SIMULATION MODELS AND DESIGNS FOR ADVANCED FISCHER-TROPSCH TECHNOLOGY

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ABSTRACT

Process designs and economics were developed for three grass-roots indirect Fischer-Tropsch coal liquefaction facilities. A baseline and an alternate upgrading design were developed for a mine-mouth plant located in southern Illinois using Illinois No. 6 coal, and one for a mine-mouth plant located in Wyoming using Powder River Basin coal. The alternate design used close-coupled ZSM-5 reactors to upgrade the vapor stream leaving the Fischer-Tropsch reactor. ASPEN process simulation models were developed for all three designs. These results have been reported previously.

In this study, the ASPEN process simulation model was enhanced to improve the vapor/liquid equilibrium calculations for the products leaving the slurry bed Fischer-Tropsch reactors. This significantly improved the predictions for the alternate ZSM-5 upgrading design. Another model was developed for the Wyoming coal case using ZSM-5 upgrading of the Fischer-Tropsch reactor vapors. To date, this is the best indirect coal liquefaction case. Sensitivity studies showed that additional cost reductions are possible.

INTRODUCTION

This study is conducted under an extension of DOE Contract No. DE-AC22-91PC90027, entitled "Baseline Design/ Economics for Advanced Fischer-Tropsch Technology. This program previously has:

- Developed a baseline design and two alternate designs for indirect coal liquefaction using advanced Fischer-Tropsch (F-T) technology. The baseline case used Illinois No. 6 Coal and conventional petroleum refinery technology for product upgrading. ZSM-5 treatment of the F-T reactor vapor stream was studied as an alternate product upgrading scheme. Another alternate case used Wyoming Powder River Basin coal. All schemes contain a wax hydrocracker.

- Determined the capital cost, operating costs and utility requirements for the above three cases.

- Developed ASPEN Process Flowsheet Simulation (PFS) Models for the above cases and a discounted-cash flow (DCF) economics spreadsheet model. Product valuation is based on a linear programming (LP) analysis of the refinery which will process the liquefaction products into finished products. In concert, these closely-coupled models constitute a research tool which DOE can use to plan, guide, and evaluate its ongoing and future research and commercialization programs for the production of transportation fuels by indirect coal liquefaction.

- Performed preliminary sensitivity studies to examine the effects of key independent process variables and economic assumptions on the baseline case using the ASPEN PFS model.

Details concerning the overall design basis, process selection, and costs were reported at a previous DOE/PETC contractor's conference[1]. The PFS model development, the F-T product valuation study, and the simulation model results were presented in three separate papers last year[2-4].

This paper describes recent enhancements to the ASPEN PFS model which improved the predictions for the alternate ZSM-5 product upgrading case. This enhanced model then was used to study the effects of using ZSM-5 upgrading for Wyoming Powder River Basin coal.

PROCESS DESIGN

Overall Plant Configuration

Figure 1 is a block flow diagram showing the overall F-T process configuration for the baseline design case. The OSBL area is not shown. The ISBL plant contains three areas.

1. Area 100 - Syngas Production -- In the baseline design case, synthesis gas is generated in Shell gasifiers from ground, dried coal. The raw syngas is washed in a wet scrubber followed by single-stage COS/HCN hydrolysis and cooling, acid gas removal, and sulfur polishing. Area 100 also includes the sour water stripper and sulfur recovery plants. Because of the lower sulfur content of the Wyoming coal, Rectisol, instead of an inhibited amine, is used in the acid gas removal step, and the COS/HCN hydrolysis step is not required.
2. Area 200 - The Fischer-Tropsch Synthesis Loop -- The F-T synthesis loop includes F-T synthesis, CO2 removal, recycle gas compression and dehydration, hydrocarbon recovery by deep refrigeration, hydrogen recovery and autothermal reforming processing steps. The hydrocarbon recovery unit includes oxygenates wash columns which are not required when the F-T reactor vapor product is upgraded in the ZSM-5 reactors.
3. Area 300 - Product Upgrading -- In the baseline case, this area contains eight processing steps; 1) wax hydrocracking, 2) distillate hydrotreating, 3) naphtha hydrotreating, 4) naphtha reforming, 5) C4 isomerization, 6) C5/C6 isomerization, 7) C3/C4/C5 alkylation, and 8) saturated gas processing and product blending. The hydrocracked and hydrotreated naphthas are catalytically reformed to produce an aromatic gasoline blending component. The lighter materials are isomerized and alkylated to produce a high quality gasoline blending stock. Purchased butanes are required to alkylate all the available olefins.

   In the ZSM-5 upgrading case, this area only contains four processing steps; 1) wax hydrocracking, 2) C4 isomerization, 3) C3/C4/C5 alkylation, and 4) saturated gas processing and product blending. In addition, this case produces sufficient butanes so that extra butanes are available for sale.

These F-T liquefaction facilities produce C3 LPG, a C5 - 350 F gasoline blending component, and a 350 - 850 F combined light and heavy distillate. As mentioned above, the ZSM-5 upgrading case also produces some butanes for sale. Liquid sulfur is the primary byproduct. The hydrocarbon products contain no significant sulfur or nitrogen because of the nature of the Fischer-Tropsch reaction. Oxygen is reduced to less than 30 ppmv in the hydroprocessing steps. Olefins are saturated to low residual levels in both the gasoline and distillate products. There are virtually no aromatics in the distillate. The diesel portion of the distillate has a high cetane number (about 70), and the jet fuel and heavy distillate fractions have low smoke points. However, they do have relatively high pour points which must be carefully controlled. In the baseline case, the gasoline product is a mixture of C3/C4/C5 alkylate, C5/C6 isomerate and catalytic reformate. It is basically a raw gasoline with almost an 89 clear (R+M)/2 octane number.

Reactor Design

The F-T slurry reactor essentially is a bubble column reactor in which the slurry phase is a mixture of molten wax and catalyst. The synthesis gas provides the agitation necessary for good mixing and mass transfer of the reactants and products between the two phases. The slurry reactor design was chosen over a fixed bed reactor based on an earlier DOE sponsored Bechtel study[5,6]. The design is based on the slurry F-T reactor data from Mobil's two-stage pilot plant studies for the DOE[7]. These results are the basis for development of the yield correlations contained in the F-T slurry bed computer model used in this study[8].

In this reactor model, vapor and liquid product streams are continuously removed. In the baseline case, this vapor stream is cooled for hydrocarbon recovery before entering the carbon dioxide removal step. The heavy liquid stream goes to the hydrocarbon recovery area wherein all the hydrocarbon products are fractionated and sent to the appropriate product upgrading plant. All the C6 and heavier hydrocarbons in either the vapor or liquid streams are recovered and upgraded to useable products irrespective of the stream in which they leave the slurry bed F-T reactor. Thus, the accuracy of the vapor/liquid flash calculation is not a critical issue.

MODEL ENHANCEMENTS

In the alternate ZSM-5 upgrading case, the entire vapor stream leaving the slurry bed Fischer-Tropsch reactor passes through the ZSM-5 oligomerization reactor before cooling and hydrocarbon recovery. Consequently, in this case, it does make a difference in which stream the heavier hydrocarbon products leave the slurry bed F-T reactor. In the original ASPEN PFS model, individual normal paraffins and 1-olefins through C19 and a single lumped C20+ wax pseudocomponent were used to represent the F-T reaction products. From Schultz-Flory theory, the average carbon number of the C20+ wax can be calculated[8]. The normal boiling point and gravity of the C20+ wax pseudocomponent were determined based on its calculated average carbon number by extrapolating the properties of normal paraffins and 1-olefins assuming that 70% of the wax has a single olefinic bond. These properties (1032 F ABP, 38.6 API gravity, and a 618 MW) were then used as input to the ASPEN pseudocomponent property estimation procedures.

An examination of the model predictions of the amount of material leaving the slurry bed reactor in the vapor and liquid streams for the ZSM-5 case was made to see if the model was making a reasonable prediction of the amount of hydrocarbons leaving in the vapor stream. It predicted that 4804 lbs/hr or about 77% of the C18 material was leaving in the vapor stream, and 3716 lbs/hr or about 68% of the C19 material was leaving in the vapor stream. By simple extrapolation, one would expect about 2500 lbs/hr of the C20 material to leave in vapor stream.

However, the model predicted that only 1128 lbs/hr of the C20 and heavier material was leaving in the vapor stream. Thus, it was easy to conclude that this model is underpredicting the amount of C20 and heavier material leaving the F-T reactor in the vapor phase and the lumping of all the C20 and heavier material into a single C20+ wax pseudocomponent is the cause.

Additional components through C29 were added to improve the vapor/liquid equilibrium predictions at the F-T slurry bed reactor conditions of 488 F and 325 psia. Individual components were added for normal eicosane and 1-eicosene since they both were available in the standard ASPEN PLUS data bank. Although the ASPEN PLUS data bank contains data for the normal praffins through C30, it does not contain pure component physical property data for any olefinic compounds above C20. The API *Technical Data Book - Petroleum Refining* and other standard data compilations do not contain such data either. Thus, in an effort to minimize the number of components in the simulation, a single pseudocomponent was used to represent the olefin/praffin mixture at each carbon number from C21 up to and including C29.

At each carbon number for a mixture of 70 mole% 1-olefins and 30 mole% normal paraffins, the normal boiling point and API gravity of the C21 through C29 olefin/paraffin pseudocomponents were estimated by extrapolating the property differences between the 1-olefin and the corresponding normal paraffin. These properties and the calculated molecular weight were then used as input to the ASPEN pseudocomponent property estimation procedures. In this new situation, the wax pseudocomponent is the C30 and heavier material. Using the same methodology as before, the properties of the C30+ wax were estimated (1128 F ABP, 36.4 API gravity, and a 743 MW) and used as input to the ASPEN pseudocomponent property estimation procedures. Thus, eleven additional components were added to the ASPEN simulation model, and the wax pseudocomponent was redefined to be the C30 and heavier material (although the actual wax production still is considered to be the C20 and heavier material).

Examination of the results from this simulation showed that 715 lbs/hr or about 17% of the C28 material left the F-T reactor in the vapor stream, and 571 lbs/hr or about 14% of the C29 material left in the vapor stream. Furthermore, this model predicted that only 228 lbs/hr of the C30+ wax was leaving in the vapor stream. From a simple extrapolation of the C28 and C29 vapor flows, one would expect about 400 lbs/hr of C30 in the vapor alone. Thus, the model still underpredicts the amount of wax leaving in the vapor stream, but now to a much lesser extent.

Therefore, another approach was used to correct this underprediction without adding additional components. Figure 2 shows an extrapolation of the amount of each carbon number component in the vapor stream as a function of carbon number. This figure shows that about another 30 components are needed to reduce the weight in the vapor of a given carbon number component to below 1 lb/hr. As shown in this figure, the weight in the vapor decreases exponentially with increasing carbon number. Thus, using the average boiling point of the wax in the vapor/liquid equilibrium calculations always will underpredict the amount of wax in the vapor.

From Figure 2 the expected weight in the vapor of the C30 through C50 components is 2762 lbs/hr, and that of the C30 through C60 components is 2803 lbs/hr. Therefore, it was decided to use an effective boiling point for the C30+ wax pseudocomponent rather than the calculated average boiling point to better reproduce its vapor/liquid behavior at the F-T reactor conditions since this behavior is controlled by the lightest portion of the C30+ wax. This effective boiling point being defined as that which produces an amount of vapor equal to that predicted by Figure 2 for the C30+ material. By trial an error, it was found that reducing the average boiling point of the C30+ wax by 154 Fahrenheit degrees (to produce an effective boiling point of 974 F) results in 2834 lbs/hr of the C30+ wax component leaving the F-T reactors in the vapor stream.

Thus, an enhanced PFS model was developed which contains eleven new components, and a revised C30+ wax component which uses an effective boiling point that is 154 Fahrenheit degrees below that which would be calculated based on the average boiling point of the C30+ wax. This enhancement was incorporated in all the ASPEN PFS indirect coal liquefaction models.

COMPARISON OF MODELS

Table I compares the results from the original ASPEN simulation model with those from the enhanced model for the baseline case with Illinois No. 6 coal. As expected, there is little difference between the two models. Any differences are well within the accuracy of the experimental data upon which the models are based. Also contributing to these slight differences is a switch from ASPEN/SP software in which the original models were developed to ASPEN PLUS software which is used for the enhanced models.

The preliminary Crude Oil Equivalent (COE) prices shown at the bottom of the table are based on the 1995 EIA forecast with a 75% owners equity. Table II shows the basic economic parameters used for these COE calculations in which margins are used to relate the product values to the calculated COE. The COE values are included in Table I only to show that the differences between the two models truly are insignificant in terms of the COE.

Table III compares the results from the original ASPEN simulation model with those from the enhanced model for the ZSM-5 upgrading case using Illinois No. 6 coal. As a result of the revision, there are significant differences between the results from the two models. In the original model, 1,128 lbs/hr of C20+ wax left the F-T reactor in the vapor stream and was processed in the ZSM-5 reactors. In the enhanced model, the flow rate of the C20+ wax material increased to 18,614 lbs/hr. As a result, the total amount of C7+ material processed in the ZSM-5 reactors increased by about 12% to 163,214 lbs/hr, and the hydrocarbon feed to the wax hydrocracker was reduced by 16,489 lbs/hr. These, changes increased the performance difference between the ZSM-5 case and the baseline design case since the effect of ZSM-5 reactor is increased.

Compared to the baseline design case, the ZSM-5 upgrading case radically alters the F-T product distribution and makes more gasoline, butanes, and LPG and less distillate. The gasoline blending stock is of a higher quality[3]. The differences between these two designs previously has been discussed in detail[4], and will not be repeated here. However, with the enhanced model, the differences are somewhat greater and result in a 0.1 $/bbl reduction in the COE.

ZSM-5 UPGRADING OF THE ALTERNATE WYOMING COAL

The primary design differences between the base Wyoming Powder River Basin coal case and the baseline design case are a consequence of the coal properties, although location also is important. Wyoming coal contains less sulfur and has a higher moisture content. It has to be dried from the as-received moisture content of 31 to 8 wt% moisture before gasification as compared to drying the Illinois No. 6 coal from 8.6 to 2 wt% moisture. The lower sulfur content necessitates a Rectisol process rather than an amine guard treating system for sulfur removal. Water availability and cost and the availability of skilled labor also were considered in developing the Wyoming base case. Consequently, there are significant design changes in the syngas production area and the offsite water treatment plant for the Wyoming location. Reference 4 provides a detailed comparison between the base Wyoming coal case and the baseline Illinois No. 6 coal case.

Because reference 4 showed an economic incentive for both the ZSM-5 upgrading case (Illinois No. 6 coal) and the Wyoming coal case over the baseline case, a ZSM-5 upgrading Wyoming coal case was developed using the enhanced model. In this case a ZSM-5 reactor processes the vapor products from the F-T reactor and converts the olefins, C7+ paraffins and oxygenates to isoolefins, isoparaffins, naphthenes, and aromatics. This eliminates the catalytic reforming, C5/C6 isomerization, naphtha hydrotreating, and distillate hydrotreating plants in the Area 300 Product Upgrading Section. C4 isomerization and alkylation are still required, but the yield of alkylate is increased, and no butanes have to be purchased. The F-T wax stream is processed in the same manner as the base case by mild hydrocracking.

Table IV compares the two Wyoming coal cases. The differences are similar to those for the Illinois coal cases except that the Wyoming ZSM-5 coal case purchases more power than the base case. The ZSM-5 case has a higher gasoline to distillate ratio. Since the F-T reactor is run at a 50% wax yield in both cases, the product yield differences are due entirely to the conversion of the F-T distillate to naphtha in the ZSM-5 reactor. These upgrading reactions produce more light ends which, in turn, increases the C3 LPG, butanes and alkylate yields. The COE for the Wyoming coal ZSM-5 upgrading case is 2.8 $/bbl below that of the base case. This reduction is about 0.25 $/bbl less than that for the corresponding Illinois No. 6 coal cases.

SENSITIVITY STUDIES

Previously, the ASPEN PFS model was used to study the sensitivity of the baseline Illinois No. 6 coal design to coal feed rate, F-T syngas conversion per pass, wax yield, and several F-T slurry bed reactor design variables[4]. This paper reports the effects of coal feed rate and wax yield for the Wyoming coal ZSM-5 upgrading design. The base conditions are 19,789 tons/day of dry Powder River Basin coal and 70.5% hydrogen conversion/pass (corresponding to a 83.5% syngas conversion/pass) at a 50 wt% wax yield.

Effect of Design Plant Capacity

The effect of plant capacity on the overall capital investment of the facility is exponential with an average cost-capacity exponent of 0.92 over the range from 10,000 to 60,000 tons/day of dry ROM coal. This exponent is high because the plant contains many parallel processing trains. At the base case coal feed rate, most of the Area 100 and 200 plants are at their maximum size and have multiple trains. For each processing plant, the PFS model calculates the required number of duplicate trains from the total plant throughput.

Table V shows the overall liquefaction plant inputs and outputs as a function of design capacity. As expected, the COE decreases as the capacity increases. Figure 3 shows that doubling the plant capacity will reduce the COE by about 1.7 $/bbl. However, this plant will have a dry ROM coal feed rate of almost 40,000 tons/day and cost almost 5,900 million dollars. It will produce just over 100,000 bbls/day of C3 and heavier hydrocarbons.

Effect of Design Wax Yield

The F-T reactor simulation model is designed to simulate operations at wax yields between 10 and 75 wt% wax. The effect of wax yield between 30 and 75 wt% wax was studied by varying the F-T reactor temperature between 503 and 468 F. Table VI summarizes the results. Below a wax yield of about 30 wt%, insufficient butanes are available to alkylate the olefins, and butanes have to be purchased. The Wyoming coal ZSM-5 upgrading model is not designed to handle this situation. Above about a wax yield of 68 wt%, the model predicts that excess isobutane is going to the alkylation unit, and some reprogramming is necessary to correctly handle this situation. As currently structured, the model slightly overpredicts the plant cost for these cases.

As the wax yield increases, less gasoline and more distillate are produced. At a 30 wt% wax yield, the gasoline to distillate ratio is 4.7, and at a 75 wt% wax, it decreases to 0.96. Also, as the wax yield increases, the character and the quality of the gasoline changes dramatically. At a 30 wt% wax yield, the gasoline contains just over 28% alkylate, 19 wt% aromatics, 8 wt% olefins, and has an (R+M)/2 octane number of about 76. At a 75 wt% wax yield, the gasoline contains about 14% alkylate, 11 wt% aromatics, 4 wt% olefins, and has an (R+M)/2 octane number of about 71. These gasoline quality changes are the result of two factors. At low wax yields, more hydrocarbon products leave the F-T reactor in the vapor phase and are upgraded to isoparaffins. isoolefins, naphthenes and aromatics. Some cracking also occurs producing more butanes and light olefins in the ZSM-5 reactor. Thus, more alkylate is produced. As the wax yield increases, less material is upgraded in the ZSM-5 reactors, and more wax is processed in the wax hydrocracker. The wax hydrocracker produces both distillate and a lower octane gasoline blending component which has very low aromatic and olefin contents. As a result of these competing forces, more C4 and heavier hydrocarbons are produced as the wax yield increases. However, the propane yield decreases at a greater rate so that there is a slight loss in the yield of C3 and heavier hydrocarbons as the wax yield increases.

Figure 4 shows the effect of wax yield on the COE. These COEs were calculated using a constant delta between the gasoline and crude oil price[3]. They do not account for the changing quality of the gasoline as a function of wax yield. Reducing the wax yield from the design case of 50 wt% to 30 wt% reduces the COE by about 0.5 $/bbl.

CONCLUSIONS AND RECOMMENDATIONS

All the ASPEN PLUS PFS models for indirect Fischer-Tropsch coal liquefaction have been enhanced to better characterize the vapor/liquid equilibrium predictions of the F-T product at reactor conditions. This enhancement did not change the model predictions for the baseline design case which uses conventional technology to upgrade the F-T wax. However, it changed the predictions for the cases where the F-T vapors are upgraded using ZSM-5 catalyst since more material now leaves the F-T reactors in the vapor phase. Parametric studies have shown that conditions other than those of the base case may be more economically attractive. Additional product upgrading schemes are being studied. One uses the ZSM-5 reactor to upgrade the low quality hydrocracked naphtha. Another involves substituting a fluid catalytic cracking unit for the wax hydrocracker and using the additional C4 through C7 olefins to make ethers for use as gasoline blending components.

Additional LP studies should be done to further define the product values; especially for a situation like that of the wax yield study where the gasoline quality changes dramatically as the wax yield changes.

ACKNOWLEDGMENT

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Figure 1
INDIRECT COAL LIQUEFACTION BASELINE STUDY
OVERALL PROCESS CONFIGURATION

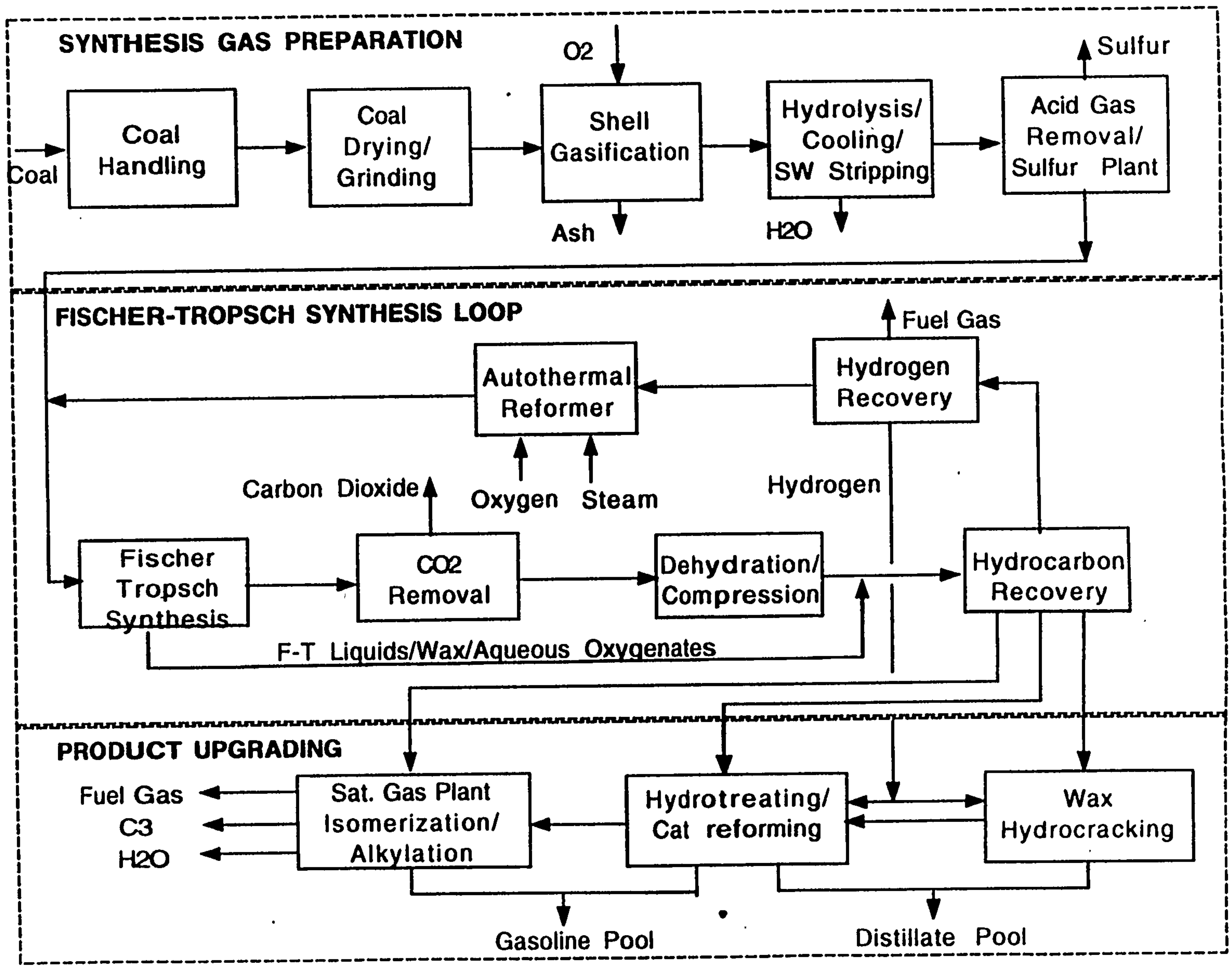

Figure 2
Weight in Vapor vs. Carbon Number

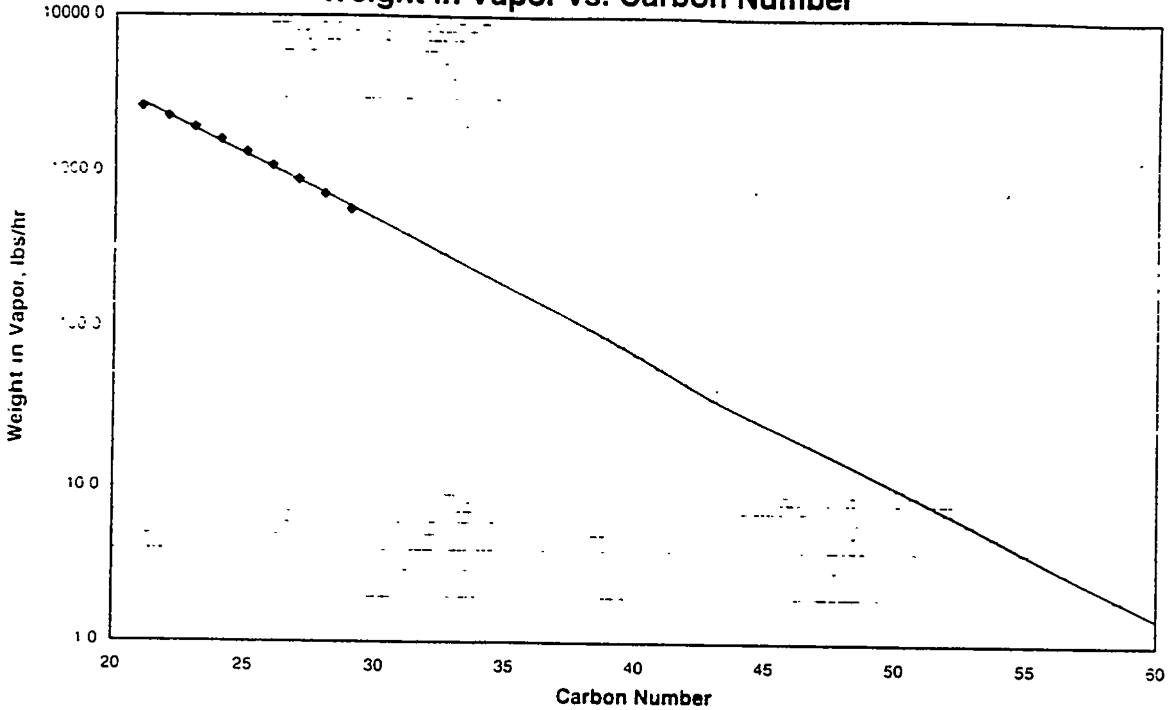

Figure 3
Effect of Design Plant Capacity on COE

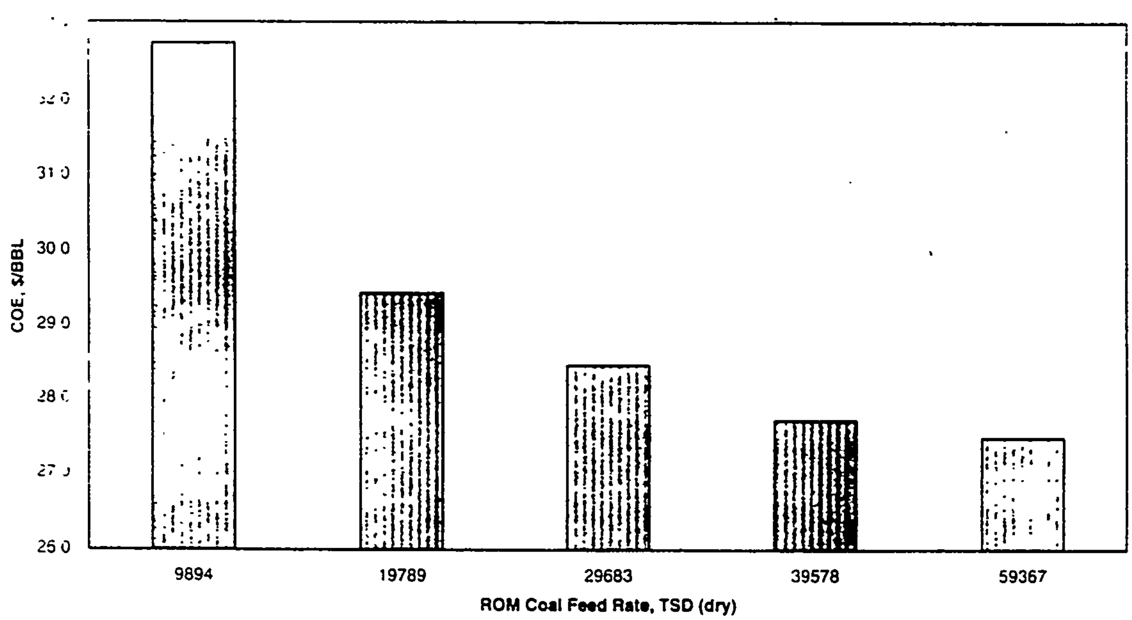

Figure 4

Effect of Wax Yield on COE

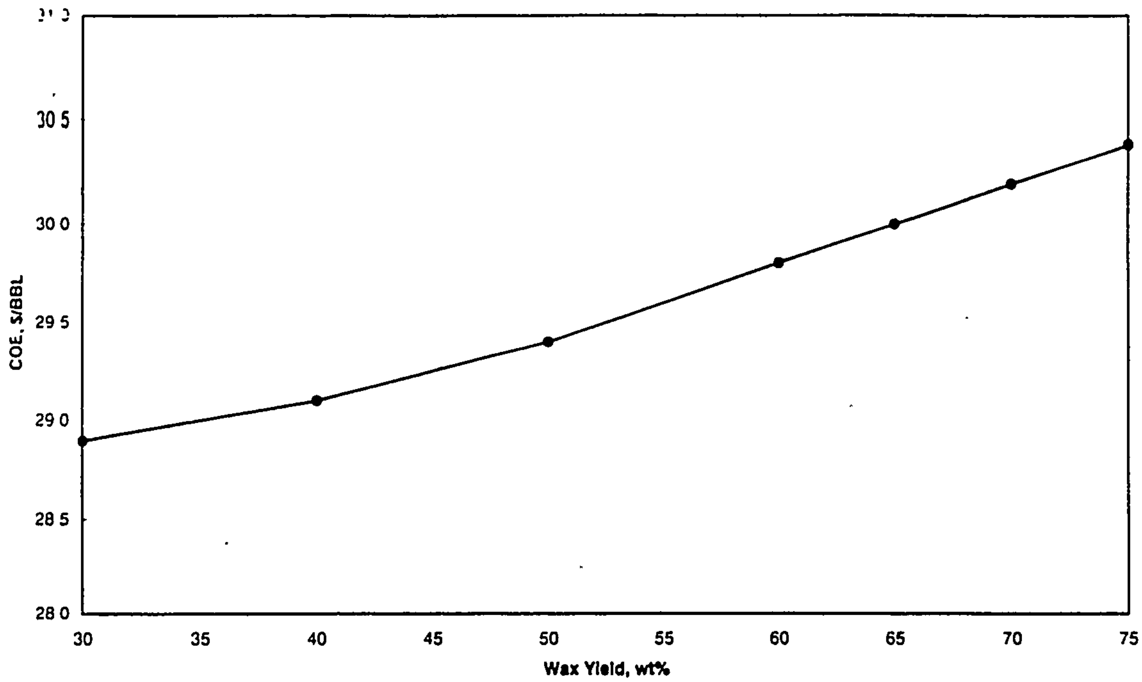

TABLE I

Comparison of the ASPEN Models for the Baseline Design Case

	Original Model	Enhanced Model	Difference
ROM Coal Feed Rate, TSD (MF)	18575	18575	0
Slag Production Rate, TSD (MF)	2244	2244	0
Electricity Purchase, MEGA-WH/SD	1304.5	1303.8	-0.7
Raw Water Make-up, MMGSD	14.460	14.460	0
Gasoline Production, BSD	23952	23943	-9
Distillate Production, BSD	24681	24686	5
Liquid Propane Production, BSD	1923	1922	-1
Mixed Butanes Production, BSD	-3121	-3110	11
Sulfur Production, TSD	560	560	0
No. of Operators and Maintenance Workers	1088	1088	0
Total Installed Capital, $MM	2963.5	2963.8	0.3
Catalyst and Chemicals, $MM	31.1	31.1	0.0
Crude Oil Equivalent Price, $/BBL	35.8	35.8	0.0

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TABLE II

Economic Parameters

Coal Cost, $/ton (31 wt% moisture)	4.5
LPG Price, $/bbl	12.3
N-Butane Price, $/bbl	14.5
Sulfur Price, $/ton	80
Electricity, cents/KWH	5
Plant Life, years	25
Depreciation, years	10
Construction Period, years	4
Owner's Cost, % of initial capital	5
Owner's Initial Equity, %	75
Bank Interest Rate, %/year	8
General Inflation, %/year	3.1
Escalation Above General Inflation, %/year	
Coal	0.8
Crude Oil	2.4
Electricity	0.3
Federal Income Tax, %	34
State and Local Taxes, %	0
Maintenance and Insurance, % of initial capital	1
Labor Overhead Factor, % of salary	40
On-Stream Factor, %	90.8
Percent Plant Operational in	
First Year	100
Second Year	100
Third Year	100

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TABLE III

Comparison of the ASPEN Models for the ZSM-5 Upgrading Case

	Original Model	Enhanced Model	Difference
ROM Coal Feed Rate, TSD (MF)	18575	18575	0
Slag Production Rate, TSD (MF)	2244	2244	0
Electricity Purchase, MEGA-WH/SD	1194.8	1262.0	67.2
Raw Water Make-up, MMGSD	14.460	14.436	-0.024
Gasoline Production, BSD	30317	31255	938
Distillate Production, BSD	16820	15858	-962
Liquid Propane Production, BSD	2509	2623	114
Mixed Butanes Production, BSD	886	998	112
Sulfur Production, TSD	560	560	0
No. of Operators and Maintenance Workers	1024	1024	0
Total Installed Capital, $MM	2901.1	2897.2	-3.9
Catalyst and Chemicals, $MM	30.7	31.4	0.7
Liquid Feed to the Wax Hydrocracker, Mlbs/hr	290.0	273.5	-16.5
Crude Oil Equivalent Price, $/BBL	32.8	32.7	-0.1

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TABLE IV

Comparison of the ZSM-5 Upgrading Model for the Wyoming Coal Case with the Base Wyoming Coal Design Case

	Base Western Coal Design Case	Western Coal ZSM-5 Upgrading Case	Difference
ROM Coal Feed Rate, TSD (MF)	19789	19789	0
Slag Production Rate, TSD (MF)	1747	1747	0
Electricity Purchase, MEGA-WH/SD	2111.7	2198.4	86.8
Raw Water Make-up, MMGSD	9.803	9.573	-0.230
Gasoline Production, BSD	23756	31026	7270
Distillate Production, BSD	24466	15772	-8694
Liquid Propane Production, BSD	1907	2613	706
Mixed Butanes Production, BSD	-3101	980	4081
Sulfur Production, TSD	108	108	0
No. of Operators and Maintenance Workers	1190	1126	-64
Total Installed Capital, $MM	3149.0	3075.0	-74.0
Catalyst and Chemicals, $MM	21.3	21.5	0.2
Total C3+ Hydrocarbons, BSD	47028	50391	3363
Total C4+ Hydrocarbons, BSD	45121	47778	2657
Crude Oil Equivalent Price, $/BBL	32.2	29.4	-2.8

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TABLE V

Sensitivity Study - Effect of Design Plant Capacity
Wyoming Coal ZSM-5 Upgrading Case

ROM Coal Feed Rate, TSD (MF)	9894	19789	29683	39578	59367
Slag Production Rate, TSD (MF)	874	1747	2621	3494	5241
Electricity Purchase, MEGA-WH/SD	1099.2	2198.4	3297.6	4396.8	6595.3
Raw Water Make-up, MMGSD	4.787	9.573	14.36	19.146	28.719
Gasoline Production, BSD	15513	31026	46539	62051	93077
Distillate Production, BSD	7886	15772	23658	31544	47317
Liquid Propane Production, BSD	1307	2613	3920	5227	7840
Mixed Butanes Production, BSD	490	980	1470	1960	2940
Sulfur Production, TSD	54	108	163	217	325
No. of Operators and Maintenance Workers	685	1126	1636	2074	3087
Total Installed Capital, $MM	1667.8	3075.0	4494.0	5870.0	8758.6
Catalyst and Chemicals, $MM	10.8	21.5	32.3	43.0	64.6
Total C3+ Hydrocarbons, BSD	25196	50391	75587	100782	151174
Total C4+ Hydrocarbons, BSD	23889	47778	71667	95555	143334
Crude Oil Equivalent Price, $/BBL	32.7	29.4	28.5	27.7	27.5

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Table VI

Sensitivity Study - Effect of Wax Yield
Wyoming Coal ZSM-5 Upgrading Case

Wax Yield, Wt%	30	40	50	60	65	70	75
ROM Coal Feed Rate, TSD (MF)	19789	19789	19789	19789	19789	19789	19789
Slag Production Rate, TSD (MF)	1747	1747	1747	1747	1747	1747	1747
Electricity Purchase, MEGA-WH/SD	2009.4	2115.1	2198.4	2282.8	2327.7	2379.8	2441.2
Raw Water Make-up, MMGSD	9.758	9.658	9.573	9.493	9.454	9.413	9.371
Gasoline Production, BSD	38696	34743	31026	27580	25942	24356	22814
Distillate Production, BSD	8162	12146	15772	19158	20782	22366	23917
Liquid Propane Production, BSD	3432	3001	2613	2258	2090	1930	1777
Mixed Butanes Production, BSD	259	556	980	1352	1520	1671	1811
Sulfur Production, TSD	108	108	108	108	108	108	108
No. of Operators and Maintenance Workers	1146	1129	1126	1129	1129	1129	1129
Total Installed Capital, $MM	3074.9	3076	3075	3089.6	3099.3	3108.7	3120.8
Catalyst and Chemicals, $MM	21.6	21.6	21.5	21.5	21.5	21.5	21.5
Total C3+ Hydrocarbons, BSD	50549	50448	50391	50348	50334	50323	50319
Total C4+ Hydrocarbons, BSD	47117	47447	47778	48090	48244	48393	48542
Gasoline Properties:							
(R+M)/2 Octane Number	76	75	74	74	73	72	71
Aromatics, wt%	19.2	17.8	16.2	14.4	13.3	12.1	10.8
Olefins, wt%	8.0	7.1	6.5	5.8	5.3	4.8	4.2
Alkylate Production, BSD	11025	9054	7161	5436	4627	3851	3104
Percent Alkylate in Gasoline	28.5	26.1	23.1	19.7	17.8	15.8	13.6
Crude Oil Equivalent Price, $/BBL	28.9	29.1	29.4	29.8	30.0	30.2	30.4

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