overhead Filter
Lean MDEA Transfer Pumps
Acid Gas Stripper Knock Out Drum
Acid Gas Stripper Preheater
Amine Reclaim Unit
Condensate Degassing Column
Degassing Column Bottoms Cooler
Sour Water Transfer Pumps
Ammonia Stripper
Ammonia Stripper Bottoms Cooler
Stripped Water Transfer Pumps
Quench Column
Quench Column Bottoms Cooler
Stripped Water Transfer Pumps
Degassing Column Reboiler
Ammonia Stripper Reboiler
Syngas Heater
Sulfur Recovery – 420
Reaction Furnace/Waste Heat Boiler
Condensate Flash Drum
Sulfur Storage Tank
Storage Tank Heaters
Sulfur Pump
Claus First Stage Reactor
Claus First Stage Heater
Claus First Stage Condenser
Claus Second Stage Reactor
Claus Second Stage Heater
Claus Second Stage Condenser
Condensate Level Drum
Hydrogenation Gas Heater
Hydrogenation Reactor
Quench Column
Quench Column Pumps
Quench Column Cooler
Quench Strainer
Quench Filter
Tail Gas Recycle Compressor
Tail Gas Recycle Compressor Intercooler
Tank Vent Blower
Tank Vent Combustion Air Blower
Tank Vent Incinerator/Waste Heat Boiler
Tank Vent Incinerator Stack

Gas Turbine / HRSG – 500
Gas Turbine Generator (GTG), GE 7FA+e, Dual Fuel (Gas and Syngas) Industrial turbine set, Including: Lube Oil Console, Static Frequency Converter, Intake Air Filter, Compressor, Turbine Expander, Generator Exciter, Mark V Control System, Generator Control Panel and Fuel skids.
GTG Erection (S/C)
Heat Recovery Steam Generator (HRSG) - Dual Pressure, Unfired, with Integral Deaerator
HRSG Stack (S/C)
HRSG Continuous Emissions Monitoring Equipment
HRSG Feedwater Pumps
HRSG Blowdown Flash Tank
HRSG Atmospheric Flash Tank
HRSG Oxygen Scavenger Chemicals Injection Skid
HRSG pH Control Chemicals Injection Skid
GTG Iso-phase Bus Duct
GTG Synch Breaker
Power Block Auxilary Power XformerS
Steam Turbine Generator & Auxiliaries - 600
Steam Turbine Generator (STG), Reheat, TC2F, complete with lube oil console
Steam Surface Condenser, 316L tubes
Condensate (hotwell) pumps
Circulating Water Pumps
Auxiliary Cooling Water Pumps
Cooling Tower
ASU – 800
Air Separation Unit Including:
Main Air Compressor
Air Scrubber
Oxygen Compressor
Cold Box (Main Exchanger)
Oxygen Compressor / Expander
Liquid Nitrogen Storage
Balance Of Plant - 900
High Voltage Electrical Switch Yard (S/C)
Common Onsite Electrical and I/C Distribution
Distributed Control System (DCS)
In-Plant Communication System
15KV, 5KV and 600V Switchgear
BOP Electrical Devices
Power Transformers
Motor Control Centers
River Water - Makeup Water Intake and Plant Supply Pipeline
Water Intake System S/C Including;
Intake Structure
Pumphouse
Makeup Pumps
Substation & Motor Control Center (MCC)
Lighting, Heating & Ventilation
Makeup Water Treatment Storage and Distribution
Water Treatment Building Equipment ;
Hydroclone Clarifier
Coagulation Storage Silo

Clarifier Lime Storage Silo
Gravity Filter
Clear Well
Clear Well Water Pumps
Water Softner Skids
Carbon Filters
Cation Demineralizer Skids
Degasifiers
Anion Demineralizer Skids
Demineralizer Polishing Bed Skids
Bulk Acid Tank
Acid Transfer Pumps
Demineralizer - Acid Day Tank Skid
Bulk Caustic Tank Skid
Caustic Transfer Pumps
Demineralizer - Caustic Day Tank Skid
Firewater Pump Skids
Waste Water Collection and Treatment
Oily Waste - API Separator
Oily Waste - Dissolved Air Flotation
Oily Waste Storage Tank
Sanitary Sewage Treatment Plant
Wastewater Storage Tanks
Reverse Osmosis Unit for Chloride Removal
Waste Water Outfall
Monitoring Equipment
Common Mechanical Systems
Shop Fabricated Tanks
Miscellaneous Horizontal Pumps
Auxiliary Boiler
Safety Shower System
Flare
Flare Knock Out Drum
Flare Knock Out Drum Pumps
Chemical Feed Pumps
Chemical Storage Tanks
Chemical Storage Equipment
Laboratory Equipment

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