BASELINE DESIGN/ECONOMICS FOR ADVANCED FISCHER-TROPSCH TECHNOLOGY

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INTRODUCTION

This Advanced Fischer-Tropsch Indirect Liquefaction Study is conducted under DOE Contract No. DE-AC-91PC90027. The objectives of the study are to develop:

- A baseline design and several alternatives for indirect liquefaction using advanced Fischer-Tropsch (F-T) technology. The Baseline design uses Illinois No. 6 Coal and conventional refining. One alternative case uses ZSM-5 treatment of the vapor stream from the slurry F-T reactor. The other alternative case uses Wyoming coal from the Powder River Basin.

- The capital and operating costs for the above. Individual plant costs for the alternative cases will be prorated on capacity, wherever possible, from the baseline case.

- An ASPEN/SP Process Flowsheet Simulation (PFS) model and an economics spreadsheet model. Product valuation is based on analysis of the receiving refinery via linear programming. When the model is complete, sensitivity studies will be performed to demonstrate the effects of key independent process variables and economic assumptions.

The baseline design, along with the PFS and the economic spreadsheet analysis models, will constitute a major research planning tool that PETC can use to plan, guide and evaluate its ongoing and future research and commercialization programs relating to indirect coal liquefaction for the manufacture of synthetic liquid fuels.

A schematic diagram showing the overall program flow for achieving the above objectives is shown in Figure 1.

This paper covers the development of the Baseline design, including design basis, process selection, capital and annual operating costs. The development of the PFS model, and preliminary results on the F-T product evaluation and economic analysis will be discussed. The alternative cases are still under development.

OVERALL DESIGN BASIS

The Baseline design basis, including overall plant capacity and key process selection, was set at the outset of the project and continuously refined as the job progressed. Process selection around the Fischer-Tropsch synthesis loop was the subject of an intensive set of optimization studies, the conclusions of which are provided here. Equipment design criteria and plant site information generally follow the guidelines set in the direct liquefaction study[1], though there are certain aspects which are unique to indirect liquefaction.

Plant Capacity

The nominal baseline plant capacity was set at 50,000 BPD of liquid products, or roughly 20,000 stpd of coal feed, by considering the size of some of the downstream processing units. This size places processing plants such as the alkylation, isomerization and catalytic reforming units within the normal range of commercial sizes. In consultation with Shell Development Company, the maximum capacity of a single gasifier train is set at 2,540 stpd of as-received (8.6 wt% moisture) Illinois #6 coal.The coal is dried to 2 wt% moisture before gasification. Eight gasifier trains would thus consume 20,320 stpd of as-received coal, and this is set as the Baseline design capacity. For sensitivity studies, the design size range is set at 5,000 to 50,000 stpd of coal. Properties of Illinois #6 coal used for design purposes are shown in Table 1.

General Considerations

The indirect liquefaction facility is designed as a generic mine mouth plant located in southwestern Illinois. Perry county is identified as the most likely location. The facility will comply with all applicable environmental, safety and health regulations including EPA and Illinois air and water emissions regulations. Air cooling is maximized in order to minimize cooling water requirements, and thus, minimize the environmental impact.

A review of possible site locations and their proximity to water transportation lead to the conclusion that a vessel diameter of 17-feet OD could be transported by a combination of water transport and special land transportation. This sets the maximum size for shop fabrication and was used as a guide in setting maximum per train capacity of several process plants such as the CO_2 Removal and F-T Synthesis plants.

The facility is designed to use all of the byproduct steam and fuel gas production to supply a portion of the in-plant power requirement. Power is purchased when the requirements exceed the byproduct power production.

The only materials delivered to the plant are coal, n-butane, raw water, catalysts and chemicals. Coal is received by conveyor from the mine during two shifts per day, five days per week. Twenty eight days of raw coal storage are provided.

Provisions are made for seven days of raw water storage and seven days of n-butane storage. Other storage specifications are as follows:

Chemicals	30 days
Intermediates	2 days
Sour Water	5 days
Products	30 days
Sulfur	14 days

Plant availability studies were made which established that an overall on-stream factor of 90.8% was feasible provided that a spare gasifier train and a spare F-T reactor were incorporated into the facility design. A spare sulfur removal unit also was used in order to meet environmental restrictions.

BASELINE DESIGN

Overall Configuration

Figure 2 is a block flow diagram showing the overall process configuration. The facility is divided into three main areas:

1. Syngas production. Synthesis gas is generated in the Shell gasifiers from ground, dried coal. Processing of the raw syngas from the gasifiers is conventional, with wet scrubbing followed by single stage COS/HCN Hydrolysis and Cooling, Acid Gas Removal by inhibited amine solution and Sulfur Polishing. These units are supplemented by Sour Water Stripping and Sulfur Recovery units.

2. The Fischer-Tropsch synthesis loop. The synthesis loop includes Fischer-Tropsch (F-T) Synthesis, CO_2 Removal, Recycle Gas Compression/Dehydration, Hydrocarbon Recovery by deep refrigeration, Hydrogen Recovery for downstream usage and Autothermal Reforming. The Hydrocarbon Recovery Plant includes deethanization, depentenization, fractionation and an oxygenates wash column. At low H_2/CO ratios, CO_2 is the primary byproduct of the F-T reaction so a large CO_2 removal unit is required. In the Autothermal Reformer, unrecovered light hydrocarbons in the recycle gas are converted to additional syngas which raises the H_2/CO ratio to the F-T reactors.

3. Product upgrading. The downstream upgrading plants include Wax Hydrocracking, Distillate and Naphtha Hydrotreating for oxygenate removal and olefin saturation, Catalytic Naphtha Reforming, C_4 Isomerization, once-through C_5/C_6 Isomerization, C_3 /C_4/C_5 Alkylation and a Saturate Gas Plant. Liquid wax from the reactor, after catalyst recovery, is sent to the hydrocracker where high quality distillates are produced along with some naphtha and light ends. The naphtha, along with hydrotreated F-T naphtha, is catalytically reformed into aromatic gasoline blending components. The lighter materials are isomerized and alkylated into quality gasoline blending stocks.

Process Selection

Three key process selections were made on the basis of earlier studies by Bechtel and others. These three selections were: (1) syngas production via the Shell dry coal feed gasifier currently being demonstrated in integrated gasification combined cycle (IGCC) power plant applications but using byproduct CO_2 as the carrier gas to avoid inert buildup in the Fischer-Tropsch synthesis loop, (2) F-T synthesis using a slurry reactor with iron-based catalyst, and pilot plant data reported by Mobil Oil[2] and (3) F-T wax upgrading via mild hydrocracking, based on data reported by Mobil[2], PARC[3], and UOP[4].

The Shell gasifier was selected for its high efficiency. While it produces a gas which is substoichiometric in H_2/CO ratio, a study indicated that this could be compensated for by the addition of steam to the Fischer-Tropsch reactor while using a catalyst active for the water gas shift reaction (e.g. the precipitated iron catalyst used by Mobil).

The F-T slurry reactor is essentially a bubble column reactor where the liquid phase is a mixture of liquid wax and catalyst. The gas provides the necessary agitation for good mixing and mass transfer of reactants to, and products from, the liquid phase. A diagram of the reactor configuration, showing the placement of the internal cooling tubes, is given in Figure 3. The slurry reactor was chosen over the fixed-bed reactor for the Fischer-Tropsch section based on an earlier Bechtel study[5].

Slurry reactor sizing is based on conditions actually demonstrated by Mobil (10 cm/s inlet superficial velocity and 22.5 wt% catalyst), leaving higher velocities and concentrations to the sensitivity analysis. These conditions primarily affect the number of reactors. Bechtel's model is used for reactor sizing in the process simulation model, although consideration is also given to results from the Viking[6] model in establishing the baseline design. Maximum syngas conversion per pass is limited to about 82%.

In selecting mild hydrocracking for upgrading the F-T wax, a review was made of Mobil's comparison of mild hydrocracking with fluid bed catalytic cracking[2]. Results indicate that both processes are equally economical. The hydrocracking alternative was selected because the hydrocracked distillates thus produced are highly superior products and complement the distillates made directly in the F-T reactor. Fluidized bed catalytic cracking remains an interesting alternative for future studies. It produces a completely different slate of products with more gasoline, alkylate and, possibly, MTBE.

Before selection of the remaining process steps, a series of trade-off studies was performed. First, a detailed mathematical model of the Fischer-Tropsch reaction yields was developed based on the Mobil pilot plant data. With this model, a desired wax yield is selected and the operating temperature and complete product distribution are predicted. A spreadsheet design material balance model was developed for the F-T Synthesis Loop (also used to guide the development of the ASPEN/SP process simulation model). With these tools, the effects of the following variables and process steps were studied:

- Wax yield
- F-T pressure level
- Oxygen purity
- Inclusion of an autothermal reforming step in the F-T Synthesis Loop
- Method of recovering hydrogen for downstream use
- CO_2 removal process selection
- Temperature level for hydrocarbon recovery
- Single stage vs two stage COS hydrolysis

The conclusions were that the baseline study should use a 50 wt% wax yield, use 99.5 mole% oxygen, employ autothermal reforming, use PSA for hydrogen recovery, use inhibited amine for CO_2 removal, use deep refrigeration for hydrocarbon recovery and use single stage COS hydrolysis. There was insufficient information to define the true effect of pressure level and, consequently, pressure was set at a level dictated by the gasifier outlet pressure (403 psig) without further compression. The above findings set the overall Area 200 design guideline for the Baseline study.

The process design and the annual operating requirements for the Baseline study are documented in a Topical Report to be issued in October 1993.

<u>Products and Byproducts</u>

The Baseline case produces C_3 LPG, a C_5-350 °F naphtha and a 350-850 °F combined light and heavy distillate. The primary byproduct is liquid sulfur. A brief summary of the major feed and product streams to and from the Baseline plant is given in Table 2. The import power requirement is also shown.

The hydrocarbon products have no measurable sulfur or nitrogen. Oxygen is removed to less than 30 ppmv. There are virtually no aromatics in the distillate. Olefins are saturated to low levels of residual olefin concentration in both the naphtha and the distillate hydrotreaters. The diesel fraction has a very high cetane number, on the order of 73.

No product specifications were set other than boiling range. The quality of the products are dictated by the nature of the processing. The upgrading section design is premised on producing shippable products similar to conventional refinery streams from crude oil (i.e. low in olefins and oxygenates) and to produce quality gasoline blending stocks from recovered light ends. A high severity catalytic reformer is included since the Fischer-Tropsch naphtha is primarily low octane straight chain paraffins.

The naphtha product is a mixture of $C_3/C_4/C_5$ alkylate, C_5/C_6 isomerate and catalytic reformate. It is basically a raw gasoline with a clear (R+M)/2 octane number of about 88. Byproduct sulfur is produced and shipped as a high quality liquid product.

The products will be shipped to a refinery for possible further processing and blending into refinery product pools. While less upgrading is a possible alternative, typical refinery hydrotreaters and catalytic reformers would require a revamp in order to handle large concentrations of raw Fischer-Tropsch products. This arrangement would, however, reduce the capital investment of a F-T coal liquefaction plant.

No oxygenate byproducts are produced. Oxygenates which are soluble in the hydrocarbon fractions are converted to product in the hydrotreaters and the hydrocracker. Oxygenates soluble in the product water plus those removed from the light hydrocarbon vapors by water washing are combined and sent to a stripper where the light oxygenates are removed and sent to the fuel gas system. The stripped water is sent to the water treating unit where the remaining oxygenates are removed by biotreatment so that the water may be recycled.

The Baseline plant has an overall thermal efficiency of 57%. This efficiency is defined as the sum of the gross heating values of the liquid products divided by that of the inputs which include coal, n-butane and imported power. F-T product heating values were estimated from the API Technical Data Book, Figure 14A 1.1. A slight extrapolation was required. This is considered an excellent thermal efficiency, but should be considered preliminary until actual heats of combustion are measured.

Product Valuation

F-T products, other than the byproduct LPG and sulfur, are high premium blending components for gasoline and diesel fuels. In order to determine their relative values, Burns and Roe Services Corp., used Bechtel's linear programming modeling tool, PIMS (Process Industry Modeling Systems) to determine the values for the F-T products when they are blended with petroleum-derived stocks to produce specification gasoline and diesel fuels. The PIMS model was developed to represent a generic Midwest (PADD II) petroleum refinery in the year 2000 (the earliest a F-T liquefaction plant would possibly be constructed and operated). Provisions were provided to allow the refinery to expand the capacity of numerous upgrading operations such as hydrotreating and oxygenate production, in order to meet the stringent fuel specifications dictated by the Clean Air Act Amendments (CAAA) of 1990.

The properties of the crude oil used in the analysis (33° API gravity and 1.3 wt% sulfur) were projected from past data which indicate that the world's crude pools are becoming heavier and more sour. The product slate and pricing margins for the year 2000 were based on projections from SRI and Pace consultants, and the base crude price was set at 18.0 $/bbl.

The first scenario uses the CAAA 1990 Phase I standards and assumes that 15% of the gasoline pool in the Midwest will need to be reformulated and 83% of the diesel fuel will be low sulfur diesel. The second assumes additional regions will comply to the reformulation program such that 70% of the gasoline pool will be fully reformulated by the end of this decade. The Federal Phase II Standards which closely resemble the published California Phase II Standards is assumed.

For the first scenario, the ratio of the value of F-T naphtha blending stocks to crude oil was determined to be 1.50, and the ratio of the F-T diesel blending stock to crude oil was 1.38. For the second scenario which contains more stringent fuel specifications, these ratios were determined to be 1.56 and 1.40 respectively. Both scenarios found the F-T products to be premium components for blending. Their negligible sulfur content allows the refinery to reduce capital expenditures to meet future CAAA 1990 specifications. These results are shown in Table 3.

A detailed description of the F-T production valuation work will be documented in a separate report prepared by Burns and Roe Services Corp.

Capital and Operating Costs

In the development of the Baseline case, both licensed (proprietary) and open art technologies were used. A list of all the process plants considered in the Baseline design and the source of data used for their design purposes is given in Table 4.

For proprietary technologies, licensors or in-house Bechtel data were used. Only an overall material balance and utility summary are provided and the cost estimate is based on cost-capacity information. For process plants using open art technology or technology developed in the course of this study, a process flow diagram, material balance, utility summary and a four-line description for each piece of major equipment are provided. The total installed cost estimate is based on the detailed equipment list[7].

The OSBL plants were sized based on a combined utility summary, steam balance and water balance, storage requirements and plot layout. The cost estimate is based on proration from OSBL units in comparable past Bechtel jobs, including the Direct Coal Liquefaction Baseline Study[1]. Considerable care has been taken to assure that the cost estimate is, as far as possible, comparable to that study. The breakdown of the estimated capital cost for the Baseline design is given in Table 5.

ECONOMICS

Baseline Design Assumptions

An economic analysis of the Baseline design was carried out using the LOTUS 1-2-3 spreadsheet discounted cash flow model developed in the Direct Coal Liquefaction Study. The key assumptions used are summarized in Table 6. In order to realize a 15% return on equity, the results of this analysis indicate that the required selling prices of the F-T naphtha and distillate are 46.1 and 42.4 $/bbl respectively.

Results also are expressed in term of an Crude Oil Equivalent (COE) price which is defined as the price of crude oil when the Baseline design will have the specified 15% return of equity. For the Baseline study, the COE equals to 30.7 $/bbl.

Sensitivity Analysis

An economic sensitivity analysis was carried out to determine the impact of capital cost, coal feed price, F-T product premium, owner's equity, plant operation factor and maintenance cost on the resultant equivalent crude oil price. These results also are included in Table 7. As shown, a change in capital cost by 10% changes the Crude Oil Equivalent (COE) price by 1.8 $/bbl, and a 25% change results in a change of 4.6 $/bbl in COE. The effect of coal cost is somewhat smaller; 0.7 $/bbl and 1.6 $/bbl for 10% and 25% changes respectively.

As the owner's equity is increased from 25% to 50%, the Baseline COE is increased by 2.9 $/bbl. The Baseline case assumes a 100% operation of the entire complex during the first three years of plant operation. A reduction in operational efficiency to 25% during the first year increases the COE by as much as 3.1 $/bbl. Increasing the maintenance cost, which as defined in the spreadsheet does not include maintenance labor cost, from 1% to 2% of initial capital, increases COE by 1.3 $/bbl.

The value of a F-T product can be expressed as a ratio of its selling price to the crude oil price, as described in the Product Valuation section. An increase in this F-T product premium results in a drop in the Crude Oil Equivalent price. Under the CAAA Phase 1 scenario, a 20% increase in the F-T naphtha and distillate blend premiums decreases the COE by 2.8 $/bbl and 2.7 $/bbl respectively.

PROCESS FLOWSHEET SIMULATION (PFS) MODEL

Amoco, as the main subcontractor, has developed a process flowsheet simulation model for indirect liquefaction similar to that developed for the Direct Coal Liquefaction Study. It is designed to predict the effects of key process variables on the overall material and utility balance, operating requirements and capital costs. This model is implemented in the PC version of ASPEN/SP.

Development of the Model

Baseline design information was transmitted to Amoco for the development of the PFS model. Information transfer was expedited by Bechtel's use of a preliminary ASPEN/SP model for the design of the F-T synthesis loop.

As in the Direct Liquefaction Study, the computer model is intended as a planning/research guidance tool for the DOE and its subcontractors. The model is not designed to be a plant design and sizing program for every plant in the complex. The F-T synthesis loop design is handled in some detail, and Bechtel's reactor sizing and yield models are built into the design. For other plants, only overall yield, utility requirements and costs are handled. Costs are prorated on capacity using cost-capacity exponents and information on maximum and minimum single train capacity.

All ISBL plants shown in Table 3 are simulated; some by ASPEN/SP process simulation blocks and others by user Fortran blocks. Material balances, as well as utility consumptions, operating personnel requirements and ISBL costs for each plant are produced. A factor which combines the effect of OSBL plant costs, engineering and contingency costs is then applied to each ISBL plant so that the total is the total installed cost of the facility.

This model generates a file for direct transfer of the significant model results to the Lotus spreadsheet economics model. The spreadsheet model takes this input and uses the financial assumptions shown in Table 5 and product values generated by linear programming refinery simulations to calculate the cost of production and an equivalent crude oil price for a 15% return on investment. The overall calculation scheme is shown in Figure 1.

Sensitivity Studies

The PFS model is designed to handle the effects of the following process variables:

Primary Variables	
Coal Feed Rate	5,000 to 50,000 tpd
F-T Conversion per Pass	50 to 82%
F-T Wax Yield	10 to 50 wt %*
F-T Reactor Inlet Superficial Velocity	5 to 20 cm/s
F-T Reactor Catalyst Concentration	20 to 40 wt %
Secondary Variables (Approx. Range)	
H_2/CO Ratio	0.36 to 0.7
Heat Transfer Flux	3,000 - 10,000 $Btu/hr/ft^2$
Flow Regime	Bubble to Churn Turbulent

* Limited by heat transfer flux

Sensitivity studies will be carried out as soon as the model has been finalized and tested.

ADDITIONAL WORK

Alternative Cases

Two alternative cases will be examined:

- The first alternate case is an alternate upgrading case using close-coupled, fixed-bed processing of the vapor stream from the slurry F-T reactor over ZSM-5 catalyst. This case duplicates the two-stage process piloted by Mobil[2]. The ZSM-5 step radically alters the character of the F-T naphtha and light distillate fractions by converting 1-olefins and paraffins above C_7 and oxygenates into isoparaffins, isoolefins, naphthenes and aromatics. This results in a higher ratio of naphtha to distillate products. Wax hydrocracking, alkylation and light ends recovery are still required, but no further processing of the naphtha and distillate products is needed. The products are somewhat lower in quality than those produced in the Baseline Case but should be acceptable feed streams for refinery processing.

- The second Case uses Western coal from the Powder River Basin. The primary difference from the Baseline Case is that the western coal contains 31 wt% moisture, and it is not practical to dry the coal to less than 8% moisture. The Illinois #6 coal in the Baseline Case contains 8.6 wt% moisture and is dried to 2% moisture before gasification. Thus, considerably more fuel gas is required for drying in this case. The capacity of the Shell gasifier is reduced and nine gasifiers are required as compared to eight in the Baseline Case. The nine gasifiers handle 19,791 stpd of moisture free coal, as compared to 18,572 stpd of moisture free Illinois #6 coal from eight gasifiers in the Baseline Case. Synthesis gas to the F-T reactors is 1,214 MMSCFD of gas with an H2/CO ratio of 0.407, whereas in the Baseline Case it is 1,191 MMSCFD of gas with an H2/CO ratio of 0.361. Hydrocarbon production is slightly higher in the Western Coal Case. While there are meteorological differences due to site location, this should have only a minor effect on processing since the Baseline process design already maximizes air cooling. The Western site will affect construction costs and will limit maximum vessel size due to overland transportation limitations

The alternative cases will be documented in a later paper.

CONCLUSIONS AND RECOMMENDATIONS

The Baseline design has shown that indirect coal liquefaction is competitive with conventional crude oil production and refining at a crude cost of 30.7 $/bbl. This is considered an excellent result at this stage of development and it is quite competitive with direct liquefaction. Design conditions for the Fischer-Tropsch reactor are considered conservative and it is expected that the sensitivity studies will indicate that a significant improvement is possible.

The Baseline study represents only one of several possible processing options for Advanced Fischer-Tropsch Technology. While an intensive optimization study was carried out in designing the F-T slurry reactor loop, the same attention was not given to other key process selections which were beyond the scope of the current study. It is recognized, however, other developing gasification and upgrading options could positively impact the F-T technology economics. We recommend the Baseline study be extended to include a detailed investigation of the following:

- Texaco coal gasification technology
- FCC for F-T wax upgrading
- Economic analysis of producing raw instead of refined F-T products

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Table 1
Illinois No. 6 Seam Coal
Burning Star Mine
Washed Coal Analysis (a)

Item	As Rec'd	Dry Basis
Higher Heating value, Btu/lb (measured)	11,193	12,246
Proximate Analysis, wt %		
Moisture	8.60	-
Ash	10.50	11.49
Volatile Matter	38.60	42.23
Fixed Carbon	42.30	46.28
Ultimate Analysis, wt %		
Moisture	8.60	-
Ash	10.50	11.49
Carbon	64.90	71.01
Hydrogen	4.39	4.80
Nitrogen	1.28	1.40
Sulfur	2.92	3.19
Chlorine	0.09	0.10
Oxygen (by difference)	7.32	8.01
Ash Mineral Analysis, wt % ash		
Silica	49.70	
Alumina	19.10	
Ferric Oxide	17.50	
Sodium Oxide	0.50	
Potassium Oxide	1.90	
Calcium Oxide	6.20	
Magnesium Oxide	1.00	
Titanium Oxide	1.00	
Phosphorous Pentoxide	0.20	
Sulfur Trioxide	2.90	

(a) Source: Burning Star Mine

Table 2
Baseline Design
Major Feed and Product Streams

Feed:

ROM Coal*	1693.57 MLbs/hr (20323 Tons/day)	
Butane	26.50 MLbs/hr (3119 BPSD)	
Electric Power	50 MWh	
Primary Products:	**MLbs/hr**	**BPSD**
C3 LPG	14.22	1921
F-T Naphtha Blend	251.44	23915
F-T Distillate	278.21	24655
Sulfur	46.69	

* As received coal (8.6 wt% water)
Catalysts and chemicals are not included

Table 3
Baseline Design
F-T Product Valuation

	Phase I	Phase II
Clean Air Act Amendment		
% Reformulated Gasoline	15.0	70.0
% Low Sulfur Diesel	83.0	83.0
Crude Oil Price, $/Bbl	18.0	18.0
Petroleum Gasoline Price, $/Bbl	26.0	26.7
High Sulfur (0.25 wt%) Diesel Price, $/Bbl	22.7	22.7
Low Sulfur (0.05 wt%) Diesel Price, $/Bbl	24.8	24.8
F-T Naphtha Blend, $/Bbl	27.0	28.1
F-T Distillate, $/Bbl	24.9	25.2
F-T Naphtha/Crude Price Ratio	1.50	1.56
F-T Distillate/Crude Price Ratio	1.38	1.40

Table 4
Baseline Case - Design Basis and Work Scope

Plant No.	Process Plant	Source of Data	Deliverable
101	Coal Receiving & Handling	In-House data	A
102*	Coal Drying & Grinding	In-House data	A
103*	Coal Gasification	Shell Oil	B
104	COS/HCN Hydrolysis/Gas Cooling	In-House data	A
105	Sour Water Stripping	Open Art Unit	A
106	Acid Gas Removal	Proprietary	A
107*	Sulfur Recovery-Claus/TGT	In-House data	B
108	Sulfur Polishing	In-House data	B
109	Wet Scrubbing	In-House data	A
110	Air Separation	Proprietary	B
201**	Fischer-Tropsch Synthesis	MOBIL reports	A
202	CO2 Removal	Proprietary	B
203	Dehydration & Compression	Open Art Unit	A
204	Hydrocarbon Recovery	Open Art Unit	A
205	Hydrogen Recovery	In-House data	B
206	Autothermal Reformer	Proprietary	B
301	Wax Hydrocracker	UOP/PARC data	B
302	Distillate Hydrotreater	UOP data	B
303	Naphtha Hydrotreater	UOP data	B
304	Catalytic Reformer	UOP data	B
305	C4 Isomerization	Proprietary	B
306	C5/C6 Isomerization	Proprietary	B
307	H_2SO_4 Alkylation	Proprietary	B
308	Saturated Gas Plant	Open Art Unit	B

A - Material Balance with Utility Summary and Equipment List
B - Material Balance with Utility Summary
* With spare train
** With spare FT reactor only

Table 5
Capital Cost (in MM 1993 Dollars)

Area No.	Description	Direct Material	Direct/Indirect Field Labor	Direct Field Subcontract	Total
100	Syngas Production	811	477	20	1308
200	F-T Synthesis	271	167	15	453
300	Upgrading/Refining	89	62	1	152
	Offsite	151	160	173	484
	Total:	1,322	866	209	2,397
	Home Office Services/Fee and Contingency:				565
			Total Cost:		2,962

Table 6

Baseline Case - Economic Assumptions

Coal Cost, $/Ton	24.0
LPG Price, $/Bbl	12.3
Sulfur Price, $/Ton	80.0
n-Butane, $/Bbl	14.5
Plant Life, Years	25
Depreciation, Years	10
Construction Period, Years	4
Owner's Cost, % of Initial Capital	5
Initial Equity, %	25
Bank Interest Rate, %	8
General Inflation, %	3
Federal Income Tax, %	34
State and Local Taxes, %	0
On Stream Factor, %	90.8
% Plant Operation	
First Year	100
2nd. Year	100
3nd. Year	100
Labor Overhead Factor, % of salary	40
Maintenance and Insurance, % of Initial Capital	1

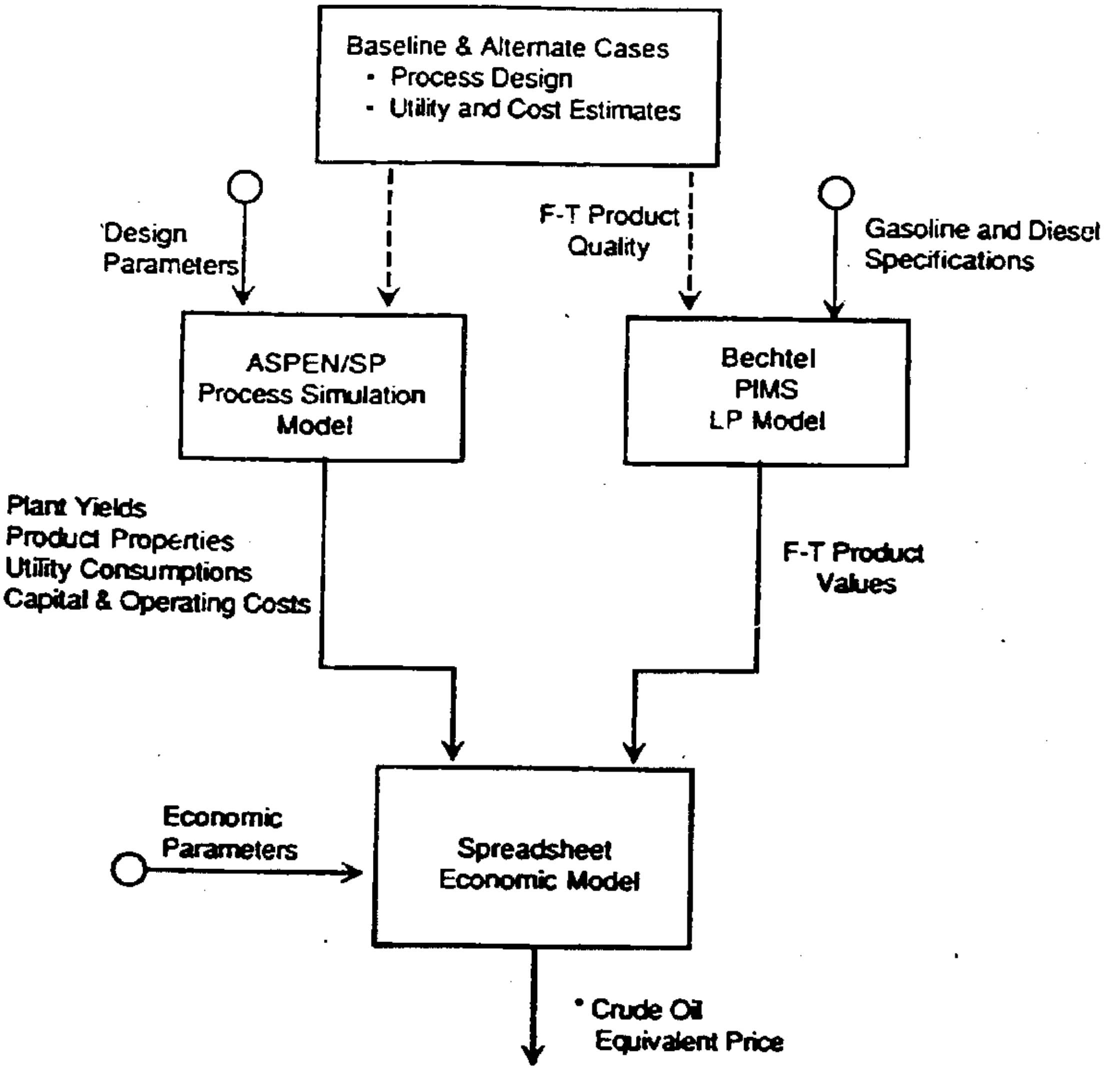
Table 7

Baseline Economic and Sensitivity Studies

	Baseline	Sensitivity Case		
	Value	Change	$/Bbl	Delta ($/Bbl)
Baseline Design			30.7	
Total Installed Capital, MM$	2961	+/- 10%	32.5/28.9	+/- 1.8
		+/- 25%	35.3/26.1	+/- 4.6
Coal Cost,$/Ton	24.00	+/- 10%	31.3/30.0	+/- 0.7
		+/- 25%	32.3/29.1	+/- 1.6
F-T Product/Crude Price Ratio				
Naphtha Blend	1.50	1.8	27.9	- 2.8
		1.2	34.2	+ 3.5
Distillate	1.38	1.7	28.0	- 2.7
		1.1	34.1	+ 3.4
Owner's Equity, %	25.0	+ 100%	33.6	+ 2.9
First Three Years Operation, %				
(On-Stream Factor = 90.8%)	100/100/100			
25/50/100			33.8	+ 3.1
25/75/100			33.1	+ 2.4
Maintenance & Insurance, %	1.0	2.0	32.0	+ 1.3

Figure 1

F-T INDIRECT COAL LIQUEFACTION STUDY
PROCESS DESIGN/SIMULATION MODELS DEVELOPMENT



* Crude Oil Equivalent Price is defined as the crude oil price needed to achieve a 15% internal rate of return

Figure 2
INDIRECT COAL LIQUEFACTION BASELINE STUDY
OVERALL PROCESS CONFIGURATION

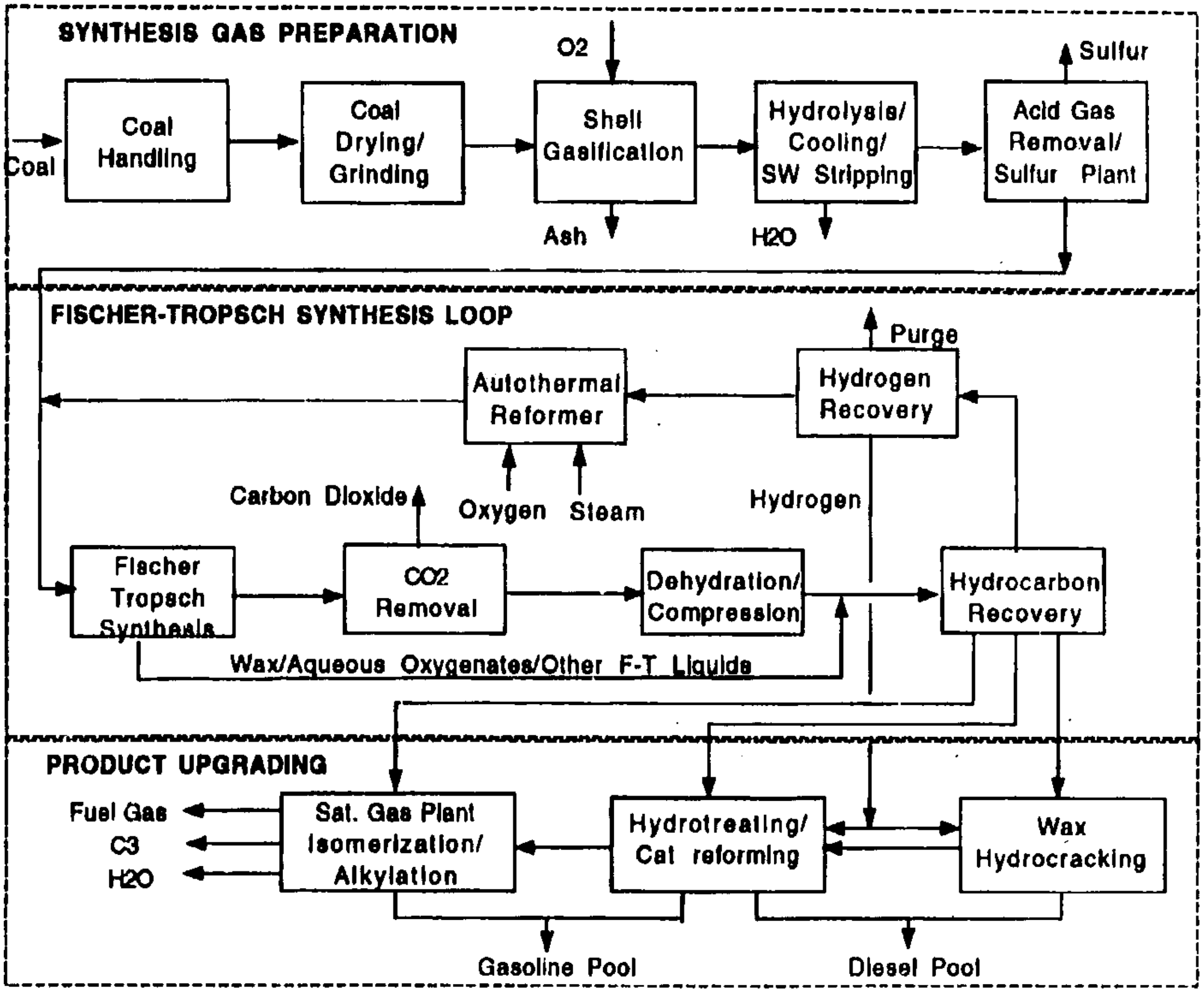


Figure 3
SLURRY FISCHER-TROPSCH REACTOR