

Early Entrance Coproduction Plant Quarterly Report No. 2

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Abstract:

The overall objective of this project is the three phase development of an Early Entrance Coproduction Plant (EECP) which produces at least one product from at least two of the following three categories: (1) electric power (or heat), (2) fuels, and (3) chemicals. The objective is to have these products produced by technologies capable of using synthesis gas derived from coal or coal in combination with some other carbonaceous feedstock.

The objective of Phase I is to determine the feasibility and define the concept for the EECP located at a specific site and to develop a Research, Development, and Test Plan (RD&T) for implementation in Phase II.

The objective of Phase II is to conduct RD&T as outlined in the Phase I RD&T Plan to enhance the development and commercial acceptance of Coproduction technology that produces high-value products, particularly those that are critical to our domestic fuel and power requirements. The project will resolve critical knowledge and technology gaps on the integration of gasification and downstream processing to coproduce some combination of power, fuels, and chemicals from coal and other feedstocks.

The objective of Phase III is to develop an engineering design package and a financing plan for an EECP located at a specific site.

The project's intended result is to provide the necessary technical, economic, and environmental information that will be needed to move the EECP forward to detailed design, construction, and operation by industry.

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II. List of Acronyms

The following acronyms were used in this report:

AFDU.....	Alternate Fuels Development Unit
AGR	Acid gas removal
API	American Petroleum Institute
ASU	Air Separation Unit
CCP.....	Commercial Coproduction Plant
DOE	United States Department of Energy
EECP.....	Early Entrance Coproduction Plant
F-T.....	Fischer Tropsch
GE	General Electric Power Systems
GT	Gas turbine
HCU	Hydrocracking Unit
HRSG.....	Heat Recovery Steam Generator
KBR	Kellogg Brown & Root, Inc.
LS.....	Low sulfur
LS FO.....	Low sulfur fuel oil
LS VGO	Low sulfur vacuum gas oil
MDEA.....	Methyldiethanolamine
PAR.....	Motiva’s Port Arthur Refinery
PUC.....	Public Utilities Commission
RD&T	Research, Development & Test
SAP	Sulfuric Acid Plant
SBCR	Slurry Bubble Column Reactor
SMR	Steam Methane Reforming
SRU.....	Sulfur Recovery Unit
TECO	Tampa Electric Company
TGTU.....	Tail Gas Treating Unit
ULS	Ultra low sulfur
USGC.....	United States gulf coast
VGO.....	Vacuum gas oil
WGS.....	Water gas shift

III. Executive Summary

This is the second of five quarterly reports which summarize the progress of Phase I of the development of the Early Entrance Coproduction Plant (EECP) concept which is covered by DOE Cooperative Agreement No. DE-FC26-99F-T40658. Phase I objective is to determine the feasibility and define the concept for the EECP located at a specific site and to develop a Research, Development, and Testing Plan. Phase I is scheduled for completion by the end of the year 2000. Phase II is to conduct the research as outlined in Phase I and is scheduled for two calendar years (2001 through 2002). Phase III is scheduled for the calendar year 2003 and is to develop an engineering design package and financing plan for the EECP. The overall project's intended result is to provide the necessary technical, economic, and environmental information needed to move the EECP forward to detailed design, construction, and operation by industry.

During this reporting period, work was performed in several areas. Preliminary "basis for design" packages were developed for two potential EECP sites. One site represents a typical refinery (Motiva's Port Arthur refinery) and the other a typical power plant (Tampa Electric's Polk Power Station). These basis of design documents include process descriptions, block flow diagrams, heat & material balances, sized equipment lists, utility-catalyst-chemical summaries, etc. Two cases using different product slates were developed for each site. For the Port Arthur refinery, each case produced power, steam, hydrocracked diesel and naphtha and the second case also produced finished wax. At Polk Power Station, each case maximized power production but one case also produced syncrude while the other case produced hydrocracked diesel and naphtha. Capital cost estimates for all four cases were started late in the quarter and are expected to be complete early in the second quarter. To support this work, process studies were conducted in the following areas:

- Fischer-Tropsch (F-T) Product Upgrading options
- F-T waste water treating / utilization
- F-T feed gas optimum H₂:CO ratio
- F-T tail gas utilization
- Percent CO₂ in F-T feed gas
- Acid gas removal scheme

Additionally, market assessment work began with an analysis of product slates and valuations.

Work in the upcoming quarter will include completion of the capital and operating cost estimates, selection of an EECP site to be the basis for further study, and commencement of more detailed process studies, as well as market, economic, and environmental assessments.

IV. Results, Discussion, and Preliminary Conclusions

Task 1 – Project Management Plan

The Project Management Plan was prepared in the 4Q1999 and submitted for DOE review. The Plan was updated in early January, 2000 to incorporate comments from DOE and the final version issued on January 12, 2000. DOE approved the Plan on January 26, 2000. Since then, the Plan has been used to status the progress of the work. Please refer to the cost curves and schedule update in sections VII and VIII of this report.

Task 2 – Concept Definition, Development & Technical Assessment

The objective of this task is to perform generic (non-site specific) and site specific studies to further define and develop the Early Entrance Co-production Plant (EECP) concept and select a site. The generic studies: F-T Reactor Feed Gas Optimum H₂/CO Ratio, F-T Product Upgrade Options, and F-T Wastewater Treating/Utilization that were initiated in 4Q1999 were completed. New generic studies: F-T Reactor Feed Gas Percent CO₂, and F-T Catalyst/Wax Separation were initiated and completed. New “site specific” studies: Site Selection Study, Acid Gas Removal, Process and Heat Integration, and F-T Tail Gas utilization were initiated. The following is a summary and status of these studies:

Generic Process Studies

F-T Reactor Feed Gas Optimum H₂ to CO Ratio

The main objective of this study is to determine the optimum F-T feed gas H₂/CO ratio for the EECP. The EECP contains several major process units, including a Gasification Unit, Air Separation Unit, Acid Gas Removal & Sulfur Recovery Unit, F-T Synthesis Unit, F-T Product Upgrading Unit, and Power Generation Unit. Individual cost of these units and how these units are integrated for various H₂/CO ratio influences the economics of the EECP.

The F-T Synthesis section converts syngas (a mixture of CO and H₂) into hydrocarbon liquids using either a cobalt-based catalyst or iron-based catalyst. For the cobalt-based catalyst, the general consensus is a H₂/CO ratio of 2 in the syngas feed to F-T is desirable. For the iron-based catalyst, due to water gas shift reaction, there is uncertainty of the optimum H₂/CO ratio. When petroleum coke is fed into the gasifier, the syngas produced from the gasifier has a low H₂/CO ratio of 0.6 to 0.8. One option is to send the syngas into the F-T unit without making any adjustment to the H₂/CO ratio. This option is called the Low H₂/CO ratio feed gas. The other option is to raise the H₂/CO ratio in the syngas to close to 2; similar to what is used for cobalt based F-T synthesis and then send it to the F-T Synthesis Section. This option is called High H₂/CO ratio feed gas. There are a number of ways to raise the H₂/CO ratio, three of which were examined in this study.

Following is a brief description of all the cases that were studied:

- Case A: Low H₂/CO ratio, no H₂/CO adjustment is made to the syngas produced from gasifying petroleum coke. The syngas feed to the F-T reactor has a H₂/CO ratio of 0.76.
- Case B: High H₂/CO ratio, the syngas produced from the gasifier is sent to a shift reactor, which shifts a portion of CO present in the syngas into H₂, thus raising the H₂/CO ratio to 1.9.
- Case C: High H₂/CO ratio, the syngas produced from the gasifier is mixed with H₂-rich syngas produced from a natural gas Steam Methane Reforming Unit (SMR). The combined syngas has a H₂/CO ratio of 1.9
- Case D: High H₂/CO ratio, the syngas produced from the gasifier is mixed with pipeline H₂ to reach a H₂/CO ratio of 1.9.

For this study, the preliminary conclusion is that the lower hydrogen to carbon monoxide ratio (0.76 as compared to 1.9) syngas is more economical for an Early Entrance Coproduction Plant when the Fischer Tropsch Synthesis is based on use of iron based catalyst. This conclusion is based on the economic analysis that was done for all four cases. The economic analysis shows that Case A, the low H₂/CO, case has highest rate of return and greatest net present worth. This is primarily due to the cost of hydrogen being sourced from natural gas for the other cases. There are currently limited data available at lower end of H₂/CO (0.6 to 1.4) ratio syngas to evaluate its impact on EECP economics. It is recommended that Phase II work include a determination of an optimum lower H₂/CO ratio (0.6 to 1.4) syngas for Fischer Tropsch Synthesis. The low H₂/CO ratio syngas feed (as produced in the Gasification section) without any H₂/CO ratio adjustment will be used for the completion of remaining tasks of the Phase I feasibility study.

F-T Reactor Feed Gas percent of CO₂

The purpose of this study is to evaluate the impact of the presence of carbon dioxide on the performance of the Fischer Tropsch Synthesis section and on the Early Entrance Coproduction Plant. There are currently no experimental data available to evaluate this impact on Rentech F-T Synthesis section. It was proposed to use an in-house numerical model developed using information available in literature to estimate the optimal amount of CO₂ in the syngas feed to the F-T Synthesis section.

Figures 1 and 2 show the CO conversion and yield of hydrocarbons (including alcohol, acids and light tail gases) as a function of the percentage of CO₂ in the feed. Since the residence time in a typical slurry column reactor is rather low, it is assumed that there is no hydrocarbon production from the conversion of CO₂, and only the additional flow will affect the reactor performance. As the amount of CO₂ in the feed is increased, the total flow rate of gas increases, and for a fixed size of reactor, the superficial gas velocity also

increases. Therefore, the residence time of the syngas in the reactor decreases producing lower conversions and yields.

It is possible that the additional CO₂ induces a reverse water gas shift (WGS) reaction to occur. This reaction will produce more water (relative to syngas) lowering the F-T reaction rate (inhibition by water) and yielding even lower conversions and yields than previously expected. These lower values as a function of the percentage of CO₂ in the feed are depicted as dashed lines in Figures 1 and 2. The most probable scenario lies somewhere between the two curves. Nevertheless, optimal reactor performance is obtained with no CO₂ in the feed. To validate these results, the effect of CO₂ will be determined in laboratory experiments outside the scope of this work, but its effect or lack of effect, will be confirmed in Phase II of the EECF project.

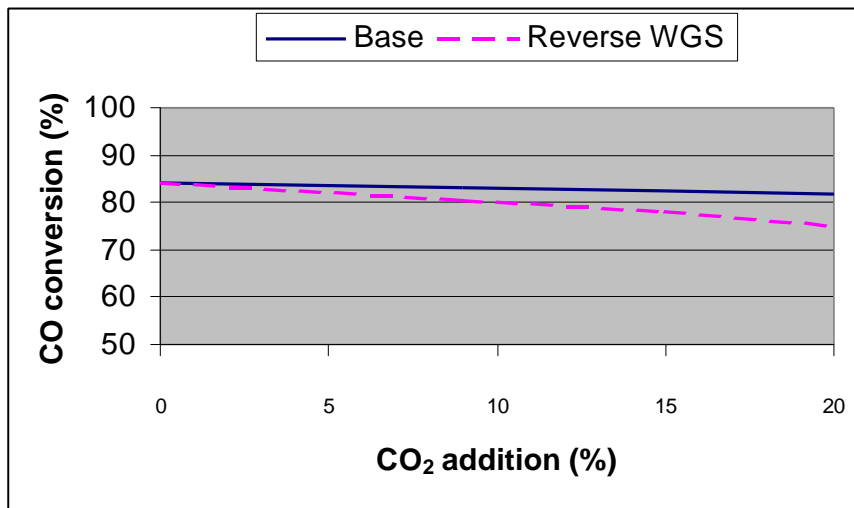


Figure 1: CO conversion as a function of CO₂ in the feed

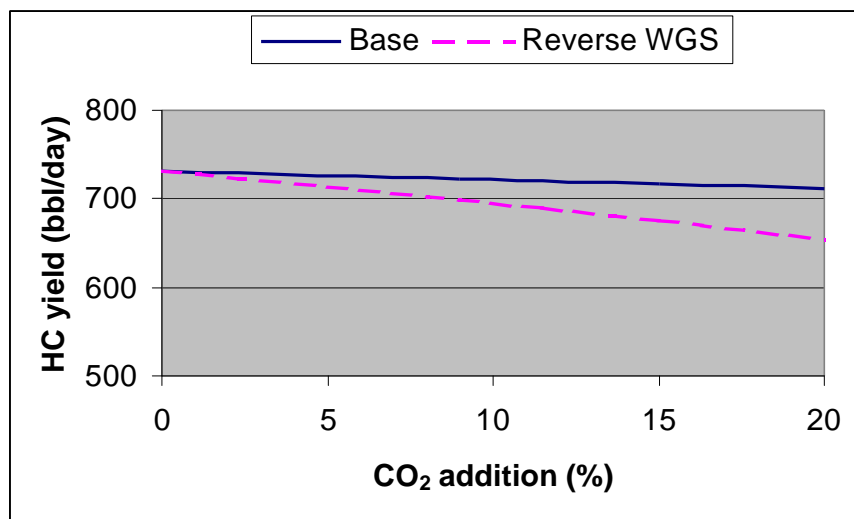


Figure 2: Hydrocarbon yields as a function of CO₂ in the feed

F-T Product Upgrade Options

The purpose for this study is to develop flow schemes and cost estimates for upgrading a nominal 2500 bpsd of F-T liquids. The information from this study is to be used to define the type of upgrading and corresponding products that would be economically attractive at selected sites, and for subsequent commercial implementation and marketing studies.

Various options for F-T products upgrading were reviewed for their suitability as part of the overall EECF/CCP concept. From these options, the following six cases were selected for further study:

- Stabilized Syncrude
- Dewaxed Syncrude
- Hydrocracker (Diesel) product
- Lube Oil Products
- Wax Products
- Alpha Olefins

The overall configurations, the advantages/disadvantages, yields and budget-type capital cost estimates were developed and summarized for these cases.

As shown in Figure 3, the cost of the F-T upgrading plant increases as the complexity of the process increases. The alpha olefin case may be slightly simpler and lower cost than the wax case, but requires more analysis and marketing research. Eventually the site location and the local marketing needs determine the desirable type of product upgrading. However, there is an incentive to take advantage of the low volume, high valued components inherent in the F-T liquids. Figure 4 supports a marketing strategy for high value specialty products (lube oil, wax, alpha olefins) rather than commodity transportation fuels. Niche market for premium “environmental” type diesel is a potential exception that can also be readily implemented.

In general, the F-T Liquids can be considered as premium feedstock with exceptional inherent qualities, such as negligible sulfur, nitrogen and aromatic content. Hydro processing is directionally the most suitable mode of product upgrading to isomerize (dewax), saturate the olefins and remove oxygenates. With hydrocracking, there is a diesel volume increase of over 10% and a cetane number of more than 70. For this kind of diesel there may be a niche market with public transportation companies. Also hydro processes for lube oils and/or waxes meets the low volume, high value product criteria and offers the best incentive.

FIGURE 3

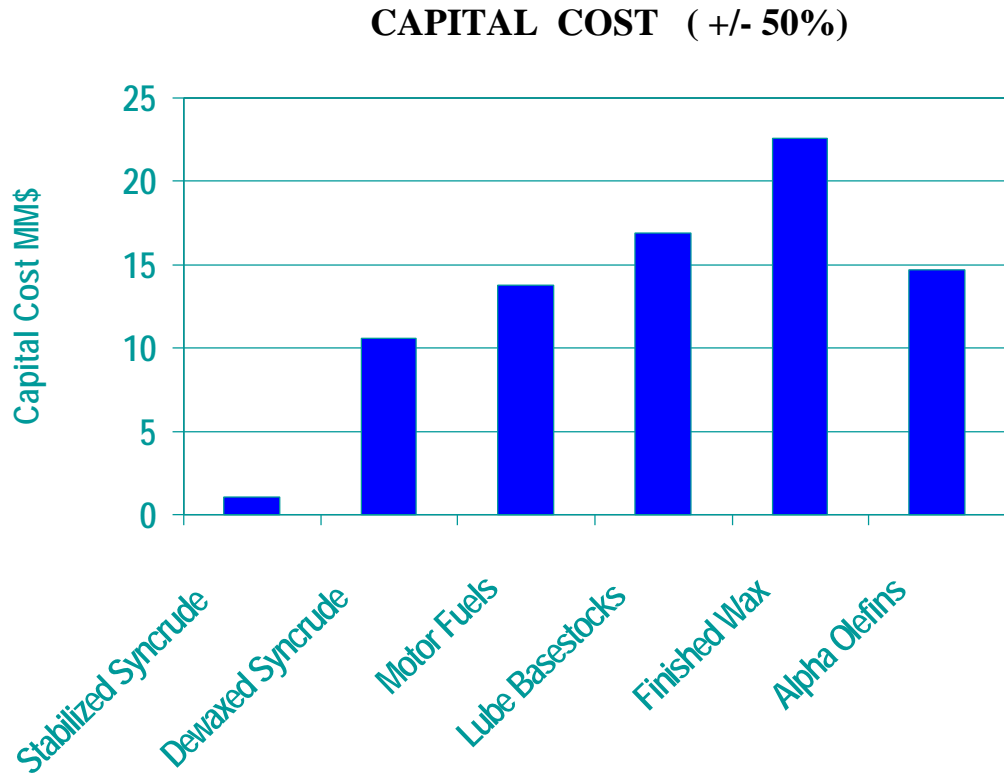
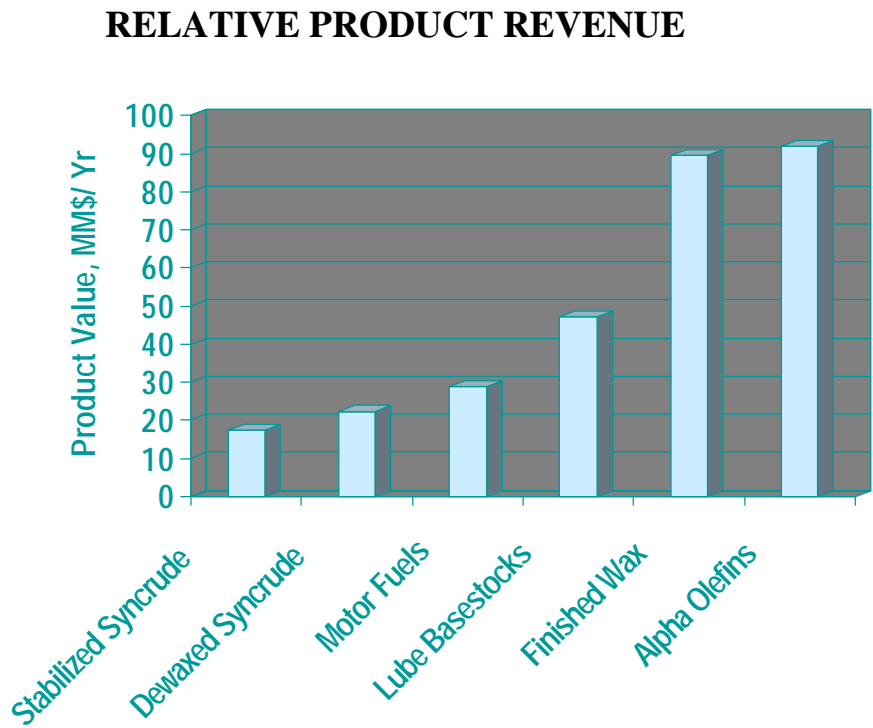


FIGURE 4



F-T Wastewater Treating/Utilization

The purpose of this study is to develop and evaluate options for treating or utilizing the wastewater stream generated by the F-T synthesis step.

The relatively small quantities of mixed alcohol and/or organic acids have little or no market value as chemicals or chemical feedstock.

Three main options (with some variations on some options) for treating and/or utilizing the F-T wastewater stream were evaluated. These included recycling to the syngas generation (Gasification) section, pretreatment to remove alcohol before sending the waste stream to the bio-treatment section, and sending the entire waste stream directly to the bio-treatment section.

Cost estimates were not developed for the different options, but analysis showed the recycling option to be superior in eliminating the waste stream by totally re-utilizing it in a manner that involves low capital and operating cost.

F-T Catalyst/Wax Separation

For low temperature F-T reactor operating conditions in a Slurry Bubble Column Reactor (SBCR), the heavier hydrocarbon products (wax) remain in liquid phase thus increasing the height of slurry catalyst bed with time. Therefore it is necessary to remove wax continually to keep the slurry height constant without removing catalyst from the reactor. Texaco does not have a proven commercial method for removing the catalyst from the wax. This is a critical issue that must be resolved. Currently, various separation methods are being evaluated outside DOE EECF funding. Prior to the detail design and construction of the EECF project, sufficient work will have identified an effective means to separate catalyst and wax. Texaco will demonstrate the effectiveness of the separation on a stand-alone system, and a small SBCR outside of DOE funding. Texaco will also privately fund construction and testing of a Demonstration Separator for catalyst/wax separation on the DOE AFDU at La Porte, Texas.

Site Specific Studies

Site Selection Study

As part of Task 6, two potential sites: (1) representative of a typical refinery application - the Motiva Port Arthur Refinery and (2) representative of a typical power generation facility - the Tampa Electric Company Polk Power Station, were selected. The purpose of this study is to develop the EECF concept tailored to these sites by the following: develop overall heat and material balances, develop capital and operating cost estimates, perform preliminary marketing, environmental and economic studies in order to select a site. This study was initiated this quarter and the following is the summary of the progress made in the 1Q2000.

A design basis was developed for both sites. A total of four concepts were developed. Overall heat and material balances were completed. Sized preliminary equipment lists and utility and catalyst/chemical summary were produced for gasification, acid gas removal, sulfur recovery, F-T synthesis and F-T product upgrading sections. Praxair provided oxygen plant, GE provided power plant and Monsanto provided H₂SO₄ plant information.

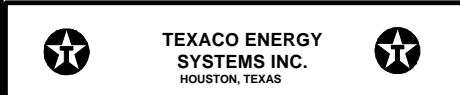
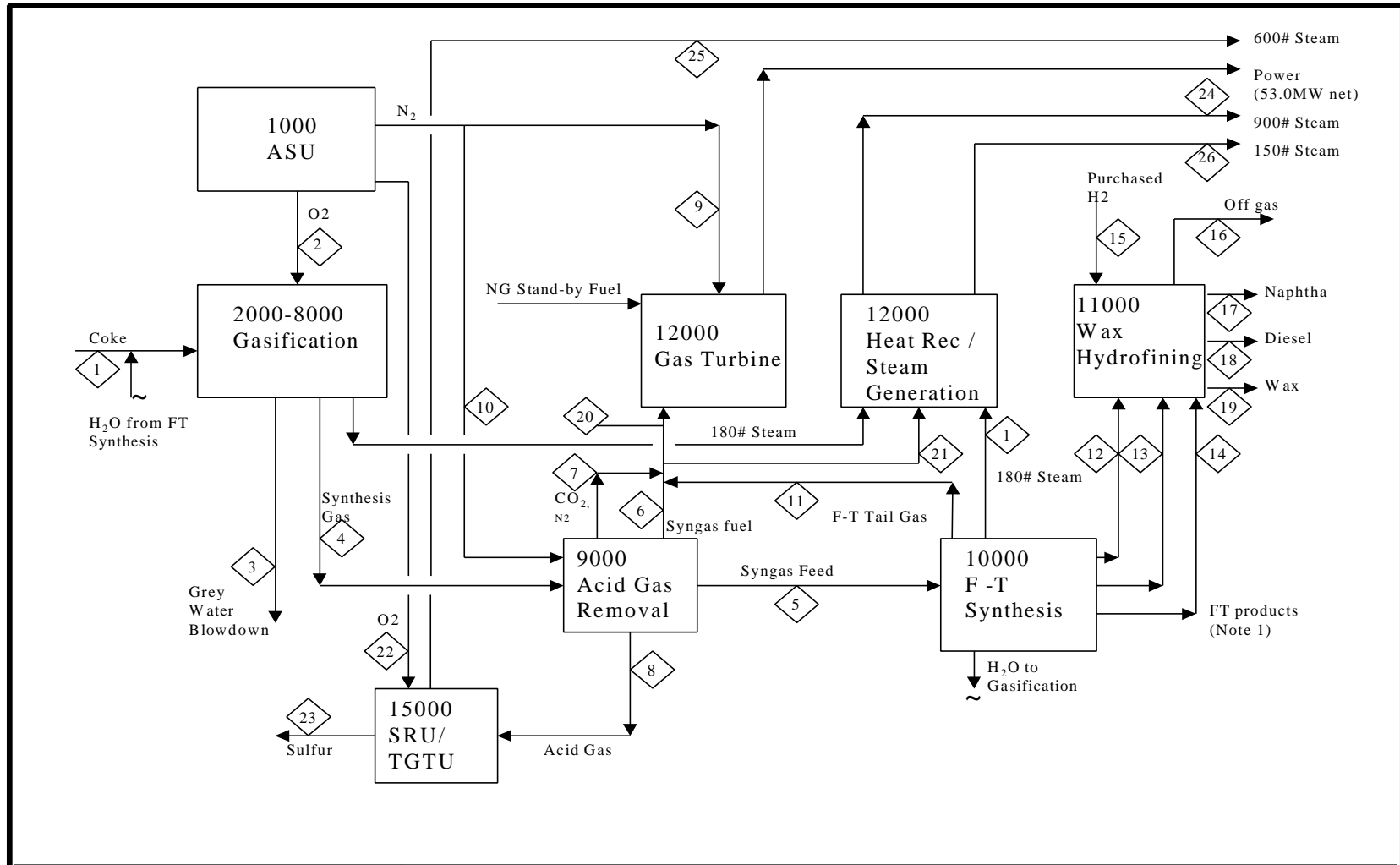
The proposed EECP is designed to process 45,926 kg/hr (1215 short tons per day) of petroleum coke to produce electric power and clean fuels. The following four cases were developed:

- Case PARFW:** Port Arthur Refinery Finished Wax. The hydrocarbon liquids produced from the F-T reactor are upgraded into three different products – hydrotreated naphtha & diesel, and finished wax.
- Case PARHCU** Port Arthur Refinery Hydrocracking Unit. The hydrocarbon liquids produced from the F-T reactor are sent to a Hydrocracker, which produces diesel and naphtha.
- Case TSC** Tampa Syncrude. The hydrocarbon liquids produced from the F-T reactor are dewaxed to produce synthetic crude.
- Case THCU** Tampa Hydrocracking Unit. The hydrocarbon liquids produced from the F-T reactor are sent to a Hydrocracker, which produces diesel and naphtha.

Following is a summary of the four cases:

Case	PARFW	PARHCU	TSC	THCU
Feed	Petroleum Coke	Petroleum Coke	Petroleum Coke	Petroleum Coke
Stand-by Fuel - GT	Natural Gas	Natural Gas	Diesel	Diesel
Auxiliary Fuel – HRSG	Natural Gas	Natural Gas & HCU Offgas	Natural Gas	Natural Gas & HCU Offgas
Products	Power Naphtha Diesel Finished Wax Sulfur 6307 kPa steam (900 psig steam) 4238 kPa steam (600 psig steam) 1136 kPa steam (150 psig steam)	Power Naphtha Diesel - Sulfur 6307 kPa steam (900 psig steam) 4238 kPa steam (600 psig steam) 1136 kPa steam (150 psig steam)	Power Syncrude - - Sulfuric Acid - - -	Power Naphtha Diesel - Sulfuric Acid - -

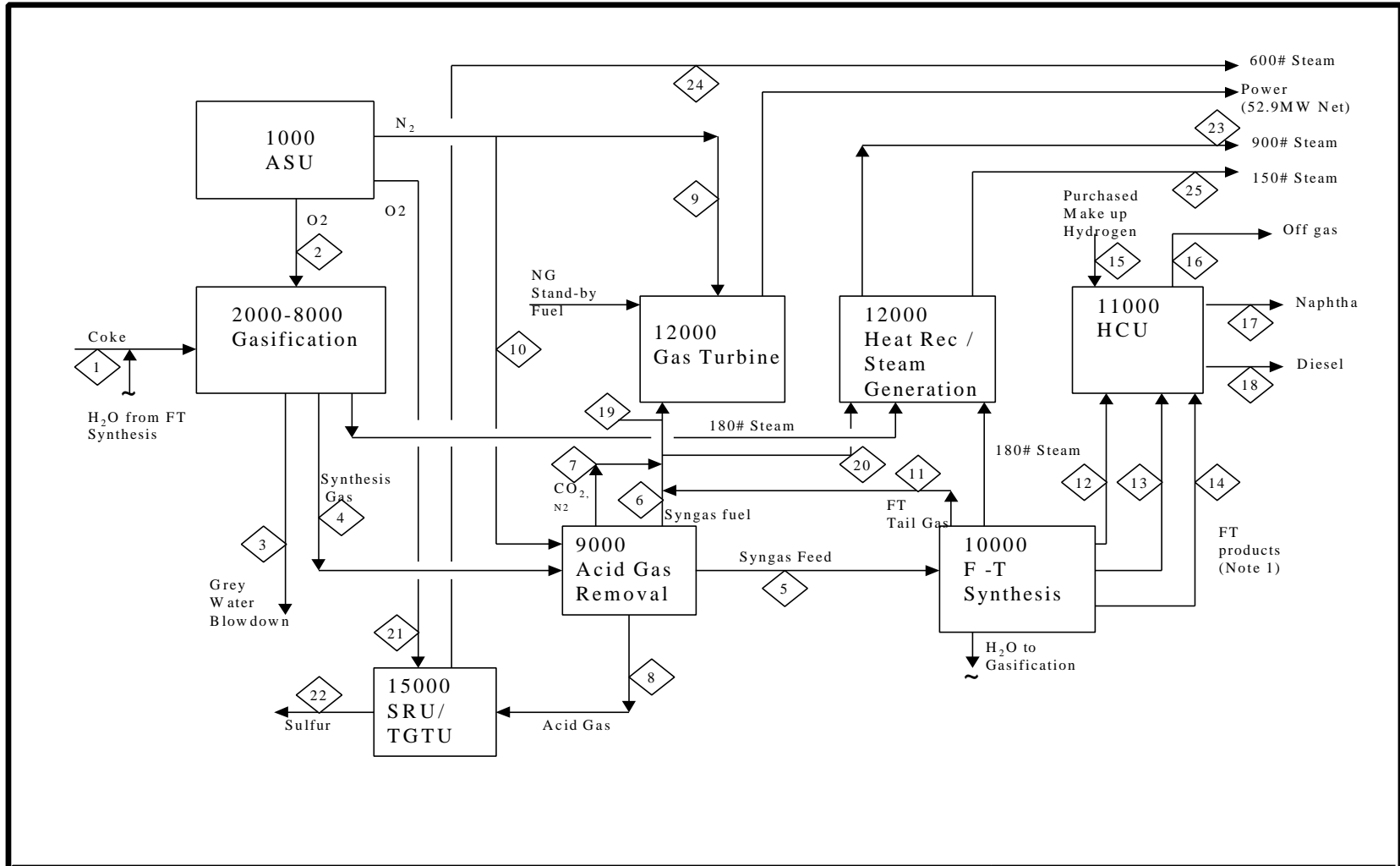
Refer to the following simplified block flow diagrams of the plant which show the major processing units and flow of primary streams and utilities.



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NOTES:
 1. Includes Light, Medium, and Heavy FT Liquid product streams

DOE EARLY ENTRANCE COPRODUCTION PLANT PORT ARTHUR REFINERY BLOCK FLOW DIAGRAM CASE: PARFW (FINISHED WAX)			
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CHECKED	DATE	TASK NO:	
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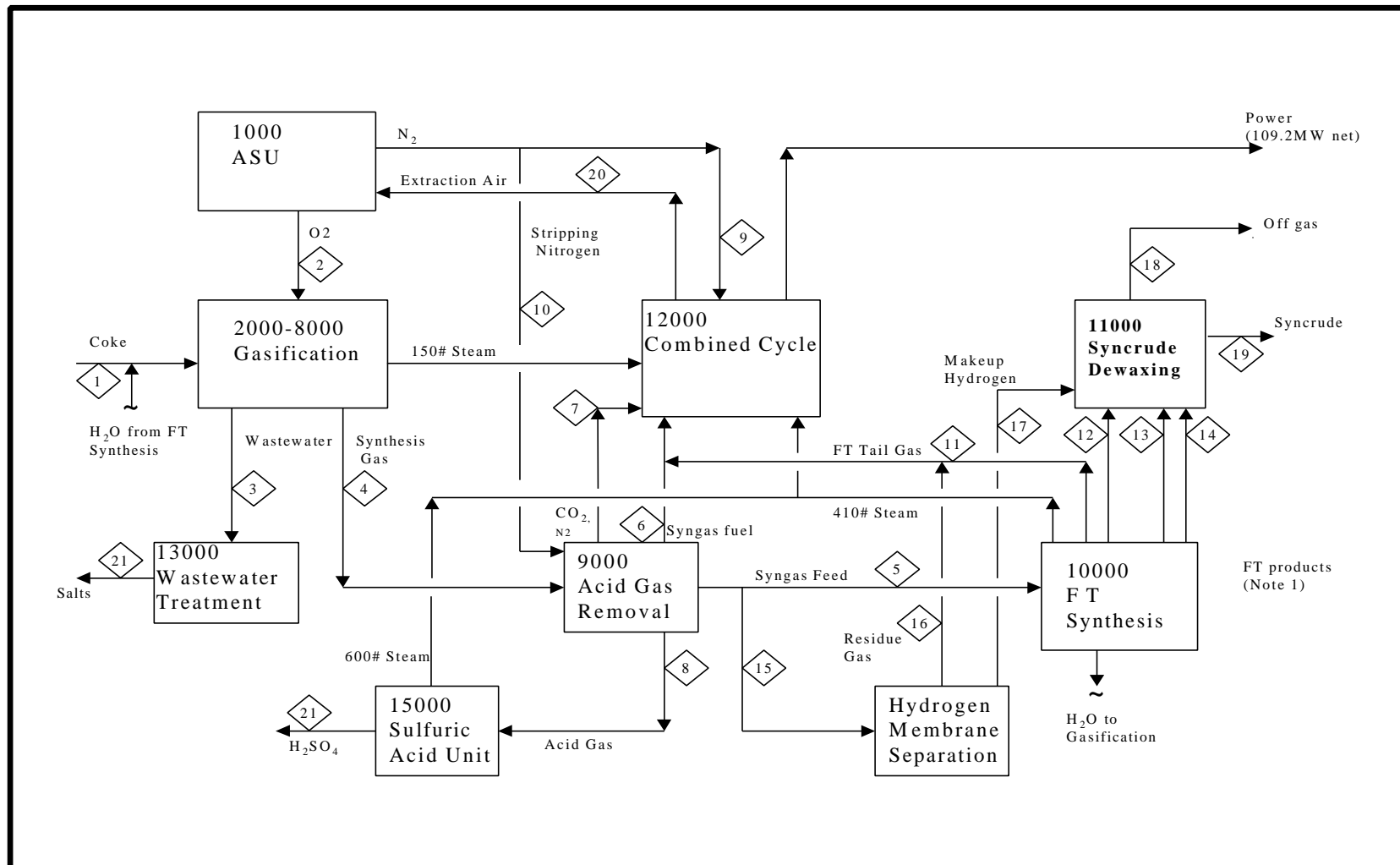
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NOTES:
1. Includes Light, Medium and Heavy FT Liquid product streams

**DOE EARLY ENTRANCE COPRODUCTION PLANT
PORT ARTHUR REFINERY
BLOCK FLOW DIAGRAM
CASE:PARHCU (HYDROCRACKING)**

DRAWN JSA	DATE 5/5/00	DOE AWARD NUMBER: DE-FC-99FT40658	REV.
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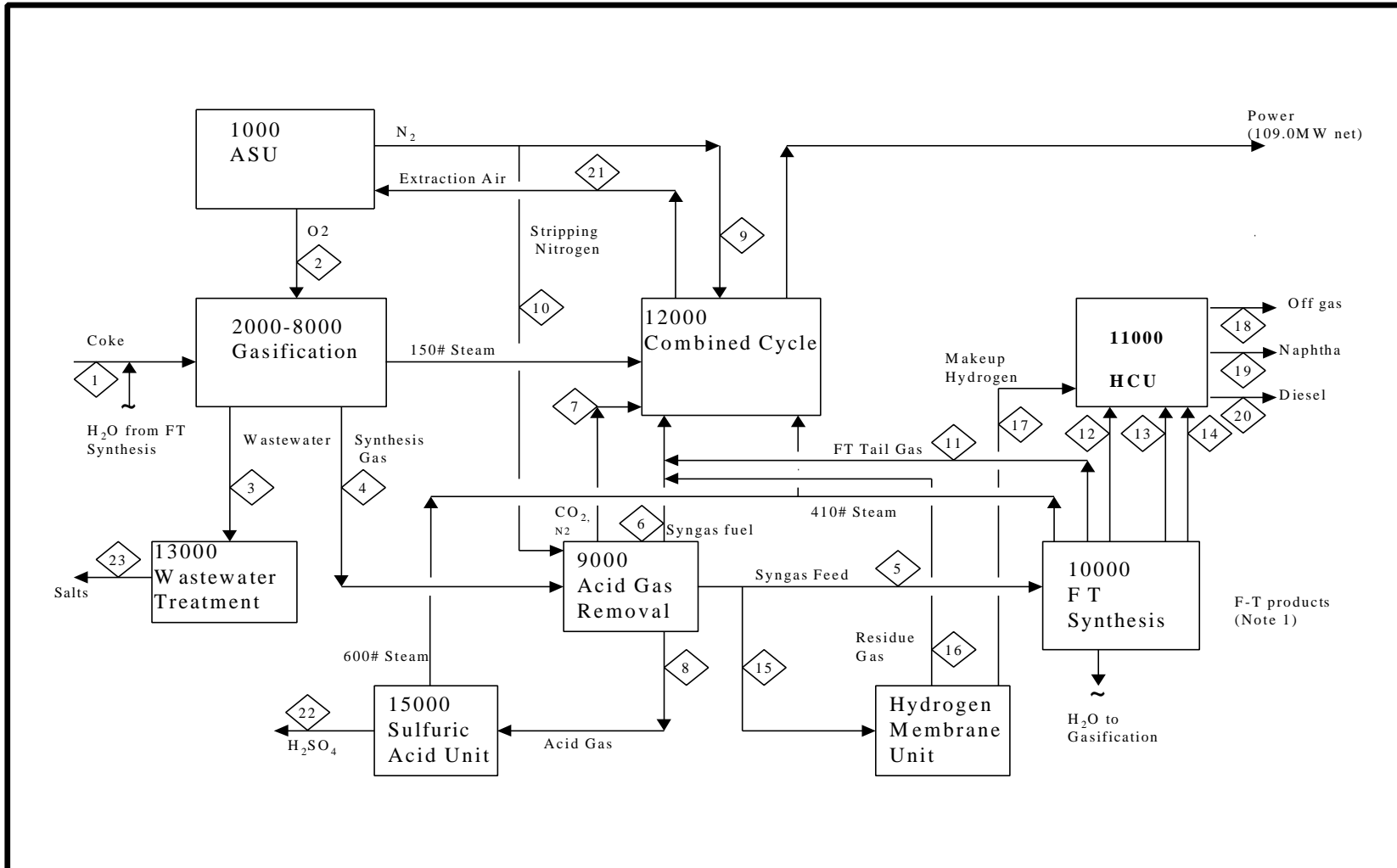



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NOTES:
1. Includes Light, Medium and Heavy FT Liquid product streams

DOE EARLY ENTRANCE CO-PRODUCTION PLANT TAMPA ELECTRIC POLK POWER STATION BLOCK FLOW DIAGRAM CASE: TSC (SYNCRUDE DEWAXING)			
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 <p>TEXACO ENERGY SYSTEMS INC. HOUSTON, TEXAS</p>	<p>NOTES:</p> <p>1. Includes Light, Medium and Heavy FT Liquid product streams</p>	<p>DOE EARLY ENTRANCE COPRODUCTION PLANT TAMPA ELECTRIC POLK POWER STATION BLOCK FLOW DIAGRAM CASE: THCU (HYDROCRACKING)</p>																												
<p>NOTICE</p> <p>THE INFORMATION HEREIN CONTAINED IS NOT PUBLIC BUT BELONGS TO TEXACO INC. WHICH HAS PROVIDED IT SOLELY FOR AN EXPRESSLY RESTRICTED PRIVATE USE. ALL PERSONS, FIRMS, OR CORPORATIONS WHO RECEIVE SUCH INFORMATION SHALL BE DEEMED BY THEIR ACT OF RECEIVING THE SAME TO HAVE AGREED TO MAKE NO DUPLICATION OR OTHER DISCLOSURE OR USE WHATSOEVER OF ANY OR ALL SUCH INFORMATION EXCEPT SUCH RESTRICTED USE AS IS EXPRESSLY AUTHORIZED IN WRITING BY TEXACO INC.</p>		<table border="1" style="width: 100%; border-collapse: collapse;"> <tr> <td style="width: 15%;">DRAWN</td> <td style="width: 15%;">JSA</td> <td style="width: 15%;">DATE</td> <td style="width: 15%;">5/5/00</td> <td style="width: 15%;">DOE AWARD NUMBER:</td> <td style="width: 15%;">DE-FC-99FT40658</td> <td style="width: 10%;">REV.</td> </tr> <tr> <td>DESIGNER</td> <td>DATE</td> <td>SUBTRACTOR PROJECT NO</td> <td colspan="4">J9202</td> </tr> <tr> <td>CHECKED</td> <td>DATE</td> <td>TASK NO:</td> <td colspan="4"></td> </tr> <tr> <td>APPROVED</td> <td>DATE</td> <td>DRAWING NO:</td> <td colspan="4">KBR-T-BFD-001</td> </tr> </table>	DRAWN	JSA	DATE	5/5/00	DOE AWARD NUMBER:	DE-FC-99FT40658	REV.	DESIGNER	DATE	SUBTRACTOR PROJECT NO	J9202				CHECKED	DATE	TASK NO:					APPROVED	DATE	DRAWING NO:	KBR-T-BFD-001			
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Process and Heat Integration Study

The EECP is composed of a number of main elements, such as the gasifier, acid gas removal (AGR) and sulfur recovery, F-T synthesis, air separation unit (ASU), power island, etc.

The members of the EECP consortium; KBR, Texaco, Praxair and GE supply these elements. One of the tasks of developing the EECP is to ensure optimum integration of its elements. This will be fully developed in the latter part of Phase I, but enough work was done in the current stage to ensure that the configurations developed for the two site locations have incorporated the benefits of the major integration possibilities.

The process integration features incorporated at this stage are:

- Recycle F-T waste water to the gasification section
- Use of nitrogen from the ASU for GT fuel dilution
- Use of nitrogen from the ASU as stripping gas in the AGR
- Use of bleed air from the gas turbine combined cycle as feed to the ASU
- Utilize oxygen from the ASU in the sulfur recovery unit (SRU)

In addition to process stream integration, the following general heat integration steps were taken:

- Low-grade heat from the gasifier was used to preheat boiler feed water (BFW) for the heat recovery steam generator (HRSG) section attached to the gas turbine.
- Steam header levels in the process portion of the EECP were set to match HRSG, steam turbine, and host site header levels to the maximum possible extent.
- Steam usage within the EECP was rationalized to minimize the number of levels at which the steam was exported or imported.

The approach to integration between the EECP and the host site was tailored for the particular site, thus:

- Port Arthur Refinery:

The philosophy was to export all available surplus steam from the EECP (process and HRSG) to the refinery steam header systems. The EECP would not have a condensing steam turbine. This resulted in the export of:

- 139,700 kg/hr of 6,307 kPa steam (308 k lb/hr of 900 psig steam)
- 5,400 kg/hr of 4,238 kPa steam (12 k lb/hr of 600 psig steam)
- 48,500 kg/hr of 1,135 kPa steam (107 k lb/hr of 150 psig steam)

- Tampa Electric Power Station:

All surplus steam generated within the process and HRSG segments of the EECPP was passed through a condensing steam turbine, which generated an additional 52 MW of power.

Acid Gas Removal Study

The purpose of this study is to develop the best technologies to remove CO₂, H₂S and COS from the gasification syngas before it enters the F-T synthesis unit and the combined cycle Power Island. The following sites were evaluated:

- Motiva's Port Arthur Refinery (PAR)
- Tampa Electric Company's Polk Power Station (TECO)

Sulfur recovery technologies were selected to match existing technology at each site: Modified Claus Sulfur unit with hydrogenation and amine-based tail gas treating at PAR and sulfuric acid plant at TECO.

Based on past project experiences, the following solvent configurations were selected for comparison to be used in the Acid Gas Removal (AGR) unit:

- Case 1 – Chemical solvent, methyldiethanolamine (MDEA) at 37.8°C (100 °F)
- Case 2 – same as above except at 23.9°C (75 °F)
- Case 3 – Physical solvent, Selexol™ at 37.8°C (100 °F)
- Case 4 – same as above except at -1.1°C (30 °F)

Cooling the solvent below 37.8°C (100 °F) requires the use of refrigeration. For both solvents, near identical process configurations, i.e., H₂S absorption, H₂S concentration, and solvent stripping sections, were selected based on high CO₂ diluent recovery for the gas turbine operations, very low-sulfur feed gas for F-T synthesis, and high H₂S content in the stripper offgas. The MDEA process requires an upstream COS hydrolysis reactor and a H₂S scavenger unit to meet the same total sulfur specifications achieved by the Selexol™ process.

Capital costs; initial catalyst and chemical costs; annual utility costs; and annual catalyst and chemical costs were developed for each of the four cases. The capital cost estimates were based on a ratioed, sized equipment list with a projected accuracy of +/- 35%.

For the PAR site, MDEA Case 2 was selected for the following reasons:

- approximately 2% lower capital cost,
- potential to further reduce capital costs by an additional 2% by combining AGR and TGTU amine strippers, although the impact of this on overall reliability needs to be assessed,
- MDEA is already in use at PAR.

For the Tampa site, MDEA Case 1 was selected for the following reasons:

- approximately 1-2% lower capital cost,
- MDEA is already in use at TECO.

The approximately \$10MM difference between the sites for overall sulfur removal and recovery is due mostly to the higher cost of the sulfuric acid plant at TECO.

Since the capital costs for both approaches, MDEA and Selexol™, are so close, the key factors such as steam, power and byproduct values must be confirmed to validate this study once a site is selected. Other mitigating site factors may actually dictate the final choice. However, for whichever site is selected, the minimum capital cost MDEA case is suggested as the initial base case.

F-T Tail Gas Utilization Study

The F-T tail gas comprise of the unconverted syngas, CO₂, and light hydrocarbons that are produced in the F-T reactor. There are many possible options for F-T tail gas usage. The optimum use will depend on the process scheme and the desired products, which can vary from site to site. Possible options for tail gas usage include: recycle to syngas generation, recycle to F-T feed, sent as feed to a second reactor, sent as fuel for power generation, sent as fuel for burners, etc.

There are important reasons for tail gas utilization: (1) the carbon efficiency will be increased by recycling the tail gas allowing for increased conversion and F-T liquid production; (2) process efficiency can be increased by recovering the energy content of the tail gas since it is very clean and free of sulfur and other contaminants; and (3) emissions into the atmosphere will be reduced. Therefore, it is important to determine the best utilization of the tail gas from both an economic and environmental viewpoint.

Reactors in Series

The concept of a second reactor in series scheme is to maximize hydrocarbon liquid production by converting as much of the available syngas to the desired products as possible. The increased yield achieved in these schemes is generally at the expense of more capital investment for larger reactors and higher operating cost for recycle compression.

In this scheme, the tail gas from the first reactor is fed to a second reactor. This is one way to maximize overall conversion of syngas to hydrocarbon liquids. Removing CO₂ in the tail gas before sending it to the second reactor will reduce the size of the second reactor. However the question of tail gas usage is not completely answered, because there is still a tail gas from the second reactor. This scheme only creates a smaller volume of tail gas to deal with. Depending on the site of the plant, this tail gas may be used as fuel for various heating needs. This scheme requires more capital investment in a second F-T reactor and a CO₂ removal unit if the CO₂ removal between the reactors in series option is desired.

Recycle to the F-T Reactor

The recycle concept is similar to the reactors in series. Instead of a second reactor, a recycle compressor compresses the tail gas containing unconverted syngas, CO₂ and light hydrocarbons and sends the tail gas back as feed to the F-T reactor. Light hydrocarbons, CO₂ and other inert gas will not contribute to the F-T reaction and will accumulate in the system due to the recycle. CO₂ can be removed in each pass and a bleed-off is required to prevent build up of gases such as nitrogen, argon, methane, etc. In commercial size plants, this bleed stream may be too large to flare or vent. Again the tail gas disposal issue is not completely resolved. A method must be developed to dispose of the bleed stream.

Another issue that must be considered with a recycle design is the H₂:CO ratio. In most cases, the H₂:CO usage ratio will not be the same as the feed H₂:CO ratio resulting in a different H₂:CO ratio in the tail gas. Recycling the tail gas can shift the combined feed H₂:CO ratio and can change the productivity of the reaction. Therefore when designing a recycle case, one must pay attention to the H₂:CO ratio in the feed to the F-T. This scheme requires more capital investment in a recycle compressor, larger F-T reactor and an additional CO₂ removal unit. It would also require operating cost to run the recycle compressor.

Recycle to the Syngas Generation Unit

The F-T tail gas can also be recycled to the syngas generation unit. In this scheme, the CO₂ and the light hydrocarbons in the tail gas can be converted back to H₂ and CO in the syngas generation unit. Therefore these species would not build up as they would in the recycle to the F-T reactor scheme. However, nitrogen and argon can still accumulate in the system if there is no bleed-off. Again this scheme, as in the other recycle scheme, is aimed at maximizing the liquid yield. One difference in this scheme is that by recycling the tail gas to the syngas generation unit, more syngas is being made available for conversion to hydrocarbon liquids. The other advantage of this scheme is that no CO₂ removal is required as in the recycle to the F-T reactor scheme. Because the tail gas is recycled to the syngas generation unit, this scheme will require a large capacity syngas generation. In most cases, the syngas generation unit is already the most capital intensive unit. This scheme would increase the capital investment in the syngas generation section as well as in the F-T section for larger reactors.

Fuel for Power Generation

The F-T tail gas can be burned in a gas turbine to generate power. In this scheme, instead of maximizing F-T liquid production, power is produced as another product. Power generation is a good option if the available power supply is not reliable. Therefore, in order to increase the operational reliability of the plant, power can be generated for internal use. Power generation is also a good option when there is a market for power

export. In this case power can be generated as a product and exported. This scheme requires the capital investment for the power generation unit.

Other Utilization Options

Within a plant, there may be various needs that the tail gas may fulfill. F-T tail gas can be used as fuel for burners although supplemental natural gas and/or special low BTU burner design may be required since the tail gas generally has a low heating value. If there is a need for hydrogen, the unconverted CO in the tail gas can be shifted to hydrogen then extracted with the rest of the hydrogen to be used in the plant. For example, hydrogen extracted from the tail gas can be used to provide the hydrogen for hydrocracking/hydrotreating. These demands are generally very small and will not be able to accommodate all of the F-T tail gas.

EECP Concept

In the concept of this EECP for both selected sites, Motiva Port Arthur Refinery and Tampa Polk Power Station, power and hydrocarbon liquids are the desired products. Therefore, the power generation scheme has been selected for all four concepts. Syngas resulting from the gasification of coke, after the required treating, is sent to the F-T for once-through high conversion to hydrocarbon liquids including light F-T liquids, medium F-T liquids and F-T wax. These liquid products are then sent to the F-T Product Upgrading Unit. The tail gas from the F-T unit, containing unconverted syngas, CO₂ and light hydrocarbons is delivered as fuel to the gas turbine to generate power.

The other options, including the ones listed above, are continuously being developed and studied outside the scope of this EECP study. The optimizations of each of these process schemes will require significant effort. Each specific case will require detailed study to find the best process scheme. Selection of the best tail gas usage option will depend highly on the location of the plant, the size of the plant and the possible synergies or restrictions of the specific site.

Tasks 5 and 6 – Market Assessment and Preliminary Site Analysis

Plant Site Selection Criteria

The plant site criteria list developed by the Technical Team in December, 1999 has been reviewed and analyzed to select the most crucial items which required evaluation by teams outside of the technical group to ensure the economic viability of the DOE Early Entrance Coproduction Plant (EECP) project. The following items were selected:

- The product slate and pricing
- The interest of the host site in having the EECP facility and their experience in operating similar unit operations
- The existing infrastructure and impact on product slate and capital investment
- The commercial benefit of the EECP facility to the host site

This section addresses the product slate and pricing for the site selected to represent a typical refinery application, the Motiva Port Arthur Refinery, and the site selected to represent a typical power generation facility, the Tampa Electric Company Polk Power Station.

Objective of section

- Provide a summary of the recommended product slate and the justification for the selected product slate for each site.
- Provide a summary of product prices and a summary of the methodology used to arrive at pricing for each site.

Background Information

The Early Entrance Coproduction Plant is a major project in DOE’s “Vision 21 PowerPlex” Program. The objective of the program is to reduce dependence on non-domestic reserves and to have cleaner burning fuels by 2010. The program will integrate emerging concepts for high-efficiency, fuel-flexible electrical power generation facilities and coproduction of cleaner burning motor fuels.

The Port Arthur Refinery was Texaco’s first refinery and has been updated through the years to maintain a competitive refining position and has recently removed surplus equipment, giving several appropriate sites for construction of the EECF facility. On January 1, 1989, Saudi Refining, Inc. purchased 50% of Texaco’s Port Arthur refinery as part of a joint venture with Texaco. This joint venture, known as Star Enterprise, includes two other Texaco refineries in Convent, LA and Delaware City, DE and all of Texaco’s marketing in southeastern U. S. In 1999, a new joint venture between Star Enterprise and Shell Oil Co. was formed under the name Motiva Enterprises. The facility is a modern complex refinery capable of processing heavier and higher sulfur crudes to gasoline and middle distillate as well as lubricating base oils. In the past, the facility has produced and marketed slack waxes. Crude is supplied to the refinery by ship and pipeline. Major units in the refinery include:

Crude Distillation	235,000 bpd	
Vacuum Distillation	106,000	
Delayed Coking	49,500	120,960 kg/hr (3200 tons/day)
Fluid Catalytic Cracking	83,000	
Catalytic Reforming	45,000	
Hydrocracking	18,000	
Hydrodesulfurization	166,000	
Alkylation	18,000	
Lube Oil Processing	17,550	

TECO Polk Station was designed under a DOE Cooperative Agreement to demonstrate the commercial application of the integrated gasification combined-cycle concept. The

project utilizes Texaco’s coal gasification process to produce syngas which fuels a General Electric gas turbine generator whose exhaust is integrated with a heat recovery steam generator and a steam turbine to produce 250 MW. The facility processes 86,940 kg/hr (2300 tons per day) of coal and is reported to have the lowest cost power generation of any base load station on TECO’s system.

Product Slate and Pricing Issues

Product type, production volumes, market demand, location factors, niche opportunities and transportation require adequate definition to achieve reasonable estimates of product value and long term markets. The Product Valuation Team was formed to consider and weigh each of these factors and develop the following:

1. The methodologies for arriving at realistic price premises for each product at each location
2. Price forecasts for each product at each location

Below is a description of products and the rationale for the product slate for each location:

Motiva Port Arthur Refinery

Case 1	Hydrotreated Fuels	146 bpd (31 Naphtha + 115 Diesel)
	Finished Wax	<u>405 bpd</u>
		551 bpd total

Case 1 takes advantage of a niche opportunity at the Motiva refinery. The refinery has produced a number of wax materials and has the capabilities of handling and transporting slack waxes. The decision as to which products will yield the greatest economic return depends on the ratio of the value of the products to additional investment and operating cost of producing them. (Note: An alternate to this case could be considered where F-T light and medium liquid could be sold to Motiva for them to hydrotreat and EECF provides Finished Wax. This case can be considered if PAR was the selected site and Finished Wax was a desired product slate)

Case 2	Butanes	43 bpd
	Hydrocracked Naphtha	157 bpd
	Hydrocracked/Finished Diesel	<u>465 bpd</u>
		665 bpd total

The product slate for case 2 is the product slate which can be utilized at the Motiva refinery with the minimum capital requirements while receiving a slight premium value for the quality of the product. A finished diesel can provide value as a blend stock to bring off-spec material into a higher value market.

low sulfur diesel being used as a blending component. A U. S. gulf coast refiner sells a diesel blending component termed hydrocracker bottoms. It sells about 17,000b/d of this stream in bulk quantities. Test results from an August 1999 cargo are:

Sulfur	10ppm
API	40.2
Pour Pt	-3.3 °C (26°F)
Flash Pt	82.8 °C (181°F)
Cloud Pt	1.1 °C (34°F)
Cetane Index	56.5
IBP	205 °C (401°F)
10% Pt	223.9 °C (435°F)
50% Pt	267.8 °C (514°F)
90% Pt	337.2 °C (639°F)
FBP	358.9 °C (678°F)

The refiner is currently selling this product for a price equal to diesel fuel. No premium or discount is being paid despite the lower sulfur content and higher cetane index. These specifications are fairly similar to what would be expected from hydrocracking the primary F-T product and as the move to lower diesel sulfur content continues, we estimate that a premium price would be \$3-7/bbl range.

Diesel Cetane Improver

Cetane improver additives are marketed by some suppliers (Ethyl, Nalco/Exxon and Octel America) to assist refiners in meeting the cetane specification of diesel. Due to the cost of the additive (about \$1,585/m³ or \$6/gal), the refiner would only use this additive on an infrequent basis. For instance, it could be used to correct an off-spec tank of diesel at a terminal. Hence the market for this additive is periodic and small.

Based on blending modeling, the expected value of the F-T diesel is approximately \$102/bbl which does not include any costs for handling or transporting the F-T diesel to a refiner. The refiner will not pay this amount for the F-T diesel for the same reason that the refiner does not routinely use the commercially available cetane improver additive. However, the refiner may see some advantage in using F-T diesel to improve and/or increase his overall diesel pool. If the product was sold strictly as a cetane improver and in drum containers, the product may have a value of \$100/bbl.

Slack Wax & Finished Waxes

Use published market prices

Base Oils (alpha olefins and Group III)

Use published prices

Electrical Power

The electric power price will be set based either on estimates from marketing studies or published PUC pricing tables.

Steam

Steam pricing will be based on the equivalent value of natural gas required to produce the steam since the EECF project will not be designed to meet the reliability levels required for abandoning any steam generation equipment at any site. Any steam streams at less than 1,135 kPa (150 psig) will have no value at any site. There will not be any steam exported to the Polk Station operation and any steam imbalance will result in the steam being condensed and returned as water to remove gasification and F-T exothermic heat.

F-T Product Pricing Basis

Crude Price: \$25/bbl WTI

\$24.80/bbl light low sulfur crude (36.6 API, 0.34% sulfur)

Refined Product Price		
	<u>\$/bbl</u>	<u>\$/kg, (\$/lb)</u>
Natural gasoline	24.75	
Refinery fuel gas	22.25	
Paraffinic naphtha	28.00	
Full range reforming naphtha	29.00	
Jet/kerosene	29.80	
Heating oil (No. 2)	27.00	
LS Diesel (0.05% S)	28.00	
ULS Diesel (0.005% S)	29.50	
LS VGO (1.0% S)	26.50	
LS FO (1.0% S)	21.50	
Low melt slack wax	65.00	0.44 (0.20)
High melt slack wax	130.00	0.88 (0.40)
Low melt finished wax	97.50	0.66 (0.30)
High melt finished wax	162.50	1.10 (0.50)

Note: ULS Diesel refers to the proposed 50ppm max sulfur being considered by the EPA starting about 2005.

The values are all FOB the plant. Any delivery or transportation costs are included in the values. These values are meant to be screening values only.

Site Specific Product Values**Motiva Port Arthur Refinery****Case 1-Finished Waxes with Hydrotreated Fuels as a Byproduct**

Hydrotreated naphtha and diesel will sold to the refinery and processed through a crude tower where the products will be valued as finished products less a refining margin.

Naphtha = Reforming naphtha – Octane debit – Refinery margin =

$$\text{Naphtha} = 29.00 - 4.25 - 2.00 = \$22.75/bbl$$

$$\begin{aligned} \text{Diesel} &= \text{LS Diesel} + \text{Sulfur Premium} - \text{Refinery margin} = \\ \text{Diesel} &= 28.00 + 1.50 - 2.00 = \$27.50/bbl \end{aligned}$$

Wax is assumed to be a finished wax that is a combination of low and high melt point wax and will be stored in a single heated tank.

$$\begin{aligned} \text{Wax} &= (\text{Low melt} + \text{High melt finished})/2 - \text{Mixing discount} = \\ \text{Wax} &= (97.50 + 162.50)/2 - 16.2 = \$113.75/bbl \text{ or } \$0.35/lb \\ &\quad (\$0.77/kg) \end{aligned}$$

Case 2-Hydrocracked Fuels

Hydrocracked naphtha is valued as a paraffinic naphtha FOB the refinery. Hydrocracked diesel is valued at ULS Diesel sold to the refinery or a third party. If the diesel does not meet both the pour point and cloud point specifications but is close to the specs for ULS diesel then its value will drop to LS Diesel.

$$\begin{aligned} \text{Naphtha} &= \text{Paraffinic Naphtha} = \\ \text{Naphtha} &= 28.00 = \$28.00/bbl \\ \text{Diesel} &= \text{ULS Diesel} = \\ \text{Diesel} &= 29.50 = \$29.50/bbl \text{ If meets all specs.} \\ \text{Diesel} &= 28.00 = \$28.00/bbl \text{ If high pour or cloud point.} \end{aligned}$$

Tampa Electric Company Polk Station

Case 1 F-T Syncrude Production

The syn-crude is valued versus LLS crude (reference previous page), a light low sulfur USGC crude oil (36.6 API, 0.34% Sulfur).

<u>Yields</u>	<u>LLS</u>	<u>HTL</u>	<u>delta</u>	<u>Price \$/bbl</u>
Gas	1.6	0	-1.6	22.25
Naphtha	22.2	4.0	-18.2	28.00
Kerosene	19.7	0	-19.7	29.80
Diesel	20.7	18.4	-2.3	28.00
VGO	26.1	77.6	51.5	26.50
1% FO	<u>9.7</u>	<u>0</u>	<u>-9.7</u>	<u>21.50</u>
Total	100	100	0	-0.40 delta x price

F-T Syn-crude value delivered to Gulf Coast refining center = LLS - \$0.40/bbl

Freight from Tampa to Louisiana is \$3.00/bbl in small quantities.

$$\begin{aligned} \text{Syn-crude} &= \text{LLS} - \text{delta} - \text{Freight} = \\ \text{Syn-crude} &= 24.80 - 0.40 - 3.00 = \$21.40/bbl \end{aligned}$$

Case 2 Hydrocracked Fuel

All of the hydrotreated or hydrocracked liquids are commingled and sold to Tampa Electric as a turbine fuel. This replaces LS Diesel fuel that is delivered to Tampa from the USGC.

$$\begin{aligned} \text{Naphtha/Diesel} &= \text{LS Diesel} + \text{Freight} = \\ \text{Naphtha/Diesel} &= 28.00 + 1.00 = \mathbf{\$29.00/bbl} \end{aligned}$$

V. List of Major Activities Accomplished in 1Q2000

The following list is provided as a brief summary of the work performed during this reporting period:

- Issued final version of Project Management Plan and received approval in January, 2000.
- Began contract development with partners.
- Began weekly meetings for coordination of process work among all parties.
- Selected a typical power plant and typical refinery for further study as potential EECF site.
- Completed the following process studies which began in 4Q1999:
 - F-T Product Upgrading options
 - F-T waste water treating / utilization options
 - F-T feed gas optimum H₂/CO ratio
- Initiated and completed the following process studies in 1Q,2000
 - Percent CO₂ in F-T feed gas
- Initiated and continued process studies for:
 - Site Selection study
 - Acid gas removal scheme
 - Heat and process integration
 - F-T tail gas utilization
- Prepared basis of design documents for the two potential sites
- Began market assessment and product valuation evaluation for both sites
- Established Business Team to evaluation markets, product valuations, perform economic calculations, etc.
- Began cost estimating activities

VI. List of Planned Activities for 2Q2000

The following list is provided as a brief summary of the work planned for the upcoming quarter:

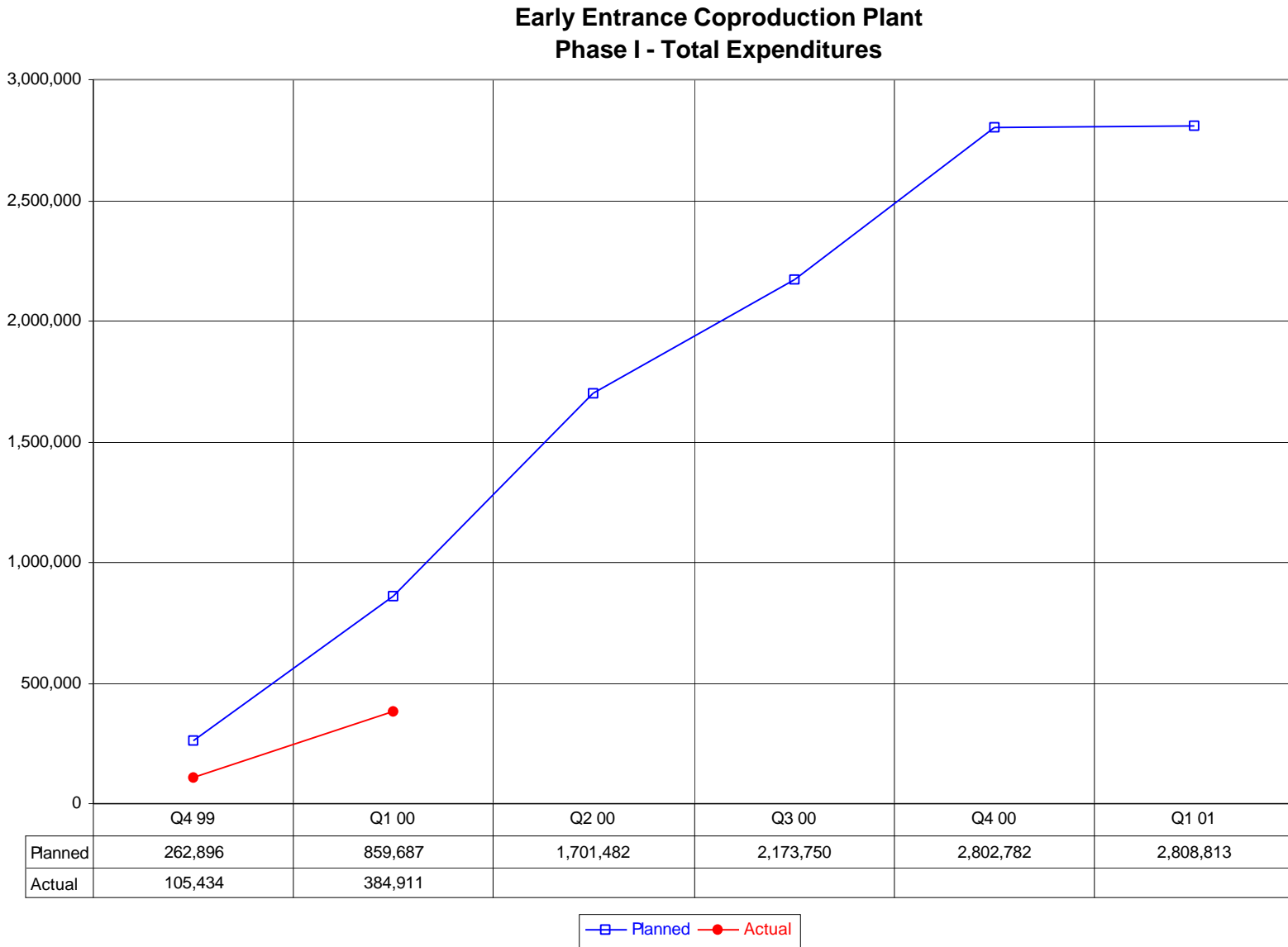
- Complete process studies for site selection work
- Complete cost estimates for the two sites
- Perform pro-forma calculations for each site
- Review basis of designs with each site
- Select a site for further EECP process studies
- Begin process studies for selected site
- Develop further the market, environmental, and economic assessments for selected site
- Perform technical assessment of subsystems
- Begin preparation of subsystem design specifications

VII. Graphs

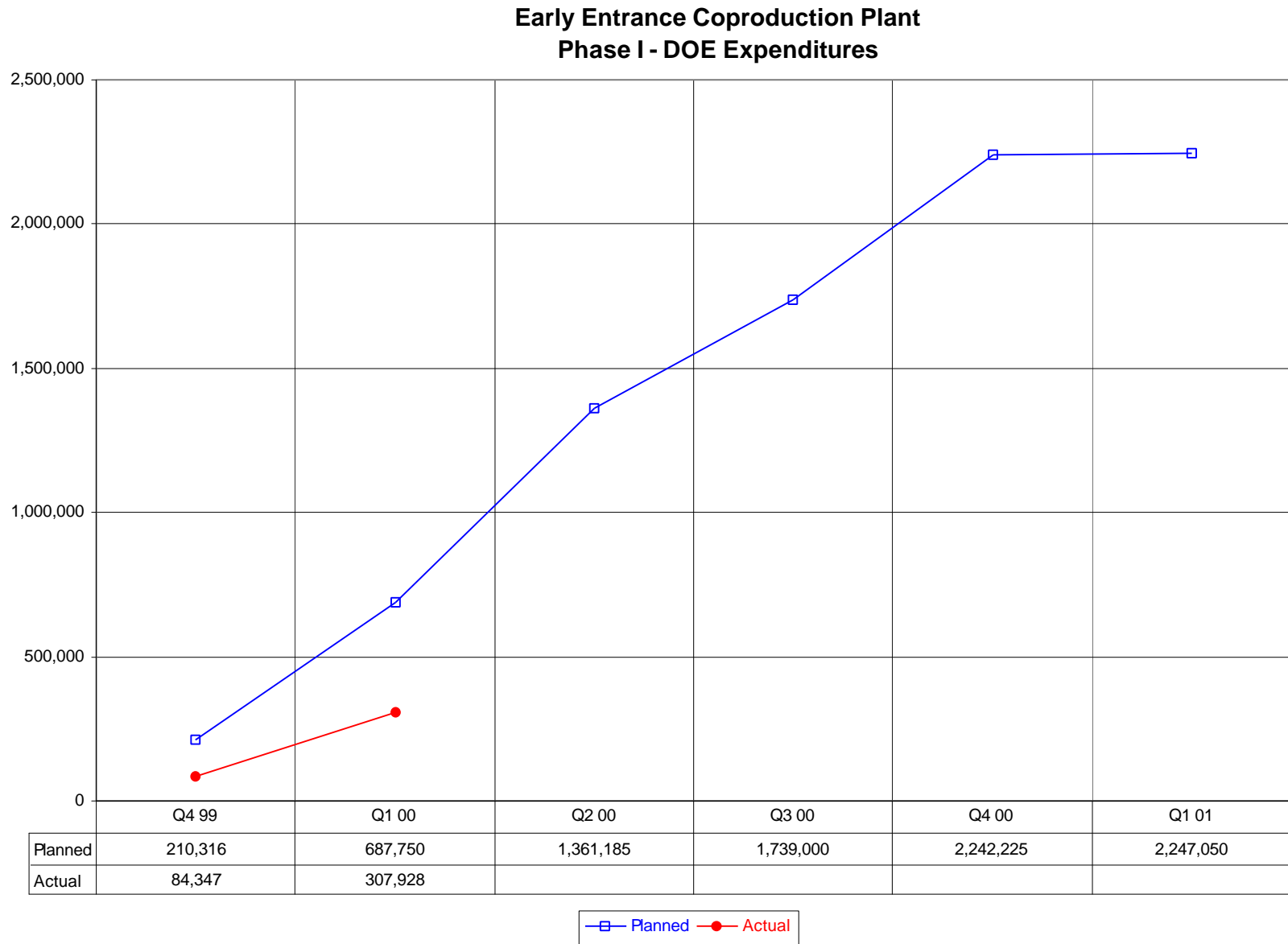
The following three graphs depict the financial status and progress of Phase I work. Graphs are shown on the following pages:

Planned vs. Actual Total Expenditures	34
Planned vs. Actual DOE Expenditures	35
Total Project Percent Complete	36

Phase I, Planned vs. Actual Total Expenditures

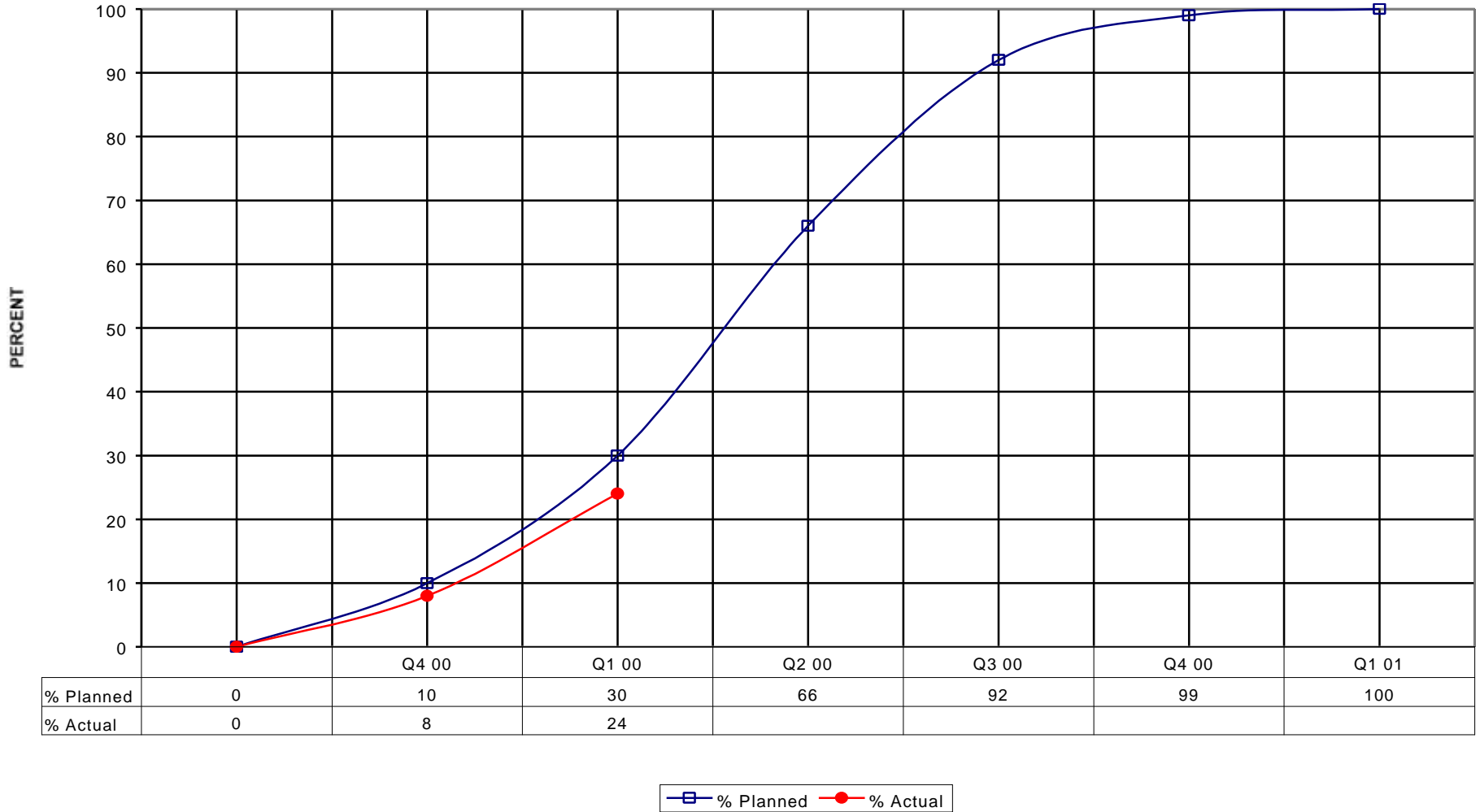


Phase I – Planned vs. Actual DOE Expenditures



Phase I – Total Project Percent Complete

Early Entrance Coproduction Plant
Phase I - Total Project % Complete



VIII. Schedule

The following two pages depict the Phase I project schedule and shows percent complete by task as of the end of 1Q1999. For a description of the work involved in each task, refer to the Cooperative Agreement. This schedule was prepared using MS Project 98 software.

Schedule:

ID	Task Name	% W/C	Q4 '99				Q1 '00			Q2 '00			Q3 '00			Q4 '00			
			Sep	Oct	Nov	Dec	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec	Jan
1	Phase 1: Concept Definition	24%	[Gantt bar from Sep '99 to Jan '00]																
2	Task 1 - Project Plan	100%	[Gantt bar from Sep '99 to Jan '00]																
3	1.1 Project Management Plan	100%	[Gantt bar from Sep '99 to Jan '00]																
7	Task 2 - Concept Definition, Development & Technical Assessment	58%	[Gantt bar from Oct '99 to Jan '00]																
8	2.1 Overall concept definition & development	87%	[Gantt bar from Oct '99 to Jan '00]																
54	2.2 Alternatives and options assessment and selection	67%	[Gantt bar from Oct '99 to Jan '00]																
64	2.3 Preliminary Block Flow Diagrams with mass and energy balance	72%	[Gantt bar from Jan '00 to Feb '00]																
69	2.4 Design considerations for advanced subsystems	63%	[Gantt bar from Feb '00 to Mar '00]																
73	2.5 Preliminary report	0%	[Gantt bar from Jun '00 to Nov '00]																
74	2.6 DOE review	0%	[Gantt bar from Nov '00 to Dec '00]																
75	2.7 Final report	0%	[Gantt bar from Dec '00 to Jan '01]																
76	2.8 Milestone - Issue final report	0%	[Gantt bar from Jan '01 to Jan '01]																
77	Task 3 - Subsystem Technical Assessment	0%	[Gantt bar from Apr '00 to May '00]																
78	3.1 ASU	0%	[Gantt bar from Apr '00 to May '00]																
79	3.2 Gasification	0%	[Gantt bar from Apr '00 to May '00]																
80	3.3 H ₂ :CO ratio adjustment	0%	[Gantt bar from Apr '00 to May '00]																
81	3.4 Fischer-Tropsch Synthesis	0%	[Gantt bar from Apr '00 to May '00]																
82	3.5 Gas Turbine	0%	[Gantt bar from Apr '00 to May '00]																
83	3.6 Steam system	0%	[Gantt bar from Apr '00 to May '00]																
84	3.7 Fischer-Tropsch product upgrading to market identifiable products	0%	[Gantt bar from Apr '00 to May '00]																
85	Task 4 - Subsystem Design Specifications	0%	[Gantt bar from Apr '00 to Sep '00]																
86	4.1 Fischer-Tropsch synthesis	0%	[Gantt bar from Apr '00 to Sep '00]																
92	4.2 Gas Turbine	0%	[Gantt bar from Apr '00 to Sep '00]																
99	4.3 Fischer-Tropsch product upgrading to market identifiable products	0%	[Gantt bar from Apr '00 to Sep '00]																
106	4.4 Risk assessment of integrated advanced subsystems	0%	[Gantt bar from Aug '00 to Sep '00]																
107	4.5 Design specifications for proven technologies	0%	[Gantt bar from Apr '00 to Sep '00]																
115	Task 5 - Market Assessment	19%	[Gantt bar from Sep '99 to Sep '00]																
116	5.1 Market analysis of products	50%	[Gantt bar from Oct '99 to Jun '00]																
117	5.2 Market analysis of technology	10%	[Gantt bar from Jan '00 to Sep '00]																
118	5.3 Product slate and quantities	30%	[Gantt bar from Apr '00 to Sep '00]																

ID	Task Name	% W/C	Q4 '99				Q1 '00			Q2 '00			Q3 '00			Q4 '00			
			Sep	Oct	Nov	Dec	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec	Jan
120	5.4 Full scale commercial plant	0%																	
121	Task 6 - Preliminary Site Analysis	93%																	
122	6.1 Site criteria	100%																	
123	6.2 Identify specific sites	100%																	
124	6.3 Identify additional commitments/parties of EECF participation	0%																	
125	Task 7 - Environmental Assessment	0%																	
126	7.1 Emission levels	0%																	
127	7.2 Adaptability for CO2 sequestration	0%																	
128	7.3 Water use and remediation	0%																	
129	7.4 Waste by-products	0%																	
130	7.5 NEPA requirements	0%																	
131	Task 8 - Economic Assessment	0%																	
132	8.1 Feed, fuel and product cost/price evaluation	0%																	
133	8.2 Cost estimates	0%																	
134	8.3 Role of government incentives for commercial viability of EECF	0%																	
135	Task 9 - Research, Development & Test Plans	0%																	
136	9.1 Design deficiency analysis	0%																	
137	9.2 Proposed test plan	0%																	
138	9.3 Preliminary report	0%																	
139	9.4 DOE review	0%																	
140	9.5 Final report	0%																	
141	9.6 Milestone - Issue final report	0%																	
142	Task 10 - Preliminary Project Financing Plan	0%																	
143	10.1 Preliminary Financing Report	0%																	
144	10.2 DOE review	0%																	
145	10.3 Final report	0%																	
146	10.4 Milestone - Issue final report	0%																	
147	Administration	40%																	