2.5 Case 5 -- Illinois No. 6 Coal with FCC Wax Upgrading with Beta Zeolite Catalyst Case

This and the following alternative case consider Fluid Catalytic Cracking (FCC) as the means of upgrading the F-T wax instead of hydrocracking it which was done in the four previous cases. In this case a beta zeolite FCC catalyst is used, and in the following case, an equilibrium USY catalyst is used. The FCC yields for these two cases were developed from pilot plant data obtained by Amoco Oil.[1] Although the beta zeolite FCC cracking catalyst used in this design is not widely used commercially, it was selected because it produces large amounts of C4, C5 and C6 olefins which can be converted to ethers, and previous economics have shown large incentives for ether production. The equilibrium USY catalyst used in the following case is a commercial FCC catalyst in wide use. Both of these FCC upgrading cases use Illinois No. 6 coal with the plant located in southern Illinois. This case is documented in Quarterly Reports of April-September 1995 and October-December 1995, as well as in a paper presented at the ACS National Meeting, Orlando, FL on August 25-29, 1996.

As in the Baseline design case, the ISBL processing area is divided into three main processing areas and an offsites area. Simplified ISBL block flow diagrams for this alternate Eastern coal with FCC wax upgrading case are shown in Figures 2.14, 2.15 and 2.16 respectively for the three main processing area of 100, 200 and 300. Since this case also processes the same Illinois No. 6 coal as the Baseline case, Area 100 is identical to that of the Baseline case. Also, Area 200, the Fischer-Tropsch Synthesis Loop, is identical to that of the Baseline design case.

Area 300, the Product Upgrading and Refining Area, for the FCC upgrading cases is different than those for the previous cases. It contains ten plants; Plants 302 through 311. In this case, Plant 301, the Wax Hydrocracking Plant, has been replaced by three new plants, Plant 309, 310 and 311. Plant 309, the FCC Plant, is a direct replacement for the wax hydrocracker. Plants 310 and 311 were added to convert the olefins produced in the FCC plant to ethers. Plant 310 converts isobutylene to MTBE (methyl tertiary-butyl ether). Plant 311, the NExTAME Ethers Plant, converts the C5, C6 and C7 reactive olefins to amyl, hexyl and heptyl ethers. These ethers are high octane oxygenated gasoline blending components.

In this case, Plants 302 through 308 are identical to those in the Baseline design case as described in Section 3.1.1. However, some have different capacities.

Plant 309, the Fluid Catalytic Cracking Plant, catalytically cracks the F-T wax over a beta zeolite catalyst to produce distillate, cat cracked naphtha, C4s, C3 olefins and fuel

[1] M. M. Schwartz et al (Amoco Oil), The Selective Catalytic Cracking of Fischer-Tropsch Liquids to High Value Transportation Fuels, DOE Contract No. DE-AC22-91PC90057, *Final Report*, August, 1994.

Figure 2.14

Block Flow Diagram - Eastern Coal\FCC Area 100 (Syngas Production)

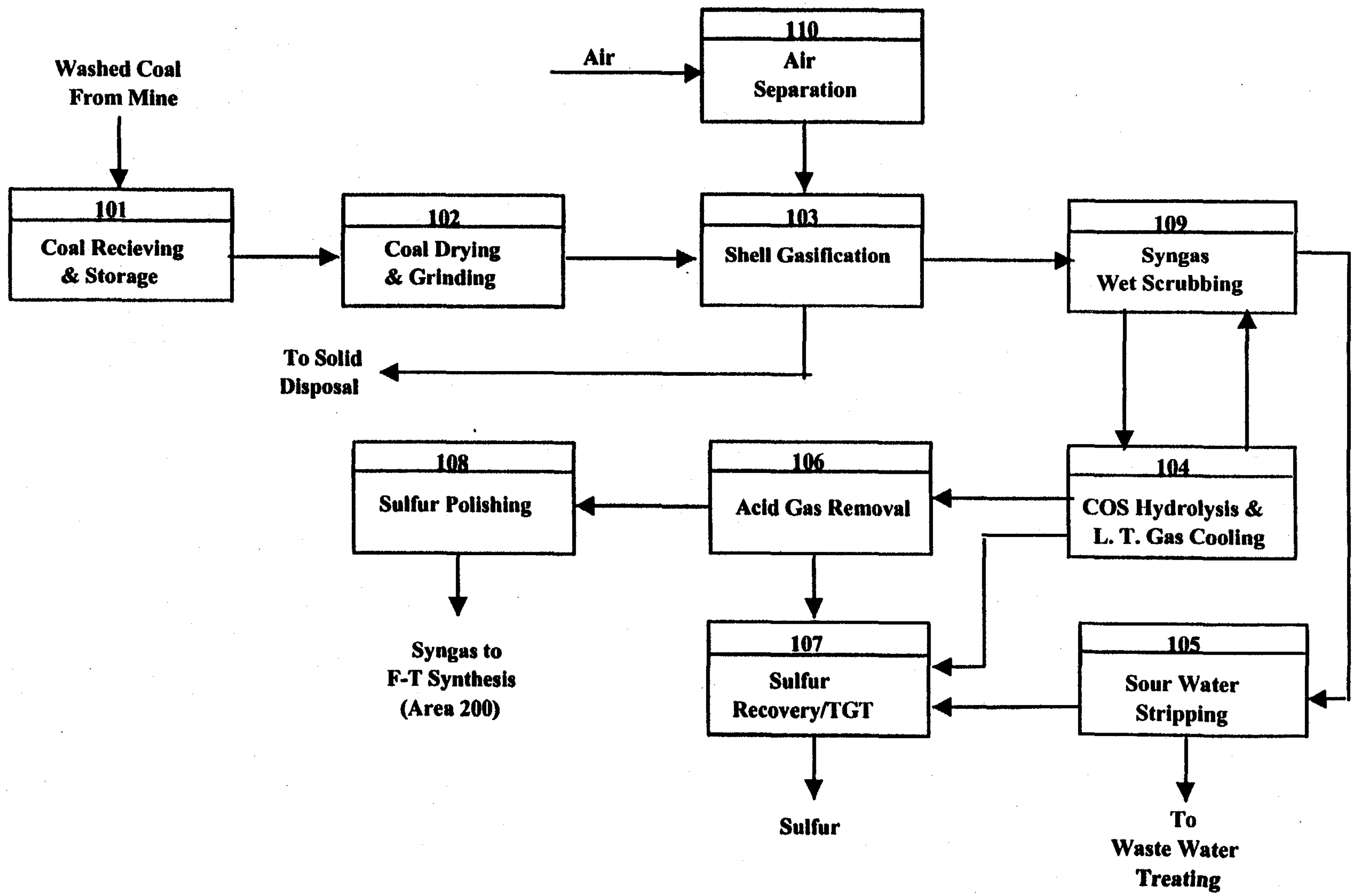

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Figure 2.15

Block Flow Diagram - Eastern Coal\FCC Area 200 (F-T Synthesis Loop)

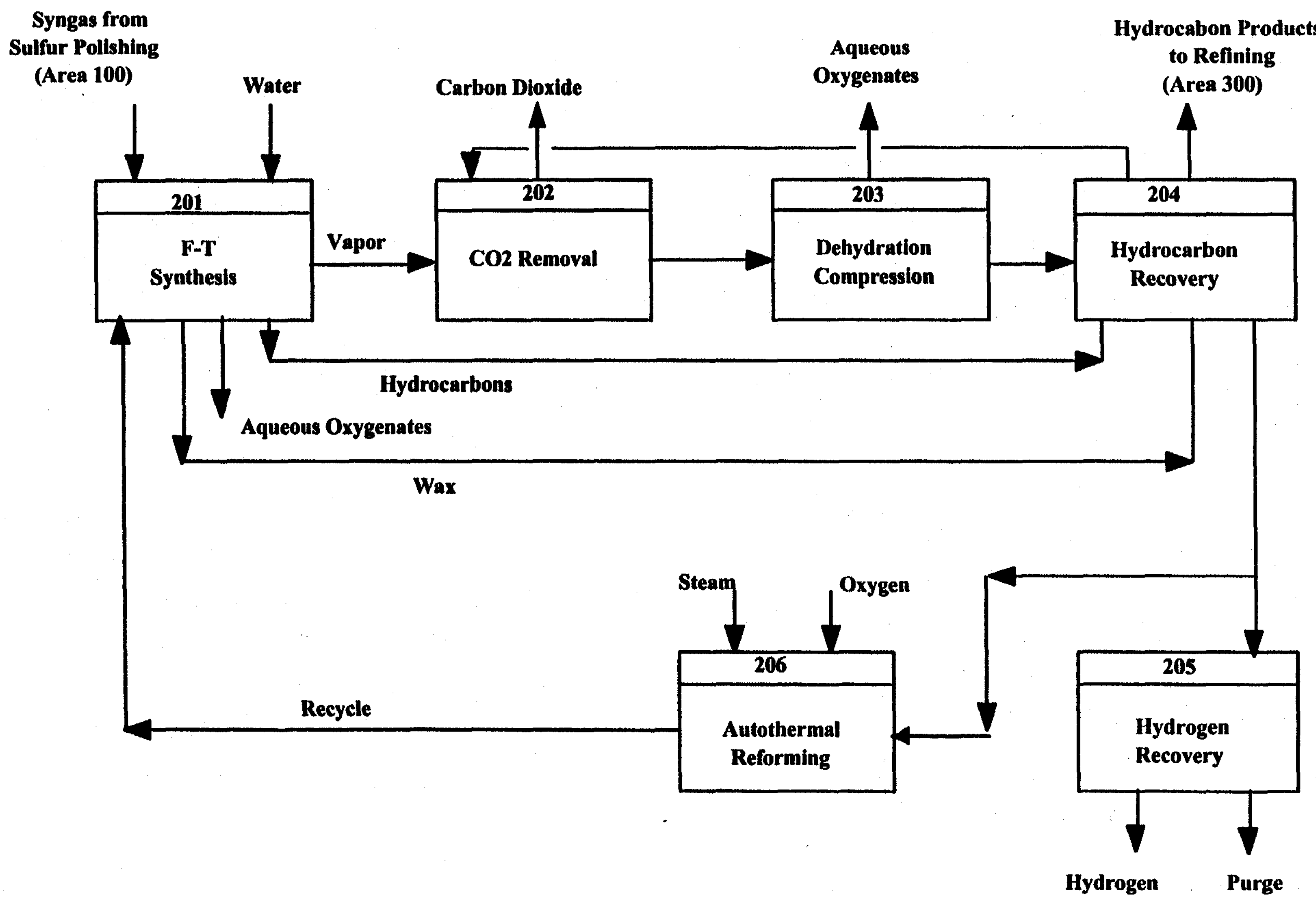

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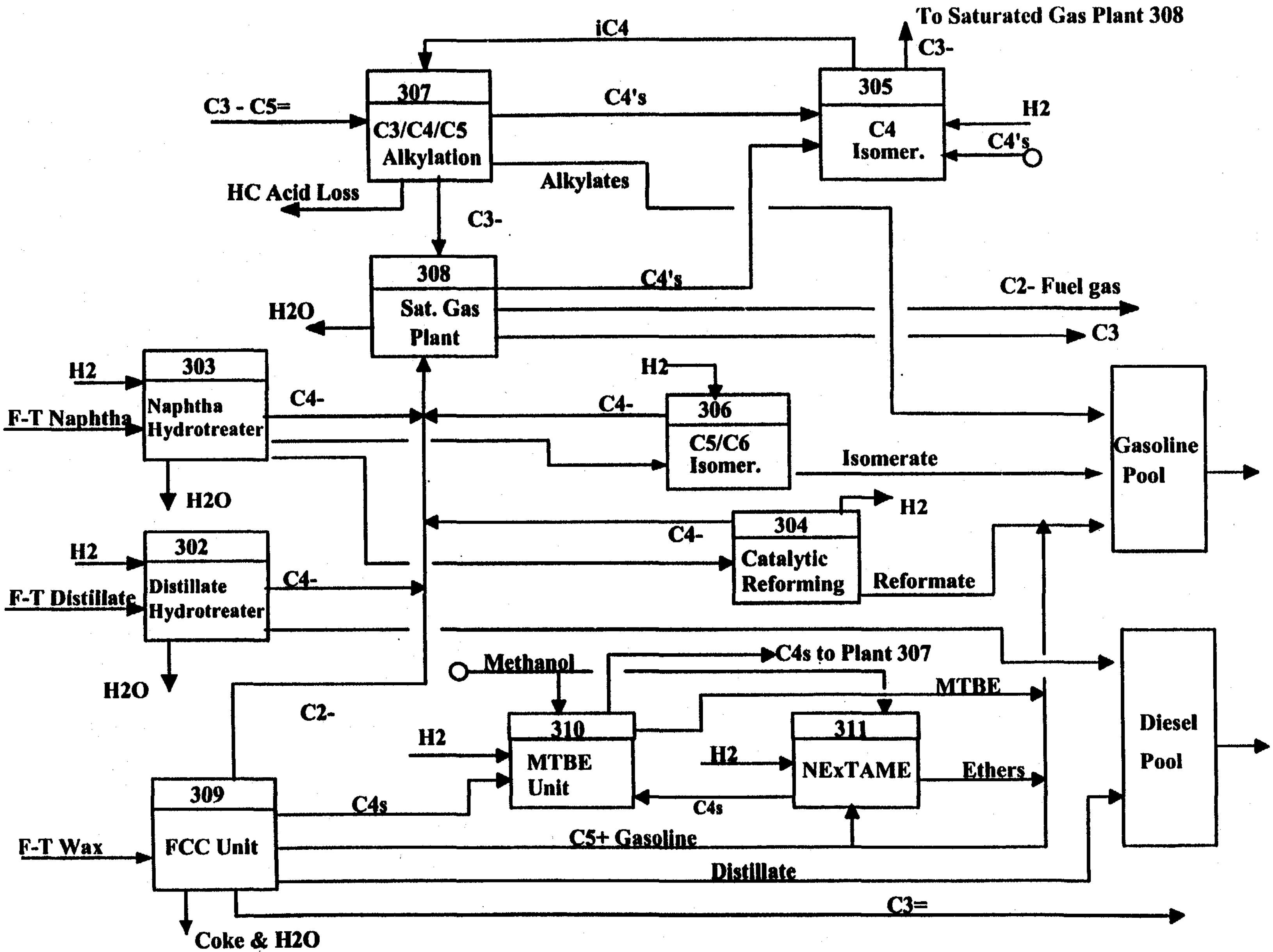
Figure 2.16

Block Flow Diagram - Eastern Coal\FCC Area 300 (Product Upgrading and Refining)



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gas. This FCC plant design includes an unsaturated gas plant which produces a polymer grade propylene stream for sale. No heavy distillate (decanted oil) product is made because all that is produced is burned in the regenerator to supplement the coke production and maintain the unit heat balance.

The design FCC yields for this case are the average yields from two Amoco pilot plant runs, runs 940-1 and 940-2.6. This beta zeolite catalyst produces large amounts of light olefins (13.8 wt % propylene and 22.35 wt% butylenes) and little heavy material (4.7 wt% 430+ °F and coke). About 3.5 wt% of the entering feed has to be consumed as fuel in the regenerator in order for this design to maintain heat balance. This is significantly more than the 1.1 wt% coke deposited on the beta zeolite catalyst. Consequently, the shortfall is made up of all the 650+ °F distillate and some of the 430-650 °F cycle oil. The C7+ naphtha has a 85.4 RON and a 76.4 MON.

Also included in this FCC unit design is an unsaturated gas plant which separates the light gases into a C2- fuel gas stream, a proplylene product stream and a butane/ butenes stream that is sent to the MTBE plant.

Plant 310, the MTBE Plant, converts the isobutylene produced in the FCC plant to MTBE, methyl tertiary-butyl ether, by reacting it with methanol. This plant also contains a selective hydrogenation unit to hydrogenate any diolefins in the feed because they will polymerize, plug and deactivate the etherification catalyst. Some isomerization of the butylenes also occurs in the hydrogenation reactor producing additional isobutylene. The unreacted normal butene and butanes from Plant 310 are sent to Plant 307 for alkylation.

Plant 311, the NExTAME Ethers Plant, converts the C5, C6 and C7 reactive olefins to amyl, hexyl and heptyl ethers by reacting them with methanol. This plant also contains a NExTSELECT selective hydrogenation unit to hydrogenate any diolefins in the feed because they will polymerize, plug and deactivate the etherification catalyst. Some isomerization of the C5, C6 and C7 olefins also occurs in the hydrogenation reactor to produce additional amounts of reactive tertiary olefins. The design of these two units was provided by Neste Engineering in Finland. The NExTAME etherification unit and its integration into a petroleum refinery has been described in two recent papers.[2,3]

The NExTAME etherfied product is a high octane oxygenated gasoline blending component having a 97.0 RON and a 85.2 MON.

The offsites area, which contains the ancillary plants and equipment that is required and does not belong in one of the three main processing areas such as tankage, interconnecting piping, buildings and product shipping, is very similar to that of the Baseline design case.

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[2] Jakkula, J. J., Ignatius, J. and H. Jarvelin, *Increase Oxygenates and Lower Olefins in Gasoline*, Fuel Reformulation 5 (1), 46-54 (January/February 1995).
[3] Tamminen, E. et al, *The Cost of Producing Oxygenates from FCC Gasoline via Two Processing Scemes, 1995 NPRA Annual Meeting, San Francisco, Paper NPRA AM-95-49, March 19-21, 1995.*

Table 2.6 shows the overall material balance and installed plant cost for the FCC wax upgrading with beta zeolite catalyst case with Illinois No. 6 coal. Again, this is the Management Summary Report generated by the ASPEN Plus process flowsheet simulation (PFS) model for this case. This report is shown here because it concisely presents the principal results for the alternate beta zeolite FCC wax upgrading case with Illinois No. 6 coal on a single page.

The plant produces 49,486 bbls/day of gasoline and distillate blending stocks, 5060 bbls/day of propylene, and 1573 bbls/day of LPG from 18,575 tons/day of moisture free Illinois No. 6 coal, 321 tons/day of methanol, and 5204 bbls/day of butanes. The plant also consumes about 58.1 MW of purchased electric power. This power consumption is in addition to the 66.6 MW of power that are produced in Plant 31 from the byproduct steam and fuel gas.

The lower half of Table 2.5 shows the estimated ISBL field cost and total installed cost of each plant as well as the number of dedicated operators required to run it. The total installed cost of each plant consists of the ISBL field cost, an apportioned allotment for the OSBL plants, an amount for home office, engineering and fees, and a contingency allotment. As shown at the bottom of the second cost column, the total cost of the alternate FCC wax upgrading with beta zeolite catalyst case with Illinois No. 6 coal in mid-1993 dollars at the southern Illinois site is 2987 MM$. The annual catalyst and chemicals cost is 37.0 MM$.

Three hundred and eighty three operators are required to run the ISBL plants in this design. An additional 728 operating and maintenance personnel are required for operating the OSBL facilities, extra and spare operators, maintenance, and laboratory personnel making a total labor requirement of 1111 people without management supervision.

As in the previous cases, both the primary liquid hydrocarbon products are nitrogen, sulfur and oxygen free. This case produces significantly more gasoline (39,722 vs. 23,943 bbls/day) and less distillate (9764 vs. 24,686 bbls/day) than the Baseline design case. Also, the gasoline properties from this alternate beta zeolite FCC wax upgrading with Illinois No. 6 coal are different than those of the Baseline case. The gasoline has a 96.8 RON and a 88.9 MON; both of which are significantly higher than those from the previous cases In addition, the gasoline has a lower Reid vapor pressure of 4.7 psi, contains less benzene (0.1 wt%) and contains less aromatics (11.0 wt%) than any of the previously described cases. However, it contains more olefins (12.7 wt%). The diesel blending component has an exceptionally high cetane index of about 74 and a pour point of about -40 °F. This is a superior diesel fuel blending component.

Table 2.6
FCC Wax Upgrading with Beta Zeolite Catalyst Case

MANAGEMENT SUMMARY REPORT

MAJOR INPUT AND OUTPUT STREAMS

INPUT	MLBS/HR	TONS/DAY	
ROM COAL*	1547.933	18575.	
METHANOL	26.789	321.	
NATURAL GAS, MM SCF/HR			0.000
ELECTRIC POWER, MEGA-WH/SD			1395.863
RAW WATER MAKE-UP, MM GAL/SD			15.685

OUTPUT	MLBS/HR	TONS/DAY	BBL/DAY
PROPYLENE	38.371	460.	5060.
PROPANE	11.647	140.	1573.
BUTANES	-44.213	-531.	-5204.
GASOLINE	415.704	4988.	39722.
DIESEL	108.525	1302.	9764.
REFUSE*	0.000	0.	
SLAG*	187.033	2244.	
SULFUR	46.689	560.	
TOTAL	763.756	9165.	50916.

* THESE STREAM FLOW RATES ARE ON A DRY BASIS.
NEGATIVE PRODUCT FLOWS DESIGNATE PURCHASED MATERIAL.

ISBL FIELD AND TOTAL INSTALLED COSTS (INCLUDING OSBL COSTS)

	NUMBER OF PLANTS		PLANT COST, MM$,		DEDICATED
PLANT	OPERATING	SPARES	ISBL	TOTAL	OPERATORS
101	1	0	41.997	63.526	12
102	5	1	101.271	153.186	17
103	8	1	702.888	1063.214	183
104	8	0	37.968	57.432	8
105	1	0	3.213	4.860	0
106	4	0	18.654	28.217	9
107	2	1	43.367	65.599	13
108	8	0	23.731	35.897	0
109	8	0	7.543	11.410	8
110	8	0	326.770	494.284	8
201	8	0	220.380	333.354	43
202A	8	0	16.794	25.403	0
202B	8	0	124.744	188.693	8
203	4	0	17.835	26.977	4
204	4	0	53.675	81.191	4
205	2	0	19.353	29.274	2
206	4	0	21.977	33.243	4
302	1	0	14.001	21.179	4
303	1	0	6.599	9.981	4
304	1	0	22.320	33.762	10
305	1	0	7.805	11.806	4
306	1	0	4.372	6.614	4
307	1	0	52.013	78.677	10
308	1	0	4.801	7.262	4
309	1	0	53.436	80.829	12
310	1	0	14.734	22.288	4
311	1	0	12.168	18.406	4
TOTAL			1974.410	2986.564	383

CATALYST AND CHEMICALS, MM$/YEAR 37.051

DEDICATED PLANT OPERATORS	383
EXTRA OPERATORS, FOREMEN AND MAINTENANCE WORKERS	728
TOTAL	1111

2.6 Case 6 -- Illinois No. 6 Coal with FCC Wax Upgrading with Equilibrium USY Catalyst Case

This and the previous alternative case consider Fluid Catalytic Cracking (FCC) as the means of upgrading the F-T wax instead of hydrocracking. In this case an, equilibrium USY FCC catalyst is used. The FCC yields for these two cases were developed from pilot plant data obtained by Amoco Oil.[1] The equilibrium USY catalyst used in this case is widely used commercially. It does not produce as much C4, C5 and C6 olefins which can be converted to ethers as the beta zeolite catalyst which was used in the previous case. Like the previous FCC upgrading case, this design uses Illinois No. 6 coal with the plant located in southern Illinois. This case is documented in Quarterly Reports of April-September 1995 and October-December 1995, as well as in a paper presented at the ACS National Meeting, Orlando, FL on August 25-29, 1996.

As in the Baseline design case, the ISBL processing area is divided into three main processing areas and an offsites area. Simplified ISBL block flow diagrams for this case are the same as those for the previous case, as shown in Figures 2.14 to 2.16 Since this case also processes the same Illinois No. 6 coal as the Baseline case, Area 100 is identical to that of the Baseline case. Also, Area 200, the Fischer-Tropsch Synthesis Loop, is identical to that of the Baseline design case.

Area 300, the Product Upgrading and Refining Area, for the FCC upgrading cases is the same as the previous FCC upgrading case as described in Section 2.5. It contains ten plants; Plants 302 through 311. In this case, Plant 301, the Wax Hydrocracking Plant, has been replaced by three new plants, Plant 309, 310 and 311. Plant 309, the FCC Plant, is a direct replacement for the wax hydrocracker.

Plant 309, the Fluid Catalytic Cracking Plant, catalytically cracks the F-T wax over an equilibrium USY catalyst to produce distillate, cat cracked naphtha, C4s, C3 olefins and fuel gas. This FCC plant design includes an unsaturated gas plant which produces a polymer grade propylene stream for sale. No heavy distillate (decanted oil) product is made because all that is produced is burned in the regenerator to supplement the coke production and maintain the unit heat balance.

The design FCC yields for the case are the average yields from two Amoco pilot plant runs, runs 939-1 and 940-2.6 This equilibrium USY catalyst does not produce as much light olefins as the beta zeolite catalyst. About 3.8 wt% of the entering feed has to be consumed as fuel in the regenerator in order for this design to maintain heat balance. This is significantly more than the 2.2 wt% coke deposited on the equilibrium USY catalyst. Consequently, the shortfall is made up of all the 650+ °F distillate and some of the 430-650 °F cycle oil. The C7+ naphtha has slightly higher octanes than those of the beta zeolite catalyst case of 86.1 RON and a 77.5 MON.

Also included in this FCC unit design is an unsaturated gas plant which separates the light gases into a C2- fuel gas stream, a proplylene product stream and a butane/ butenes stream that is sent to the MTBE plant.

The offsites area, which contains the ancillary plants and equipment that is required and does not belong in one of the three main processing areas such as tankage, interconnecting piping, buildings and product shipping, is very similar to that of the Baseline design case.

Table 2.7 shows the overall plant summary and installed plant cost for the FCC wax upgrading with equilibrium USY catalyst case with Illinois No. 6 coal. The plant produces 49,297 bbls/day of gasoline and distillate blending stocks, 3215 bbls/day of propylene, and 1584 bbls/day of LPG from 18,575 tons/day of moisture free Illinois No. 6 coal, 209 tons/day of methanol, and 4327 bbls/day of butanes. The plant also consumes about 56.4 MW of purchased electric power. This power consumption is in addition to the 66.6 MW of power that are produced in Plant 31 from the byproduct steam and fuel gas.

The lower half of Table 3.5 shows the ISBL field cost and total installed cost of each plant as well as the number of dedicated operators required to run it. The total installed cost of each plant consists of the ISBL field cost, an apportioned allotment for the OSBL plants, an amount for home office, engineering and fees, and a contingency allotment. As shown at the bottom of the second cost column, the total cost of the alternate FCC wax upgrading with beta zeolite catalyst case with Illinois No. 6 coal in mid-1993 dollars at the southern Illinois site is 2977 MM$. The annual catalyst and chemicals cost is 34.8 MM$.

Three hundred and eighty three operators are required to run the ISBL plants in this design. An additional 728 operating and maintenance personnel are required for operating the OSBL facilities, extra and spare operators, maintenance, and laboratory personnel making a total labor requirement of 1111 people without management supervision.

This case produces significantly more gasoline (39,950 vs. 23,943 bbls/day) and less distillate (9347 vs. 24,686 bbls/day) than the Baseline design case. Again, both the primary liquid hydrocarbon products are nitrogen, sulfur and oxygen free. Also, the gasoline properties from this alternate FCC wax upgrading with equilibrium USY catalyst case with Illinois No. 6 coal are different than those of the Baseline case. The gasoline has a 95.8 RON and a 87.8 MON; both of which are significantly higher than those from the conventional refining cases but less than the beta zeolite FCC case. In addition, the gasoline has a Reid vapor pressure of 4.8 psi, contains less benzene (0.1 wt%) and contains less aromatics (13.9 wt%) than any of the conventional refining cases. However, it contains more olefins (15.5 wt%). The diesel blending component has the same properties as that from the beta zeolite FCC case. It has an exceptionally high cetane index of about 74 and a pour point of about -40 °F. This is a superior diesel fuel blending component.

Table 2.7
FCC Wax Upgrading with Equilibrium USY Catalyst Case

MANAGEMENT SUMMARY REPORT

MAJOR INPUT AND OUTPUT STREAMS

INPUT	MLBS/HR	TONS/DAY	
ROM COAL*	1547.933	18575.	
METHANOL	17.425	209.	
NATURAL GAS, MM SCF/HR			0.000
ELECTRIC POWER, MEGA-WH/SD			1354.201
RAW WATER MAKE-UP, MM GAL/SD			15.579

OUTPUT	MLBS/HR	TONS/DAY	BBL/DAY
PROPYLENE	24.375	293.	3215.
PROPANE	11.728	141.	1584.
BUTANES	-36.759	-441.	-4327.
GASOLINE	418.549	5023.	39950.
DIESEL	104.957	1259.	9347.
REFUSE*	0.000	0.	
SLAG*	187.033	2244.	
SULFUR	46.689	560.	
TOTAL	756.573	9079.	49769.

* THESE STREAM FLOW RATES ARE ON A DRY BASIS.
NEGATIVE PRODUCT FLOWS DESIGNATE PURCHASED MATERIAL.

ISBL FIELD AND TOTAL INSTALLED COSTS (INCLUDING OSBL COSTS)

PLANT	NUMBER OF PLANTS OPERATING	SPARES	PLANT COST, MM$, ISBL	TOTAL	DEDICATED OPERATORS
101	1	0	41.997	63.531	12
102	5	1	101.271	153.198	17
103	8	1	702.888	1063.294	183
104	8	0	37.968	57.436	8
105	1	0	3.213	4.860	0
106	4	0	18.654	28.219	9
107	2	1	43.367	65.604	13
108	8	0	23.731	35.899	0
109	8	0	7.543	11.411	8
110	8	0	326.863	494.462	8
201	8	0	221.463	335.018	43
202A	8	0	16.828	25.457	0
202B	8	0	124.993	189.083	8
203	4	0	17.872	27.036	4
204	4	0	53.772	81.344	4
205	2	0	19.416	29.372	2
206	4	0	22.302	33.737	4
302	1	0	14.046	21.249	4
303	1	0	6.622	10.017	4
304	1	0	22.392	33.873	10
305	1	0	7.092	10.728	4
306	1	0	4.387	6.636	4
307	1	0	47.953	72.541	10
308	1	0	5.153	7.795	4
309	1	0	52.222	78.999	12
310	1	0	10.556	15.969	4
311	1	0	13.791	20.862	4
TOTAL			1968.355	2977.630	383

CATALYST AND CHEMICALS, MM$/YEAR 34.838

DEDICATED PLANT OPERATORS	383
EXTRA OPERATORS, FOREMEN AND MAINTENANCE WORKERS	728
TOTAL	1111

2.7 Case 7 -- Fischer-Tropsch Liquefaction of Natural Gas Case

Figure 2.17 is a block flow diagram showing the overall process configuration of this natural gas Baseline F-T design. In this Fischer-Tropsch liquefaction of natural gas case, the syngas is prepared from natural gas in contrast to all the previous cases where it was prepared by gasification of coal. A combination of the partial oxidation (POX) and steam reforming processes are used to generate the syngas for the F-T synthesis. Furthermore, like the previous Illinois No. 6 coal cases, the plant is located in southern Illinois. This case is documented in Topical Report VI, *Natural Gas Fischer-Tropsch Case*, Volume I, *Summary Report*, and Volume II, *Plant Design and ASPEN Process Simumaltion Model* (references TR-8 and TR-9).

As in all the previous cases, the ISBL processing area is divided into three main processing areas. In addition there is a offsites area which contains the ancillary plants and equipment that is required and does not belong in one of the three main processing areas such as tankage, interconnecting piping, buildings and product shipping. Simplified ISBL block flow diagrams for this natural gas Baseline F-T design are shown in Figures 2.18, 2.19 and 2.20 respectively for the three main processing area of 100, 200 and 300.

Area 100, the syngas production area is entirely different for this natural gas case. It consists of only three plants, Plants 110, 111 and 112.

Plant 110, the Air Separation Plant, primarily provides the required oxygen for Plant 111, the Partial Oxidation Plant. Plant 110 consists of six parallel trains each producing 2100 stpd of 99.5 mole% oxygen. No spare trains are provided. However, the design includes a backup system consisting of a liquid oxygen storage facility with a capacity of one full days production and a gaseous oxygen storage facility with a capacity of 40 minutes of production from a single train.

Plant 111, the Syngas Production by Partial Oxidation (POX) Plant, consumes about 97.8 % of the natural gas feed and produces 45.3 MMSCF/hr of syngas. Plant 111 consists of 24 POX reactor trains, each producing about 1.9 MMSCF/hr of syngas with a molar H_2/CO ratio of about 1.79/1. Plant 111 is divided into three sections. The first section, 111A, consists of two CoMo/ZnO reactors per train to remove any sulfur present in the natural gas feed. Section 111A is similar to Plant 108 except that it treats the natural gas feed rather than the syngas product. The cleaned natural gas is mixed with oxygen and 600 psig steam and sent to the non-catalytic POX reactors in section 111B. The syngas leaves the POX reactors at about 400 psig and 2550 °F. The syngas is cooled in section 111C by generating 600 psig steam, washed and dehumidified by cooling to 150 °F before being reheated and sent to the F-T synthesis loop.

Plant 112, the Syngas Production by Steam Reforming Plant, consumes about 3.2 % of the natural gas feed and produces 2.4 MMSCF/hr of syngas with a molar H_2/CO ratio of about 5.89/1. It contains only a single train. As is the case with Plant 111, Plant 112

Figure 2.17
NATURAL GAS F-T BASELINE DESIGN
OVERALL PROCESS CONFIGURATION

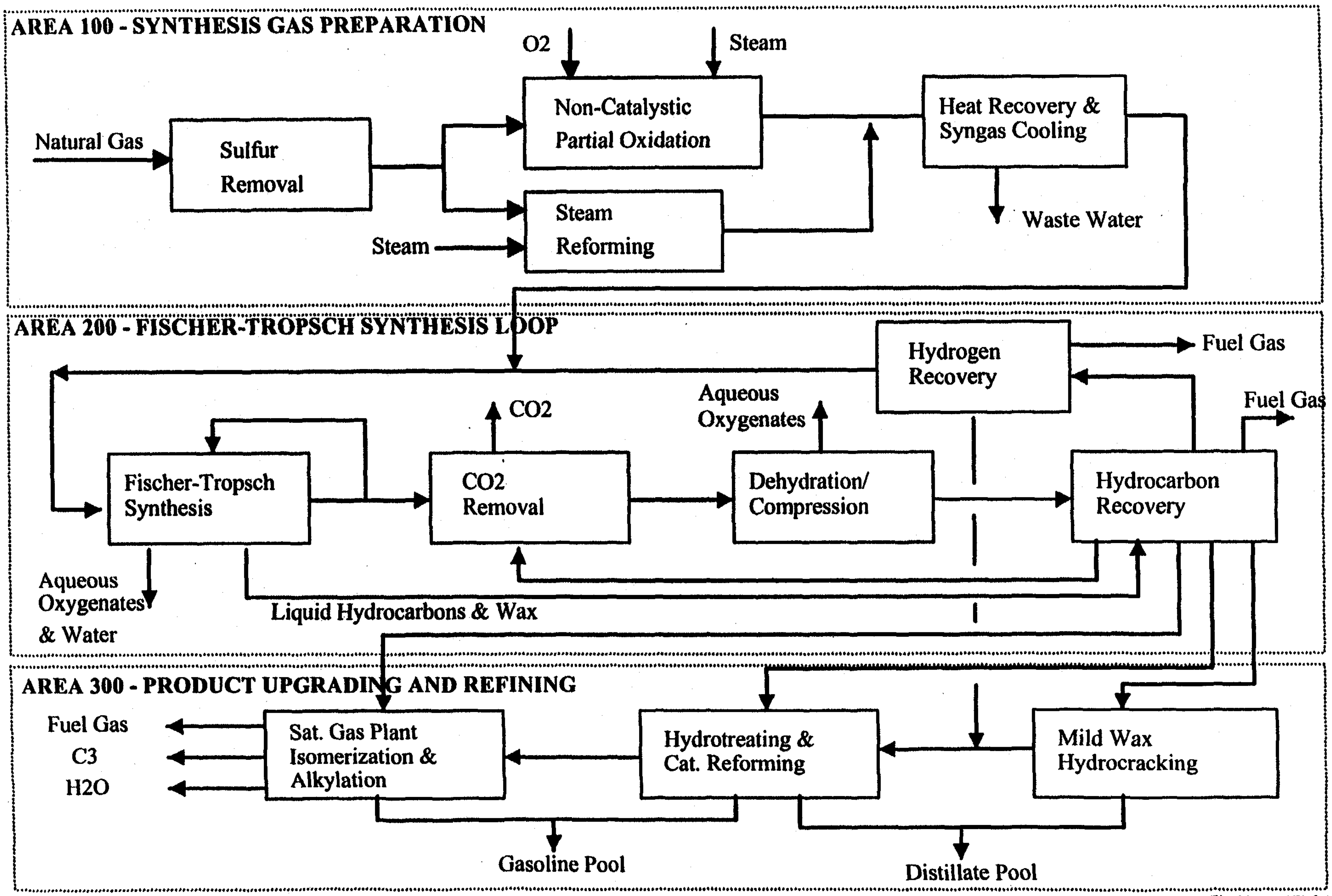

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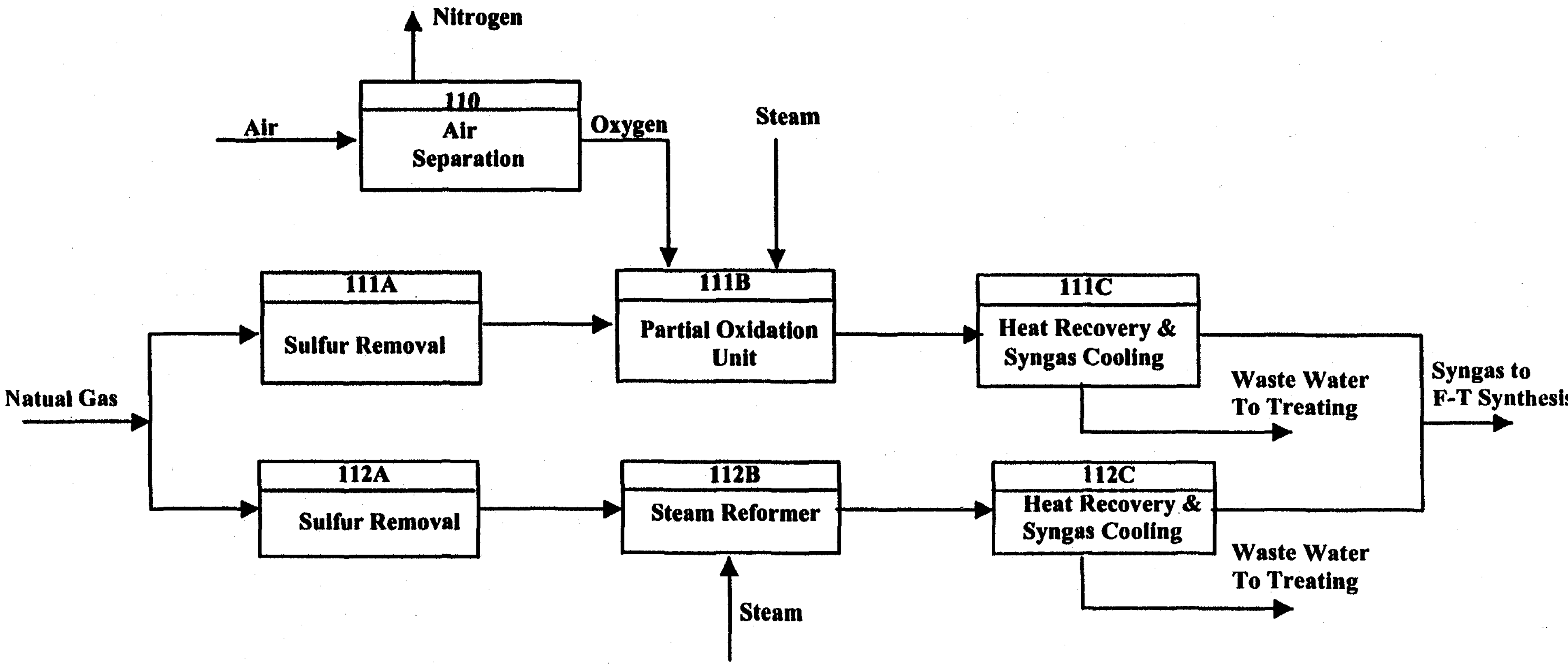
Figure 2.18

Block Flow Diagram - Natural Gas Baseline Design Area 100 (Syngas Production)



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<u>**Figure 2.19**</u>

Block Flow Diagram - Natural Gas Baseline Design Area 200 (F-T Synthesis)

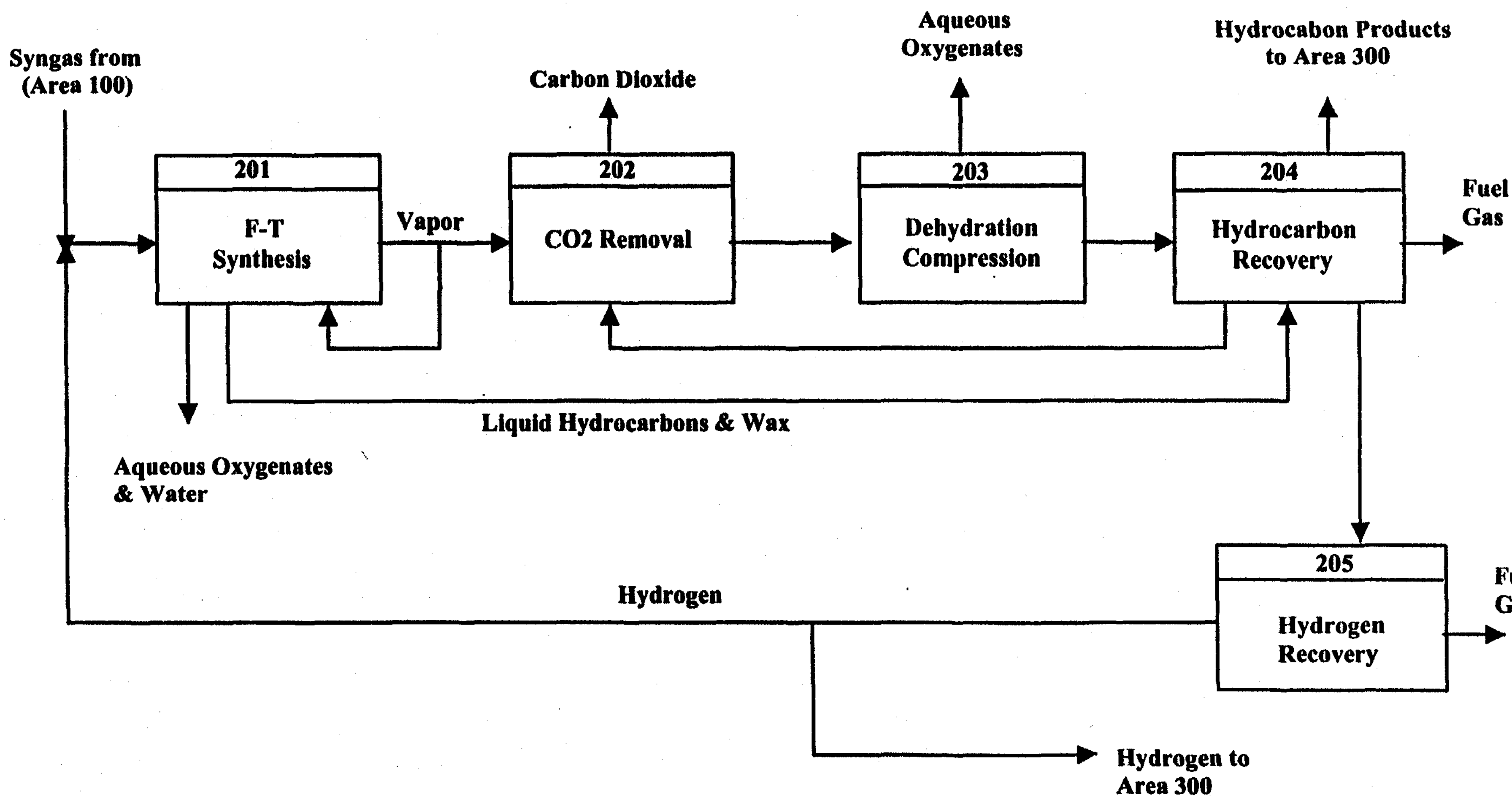

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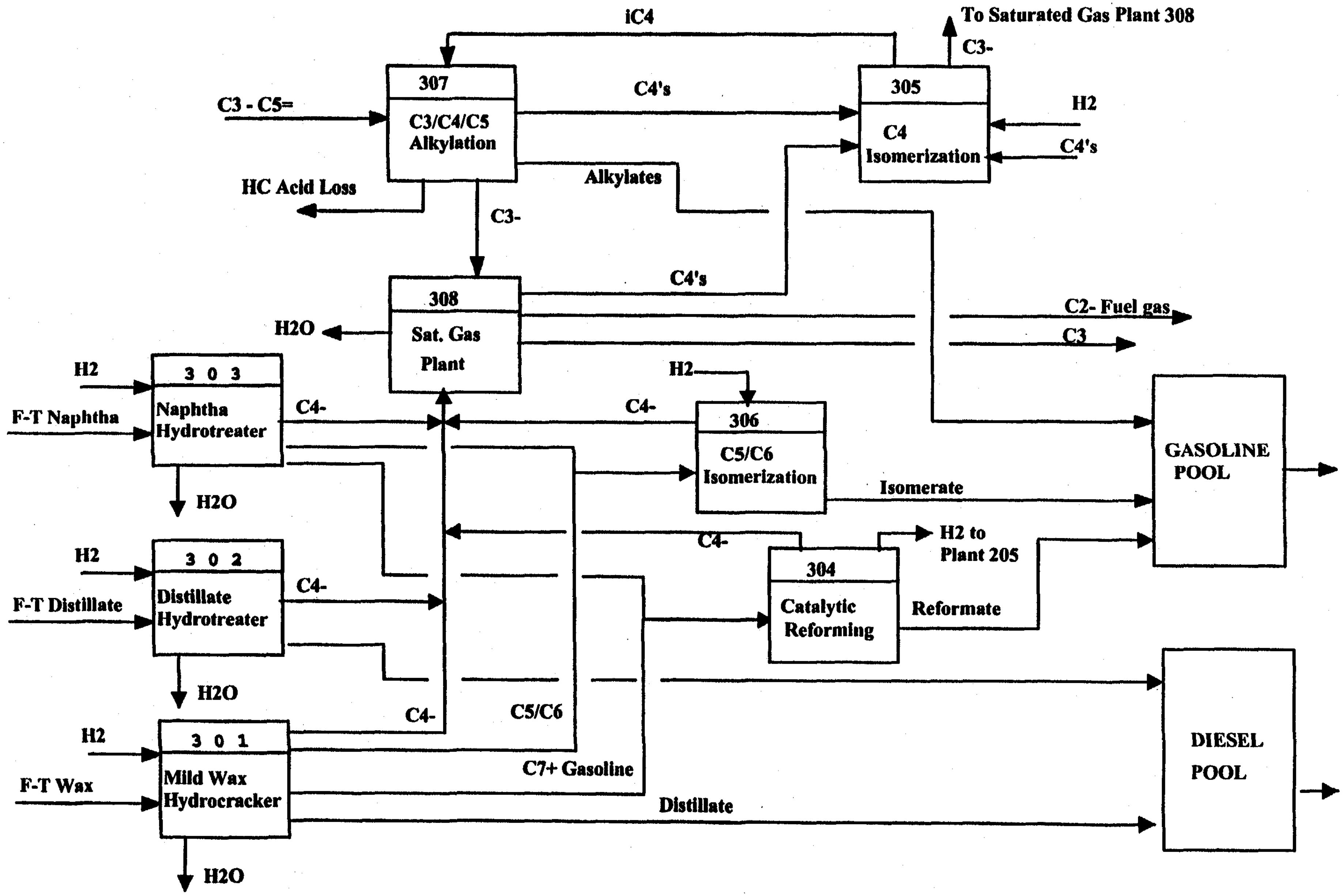
Figure 2.20

Block Flow Diagram - Natural Gas Baseline Design Area 300 (Upgrading and Refining)



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also consists of three sections. The first section, 112A, consists of two CoMo/ZnO reactors per train to remove any sulfur present in the natural gas feed. Section 112A is similar to Plant 108 except that it treats the natural gas feed rather than the syngas product. The cleaned natural gas is mixed with 600 psig steam and sent to the catalytic steam reforming reactors in section 112. The syngas leaves the POX reactors at about 400 psig and 1650 °F. The syngas is cooled in section 112C by generating 600 psig steam, washed and dehumidified by cooling to 150 °F before being reheated and sent to the F-T synthesis loop.

The combined syngas from Plants 111 and 112 entering Area 200 has a molar H_2/CO ratio of 1.87/1.

Area 200, the Fischer-Tropsch Synthesis Loop, consists of five plants. This area is significantly different from the corresponding Area 200 of the Baseline design case for Illinois No. 6 coal which consists of six plants; whereas, in this natural gas case, it consists of only five plants. Plant 201, the F-T Synthesis Plant, is different because it uses a cobalt based F-T synthesis catalyst rather than an iron based one. In the Baseline case, hydrogen is recovered from only a portion of the unconverted syngas which contrasts with this natural gas case where hydrogen is recovered from all the unconverted syngas. Furthermore, in the Baseline case, only a portion of the unconverted syngas is used for fuel; whereas, in this case, all of the unconverted syngas after hydrogen recovery is used for fuel. In effect the unconverted syngas recycle stream in all the previous cases is now an essentially pure hydrogen recycle stream. Consequently, Plant 206, the Autothermal Reforming Plant, is eliminated.

In this natural gas case, Plant 201, the Fischer-Tropsch Synthesis Plant, uses a cobalt based F-T synthesis catalyst in contrast to all the other cases which use an iron based catalyst. Cobalt based F-T synthesis catalyst was selected for this case because it has negligible activity for the water-gas shift reaction compared to iron based catalyst, and thus, it requires a syngas with a molar H_2/CO ratio near the stoichiometric value of 2.0/1. Since methane, the principal component in natural gas, has a molar H_2/CO ratio of 2.0/1, syngas produced from it has similar H_2/CO ratio. Also, for iron based catalyst with a high water-gas shift activity, CO_2 is the primary byproduct from the Fischer-Tropsch synthesis. With cobalt catalyst, water is the primary byproduct because the cobalt catalyst only has slight activity for the water-gas shift reaction which converts CO and H_2O to CO_2 and H_2.

The first-stage F-T slurry bed reactor operates at 428 °F and 335 psig, and the second-stage reactor operates at 428 °F and 290 psig. Excess heat is removed by the generation of 150 psig from tubes within the reactor. By comparison, the iron based catalyst cases generate 440 °F, 360 psig steam because the reactors operate at 488 °F.

In order to prevent the formation of a liquid water phase within the F-T reactors, the raw syngas entering Area 200 is dried by cooling it to 150 °F to condense and remove water. With cobalt based F-T catalyst, the reactor configuration also has been changed.

For iron based catalyst, each train contains three parallel slurry reactors. In this case, the unconverted syngas leaving the two parallel first-stage reactors is cooled to 150 °F to condense and remove water before being fed feed a single second-stage reactor. In the first-stage reactor, the CO conversion/pass is 56 %, and it is 59 % in the second-stage reactor resulting in an overall CO conversion/pass of about 82 %. An unconverted syngas recycle stream around Plant 201 increases the CO conversion to about 91 %. In the Baseline design, the CO conversion/pass is 86.9 %.

Area 300, the Product Upgrading and Refining Area, for this Fischer-Tropsch liquefaction of natural gas case is identical to that of the Baseline design.

The offsites area for this case is somewhat different than that of the Baseline design case. It consists of eighteen plants. These plants are similar to those which were developed for the Baseline coal liquefaction case with minor modifications as required for this natural gas case. The most notable changes involved:

1. The removal of the facilities for handling a liquid sulfur product from Plants 20 and 23 since this case does not produce liquid sulfur.
2. The removal of Plant 24, Coal Ash Disposal.
3. The interconnecting piping system of Plant 24 was modified to remove the 900 and 360 psig steam headers and to add a 150 psig steam header.
4. Plant 31, Steam and Power Generation, and Plant 32, Raw, Cooling and Potable Water, were redesigned to match the requirements for this case.

Table 2.8 shows the overall material balance and installed plant cost for the Fischer-Tropsch liquefaction of natural gas case. Again, this is the Management Summary Report generated by the ASPEN Plus process flowsheet simulation (PFS) model for this case. This report is shown here because it concisely presents the principal results for this natural gas case on a single page.

The plant produces 43,238 bbls/day of gasoline and distillate blending stocks and 1704 bbls/day of LPG from 412 MMSCF/day of natural gas and 340 bbls/day of butanes. The plant also produces about 24.7 MW of surplus electric power for sale.

The lower half of Table 2.8 shows the ISBL field cost and total installed cost of each plant as well as the number of dedicated operators required to run it. The total installed cost of each plant consists of the ISBL field cost, an apportioned allotment for the OSBL plants, an amount for home office, engineering and fees, and a contingency allotment. As shown at the bottom of the second cost column, the total cost of the Fischer-Tropsch liquefaction of natural gas case in mid-1993 dollars at the southern Illinois site is 1842 MM$. The annual catalyst and chemicals cost is 21.8 MM$.

Two hundred and six operators are required to run the ISBL plants in this design. An additional 780 operating and maintenance personnel are required for operating the OSBL facilities, extra and spare operators, maintenance, and laboratory personnel making a total labor requirement of 598 people without management supervision.

Table 2.8

Fischer-Tropsch Liquefaction of Natural Gas Case

MANAGEMENT SUMMARY REPORT

MAJOR INPUT AND OUTPUT STREAMS

INPUT	MLBS/HR	TONS/DAY	
ROM COAL*	0.000	0.	
METHANOL	0.000	0.	
NATURAL GAS, MM SCF/HR			17.163
RAW WATER MAKE-UP, MM GAL/SD			20.980

OUTPUT	MLBS/HR	TONS/DAY	BBL/DAY
PROPANE	12.618	151.	1704.
BUTANES	-2.891	-35.	-340.
GASOLINE	179.404	2153.	17027.
DIESEL	295.184	3542.	26211.
REFUSE*	0.000	0.	
SLAG*	0.000	0.	
SULFUR	0.000	0.	
TOTAL	484.315	5812.	44602.

ELECTRIC POWER, MEGA-WH/SD 591.973

* THESE STREAM FLOW RATES ARE ON A DRY BASIS.
NEGATIVE PRODUCT FLOWS DESIGNATE PURCHASED MATERIAL.

NATURAL GAS TO P111 & P112 17.163 MM SCF/HR 17797.7 MM BTUS/HR

ISBL FIELD AND TOTAL INSTALLED COSTS (INCLUDING OSBL COSTS)

	NUMBER OF PLANTS		PLANT COST, MM$,		DEDICATED
PLANT	OPERATING	SPARES	ISBL	TOTAL	OPERATORS
110	6	0	247.269	427.689	6
111	24	0	438.766	758.913	96
112	1	0	20.961	36.255	4
200	8	0	159.960	276.675	43
202A	2	0	4.130	7.143	0
202B	1	0	14.624	25.295	1
203	2	0	8.196	14.177	2
204	2	0	25.011	43.261	2
205	2	0	14.310	24.751	2
301	1	0	41.323	71.474	10
302	1	0	16.812	29.079	4
303	1	0	5.651	9.774	4
304	1	0	28.477	49.256	10
305	1	0	4.139	7.158	4
306	1	0	6.493	11.230	4
307	1	0	23.833	41.224	10
308	1	0	5.300	9.168	4
TOTAL			1065.255	1842.521	206

CATALYST AND CHEMICALS, MM$/YEAR 21.858

DEDICATED PLANT OPERATORS	206
EXTRA OPERATORS, FOREMEN AND MAINTENANCE WORKERS	392
TOTAL	598

Both the primary liquid hydrocarbon products are nitrogen, sulfur and oxygen free. The gasoline blending component has similar properties as those for the Baseline design case. It has a 90.6 RON, a 85.2 MON, and a Reid vapor pressure of 5.6 psi, and it contains a small amount of benzene (0.4 wt%), contains an insignificant amount of olefins (0.01 wt%), and contains a significant amount of aromatics (33.5 wt%). The diesel blending component also has about the same properties as that from the Baseline case. It has an exceptionally high cetane index of about 75 and a pour point of about -31 °F. This is a superior diesel fuel blending component.

2.8 Case 8 -- Once Through F-T Liquefaction of Natural Gas with Power Co-Production Case

Figure 2.21 is a block flow diagram showing the overall process configuration of this once through Fischer-Tropsch design with power co-production. In this F-T liquefaction of natural gas case, the syngas is prepared from natural gas as in the previous case. However, in this case it is prepared by autothermal reforming using enriched air containing 40% oxygen. Furthermore, this is a once-through design to reduce cost. Also, minimal product upgrading is used only to produce a shippable syncrude instead of the gasoline and diesel blending components which were produced in the previous cases. This case is documented in Topical Report VII, *Natural Gas Fischer-Tropsch Design with Power Coproduction*, (reference TR-10)

In contrast to the previous cases, the ISBL processing area is divided into two main processing areas. Because power coproduction is a key component of this plant design, the combined-cycle power plant is considered an ISBL plant and located in Area 200. In addition there is a offsites area which contains the ancillary plants and equipment that is required and does not belong in one of the two main processing areas.

Because this case is so different from all the previous cases, the ISBL processing plants have been renumbered. Thus, the plant numbers for this case do <u>not</u> correspond to the plant numbers used for all the previous cases

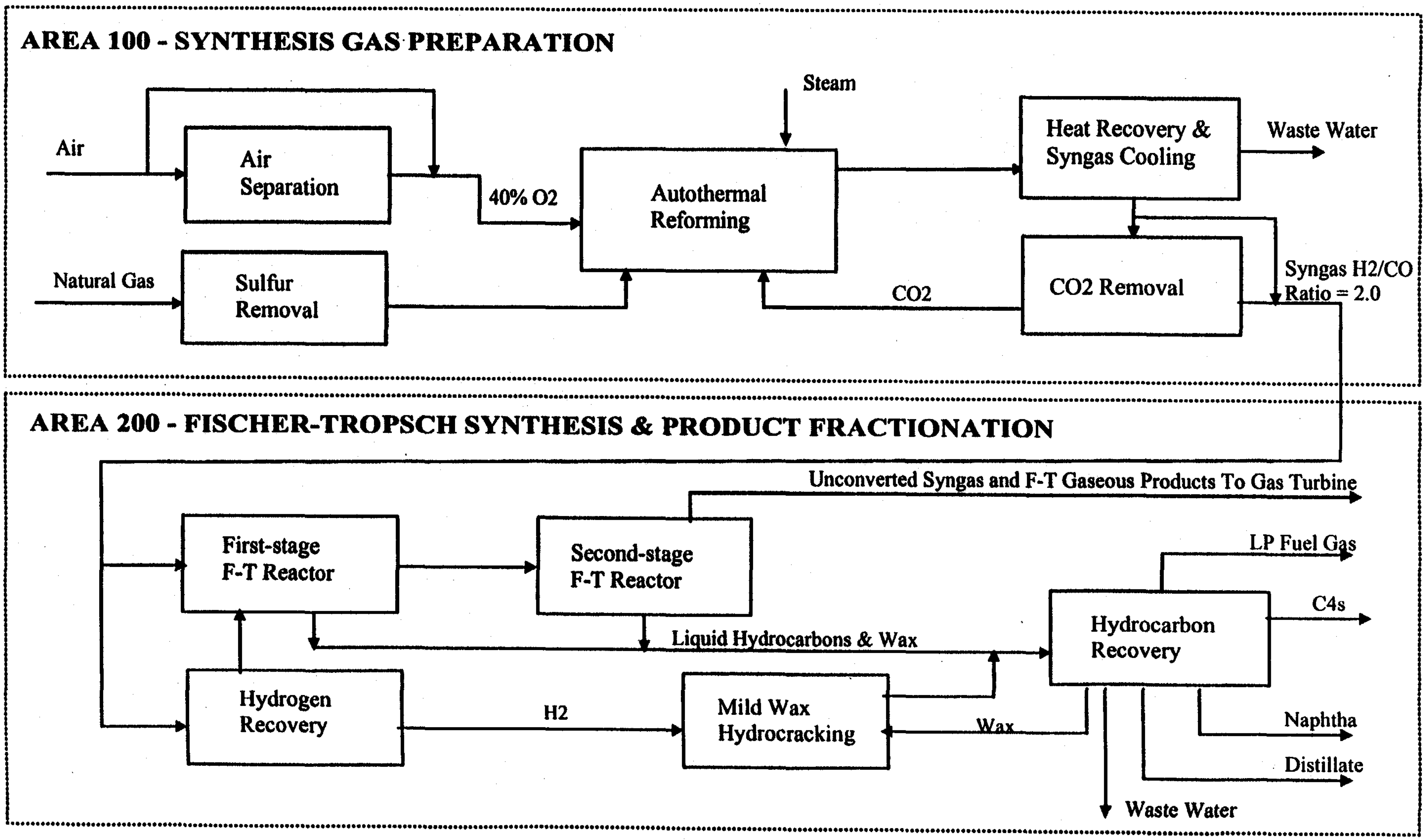
Area 100, the syngas production area is different for this once-through natural gas case from the previous natural gas case. It consists of the following three plants

Plant 101	Air Separation Plant
Plant 102	Autothermal Reforming Plant
Plant 103	CO_2 Removal and Recycle Plant

Plant 101, the Air Separation Plant, also contains an air compressor. The air separation section of Plant 101 produces a 95 mole % oxygen stream which when blended with the output from the air compressor produces an enriched air stream containing 40 mole % oxygen that is sent to Plant 102, the Autothermal Reforming Plant.

Plant 102, the Autothermal Reforming Plant, first removes trace amounts of sulfur containing compounds from the natural gas by reaction with zinc oxide. The desulfurized natural gas then is mixed with the enriched air, recycle CO_2 and steam before entering the autothermal reformer reactor where it is converted to syngas. The hot syngas product stream is cooled by feed/effluent heat exchange, steam generation, and exchange with ambient air. It is then flashed to condensed and remove water before going on to Plant 103, the CO_2 Removal and Recycle Plant.

Figure 2-21
Once-Through Fischer-Tropsch (F-T) Design With Power Coproduction
(Overall Process Configuration)



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Plant 103, the CO_2 Removal and Recycle Plant, is a standard MDEA absorption plant which removes CO_2 from a portion of the cooled syngas stream leaving Plant 102 and recycles it back to Plant 102 to control the CO_2 production within the autothermal reformer reactor. The lean syngas after CO_2 removal is mixed with that bypassing Plant 103 to produce a combined synthesis gas steam having a molar H_2/CO ratio of 2.01 which is to Plant 201, the Fischer-Tropsch Synthesis Plant.

In this once-through natural gas case, Area 200 is a combination of the previous Fischer-Tropsch Synthesis Loop and the Product Upgrading and Refining Areas with the addition of the Combined-Cycle Power Plant. It consists of the following five plants:

Plant 201	Fischer-Tropsch Synthesis Plant
Plant 202	Hydrogen Recovery Plant
Plant 203	Hydrocarbon Recovery Plant
Plant 204	Wax Hydrocracking Plant
Plant 31	Combined-Cycle Power Plant

This Area 200 is significantly different from the Fischer-Tropsch synthesis areas in all the previous cases. In each of the other two cases, a recycle design was used to obtain maximum liquids yields. Thus, the Fischer-Tropsch synthesis area contained a CO_2 removal plant and a recycle compression and dehydration plant preceding the hydrocarbon recovery plant which, using refrigeration, recovered LPG in addition to all the heavier hydrocarbons. The removal of these materials from the F-T synthesis loop is necessary in order to minimize the amount of unconverted syngas being recycled and both equipment and operating cost associated with the recycle loops.

In this once-through design, the F-T products are recovered in Plant 203 immediately after they are produced in Plant 201. All unconverted syngas is used for power production in the combined-cycle power Plant.

Plant 201, the Fischer-Tropsch Synthesis Plant, contains a two-stage reactor section with interstage cooling and withdrawal of condensed liquids. Because this case uses an enriched air-blown autothermal reformer for syngas generation, the unconverted syngas leaving the second-stage, F-T synthesis reactor contains a significant amount of nitrogen which essentially precludes a recycle design and makes recovery of the C3 and C4 hydrocarbons from the unconverted syngas both more difficult and more expensive. Thus, a less comprehensive hydrocarbon recovery plant, Plant 203, is used which does not recover LPG. However, a LiBr absorption chiller is used to recover additional hydrocarbons from the unconverted syngas leaving the Fischer-Tropsch Synthesis Plant before it becomes high pressure fuel gas.

Plant 203, the Hydrogen Recovery Plant, supplies the hydrogen needs of the wax hydrocracking plant by recovering hydrogen from only a small portion of the syngas generated in the syngas preparation area. In contrast to the other designs, the feed to the hydrogen recovery plant is fresh syngas from Area 100 instead of the unconverted syngas leaving the hydrocarbon recovery plant. This design change was required because the hydrogen concentration in the unconverted syngas is insufficient for effective, economic hydrogen recovery.

Plant 204, the small wax hydrocracking plant is the only product upgrading step included in this design. In the previous cases, the wax hydrocracking plant was part of a separate and more extensive product upgrading and refining area which produced high quality gasoline and distillate blending components. In this case, the objective is only to produce a stable, shippable product which can be upgraded to liquid transportation fuels elsewhere, such as in a conventional petroleum refinery.

Plant 31, the Combined-Cycle Power Plant, is included in the ISBL plants of Area 200 because it is such an important plant in this design. It consumes the C4- fuel gas produced in the rest of the plant as well as the excess 635 psig steam from Plant 102 and the excess 150 psig stream from Plant 201 to produce electric power, compress the inlet air to Plant 101 to 665 psia, and produce 50 psig steam for use within the process area. The fuel gas is burned in a General Electric (GE) Frame 7 gas turbine which drives the air compressors and an electric power generator. The hot exhaust gases from the air compressor and the inlet steams from Plant 102 and 201 go to the HRSG (Heat Recovery Steam Generation) section where superheated three levels of steam are generated and sent to the three-stage steam turbine which drives another power generator. 50 psig steam is withdrawn from the low pressure turbine for use in other parts of the plant.

Plant 31 supplies the complete electric power needs for the entire complex as well as producing an extra 84 MW for sale.

The offsites area for this case is similar to that of the previous natural gas case except that it does not contain the power plant. It consists of fourteen plants. These plants are similar to those which were developed for the Baseline coal liquefaction case with minor modifications as required for this once-through natural gas case.

Table 2.9 shows the overall material balance and installed plant cost for this once-through, natural gas Fischer-Tropsch case. This report is presented in the same style as the previous reports even though this case was not completely modeled in the ASPEN Plus process flowsheet simulation (PFS) program.

The plant produces 8815 bbls/day of synthetic crude and 84 MW of surplus electric power for sale. Approximately 146 bbls/day are butanes, 2933 bbl/day are in the naphtha boiling range, and 5736 bbls/day are in the distillate boiling range.

The lower half of Table 2.9 shows the plant description, total installed cost of each plant as well as the number of dedicated operators required to run it. As shown at the bottom of the second cost column, the total cost of this once-through, natural gas

Table 2.9

Once Through F-T Liquefaction of Natural Gas with Power Co-Production

MAJOR INPUT AND OUTPUT STREAMS

INPUT	MLBS/HR	TONS/DAY	
NATURAL GAS, MM SCF/HR			4.167
RAW WATER MAKE-UP, MM GAL/SD			5.3

OUTPUT	MLBS/HR	TONS/DAY	BBL/DAY
PROPANE	0.0	0.	0.
BUTANES	1.220	15.	146.
NAPHTHA	30.295	364.	2933.
DIISTILLATE	64.603	775.	5736.
SULFUR	0.000	0.	
TOTAL	96.118	1154.	8815.
ELECTRIC POWER, MEGA-WH/SD			2018.4

TOTAL INSTALLED COSTS (INCLUDING OSBL COSTS)

PLANT	DESCRIPTION	TOTAL PLANT COST, MM$	DEDICATED OPERATORS
101	AIR SEPARATION (40% O2)	70.42	2
102	AUTOTHERMAL REFORMING	22.82	6
103	CO2 REMOVAL AND RECYCLE	13.42	1
201	FISCHER-TROPSCH SYNTHESIS	35.75	6
202	HYDROGEN RECOVERY	3.63	1
203	PRODUCT FRACTIONATION	3.19	1
204	WAX HYDROCRACKING	11.83	3
31	COMBINED CYCLE PLANT	120.32	*
	OTHER OFFSITES	25.19	*
	SUBTOTAL	335.89	
	HOME OFFICE AND FEES	25.19	
	CONTINGENCY	54.16	
TOTAL		415.25	20

CATALYST AND CHEMICALS, MM$/YEAR 6.05

NOTE: THE ABOVE PLANT NUMBERS DO NOT CORRESPOND WITH THE PLANT NUMBERS USED IN THE PREVIOUS CASES.

DEDICATED PLANT OPERATORS	20
EXTRA OPERATORS, FOREMEN AND MAINTENANCE WORKERS	40
TOTAL	60

* OPERATORS ARE INCLUDED IN THE EXTRA OPERATORS CLASSIFICATION.

Fischer-Tropsch liquefaction plant in mid-1993 dollars at a hypothetical U. S. Gulf Coast site is 415 MM$. The annual catalyst and chemicals cost is 6.05 MM$.

Twenty operators are required to run the ISBL plants in this design. An additional forty operating and maintenance personnel are required for operating the combined-cycle power plant and OSBL facilities, extra and spare operators, maintenance, and laboratory personnel making a total labor requirement of sixty people without management supervision.

The syncrude produced by this once-through design is nitrogen, sulfur and oxygen free. Because they are not finished gasoline and diesel blending components, the properties of the naphtha and distillate portions of the syncrude product were not estimated. However, the distillate portion should be similar to the previous distillates because it basically has undergone the same processing. The naphtha has a lower octane than the previous cases because it has not undergone any upgrading.

2.9 The ASPEN Plus Process Flowsheet Simulation Models

ASPEN Plus process flowsheet simulation (PFS) models for all the Fischer-Tropsch, indirect coal liquefaction cases were developed as part of this project. All of the model algorithms, computer code, and training necessary to operate and modify them were provided under this contract. The models were originally developed using ASPEN/SP. Documentation of the original ASPEN/SP models development is given in Topical Report, Volume IV, *ASPEN Flowsheet (PFS) Models* (reference TR-4). Documentation for the use of the various ASPEN PFS coal liquefaction models is given in the training session report, *ASPEN Process Flowsheet Simulation Model – Training Session Report* (reference TR-5).

All the models were later converted to ASPEN Plus code. Documentation of the various ASPEN Plus models is given in Topical Report, Volume V, *ASPEN PLUS Process Simulation Models for Fischer-Tropsch Indirect Coal Liquefaction* (reference TR-6). The use of these ASPEN models requires the ASPEN Plus process simulation program which is available from ASPEN Technology, Inc., Cambridge, MA.

These ASPEN Plus PFS models, along with the LOTUS spreadsheet economic model (discussed in Section 2.10), were developed as a tool with which DOE can use to plan, guide and evaluate its future research and commercialization programs. These models are not intended as detailed process design tools. They were developed based on the various coal liquefaction designs that were discussed in the previous sections of this report. The models are expandable and changeable on a block substitution basis. All contain an enhanced version of the detailed Fischer-Tropsch slurry reactor model that Bechtel developed under the previous Slurry Reactor Design Studies project (DOE-DE-AC22-89PC89867). Besides being responsive to a variety of processing conditions, this reactor model sizes the slurry bed reactors, calculates their weight, and estimates their ISBL field cost.

These models generate and report complete material and utility balances for all the individual plants, and for the entire coal liquefaction complex. Included are utility requirements and capital costs for all the major processing plants (ISBL plants). Major equipment lists and specifications outlines may be generated only for the Fischer-Tropsch synthesis plant. The outside battery limits (OSBL) plants, including the power generation plant, are modeled for utilities generation and operating expense calculations only.

The ASPEN Plus PFS model for the natural gas conversion case is documented in Topical Report VI, *Natural-Gas Fischer-Tropsch Case, Volume II, Plant Design and ASPEN Process Simulation model* (reference TR-9). This ASPEN Plus model is not a fully integrated model and it is not a deliverable to DOE as specified in the contract scope of work.

2.10 The LOTUS Spreadsheet Economics Model

A LOTUS 2.2 spreadsheet economics model was developed to analyze the economics of various coal liquefaction process scenarios using the results generated by the ASPEN Plus PFS model. This spreadsheet economics model does detailed discounted-cash-flow calculations using the flowrates, utilities, labor, and total capital information generated by the ASPEN Plus PFS model to allow the user to study the economic sensitivities of the economic and technical parameters discussed below.

The LOTUS economics model is a two-dimensional spreadsheet into which the user can import an ASCII file generated by the ASPEN Plus PFS model. The ASPEN model output file thus becomes an input for the LOTUS spreadsheet economics model, and along with other user controlled input parameters, drives the calculation of operating costs, capital costs, and revenues. These parameters are escalated as specified by user input economic parameters to generate a cash flow summary including the calculations of revenues, expenses, capital costs, depreciation, taxes, cash flows, internal rate of return, and net present value. The model also reports a Crude Oil Equivalent (COE) price. These results allow the user to perform manual iterations by varying the economic parameters, feedstock cost, product values or other items to achieve, for example, a required rate of return, COE, or to check the sensitivities of various parameters on the project economics. Highlights of the cash flow summary also are reported.

User supplied product values are entered in terms of a premium value for the naphtha and distillate blending stocks relative to crude oil. These premium values were determined by an linear program (LP) model study using an average PADD II refinery (see publication number 4 in Appendix B).

The LOTUS spreadsheet economics model is documented in Part 6 of Topical Report, IV, *Process Flowsheet (PFS) Models, LOTUS Spreadsheet Economics Model (Restricted Addendum)* (reference TR-4, part 6). This report provides the only documentation available for the spreadsheet economics model. Users should be knowledgeable in process economics, particularly in F-T technology and process economics.