INTEGRATED COAL TO TRANSPORTATION LIQUIDS STUDY

by

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Presented at the
Indirect Liquefaction Contractors' Review Meeting
Pittsburgh Energy Technology Center
Pittsburgh, PA

November 15, 1989

1.0 INTRODUCTION

In early 1989 the Department of Energy Office of Coal Conversion Systems directed Gilbert/Commonwealth (G/C) to investigate the integration of various processes to produce transportation liquids from coal. What began as a broad survey type of initiative quickly evolved into a study to integrate novel concepts and processes to produce predominantly diesel fuel and gasoline via indirect liquefaction of coal, i.e., Fischer-Tropsch synthesis. The purpose was to determine if a pattern could be established to direct Fischer-Tropsch (F-T) liquids production toward economic viability. Accordingly there were few constraints placed on the process selection portion of the study.

The overall scope of the study consisted of sizing a plant at mine mouth with a throughput of 30,000 tons per day of dry coal. The coal would be gasified to produce synthesis gas (syngas) for the F-T reactor. The scope departs from conventional F-T synthesis in that the product gases are not recycled to extinction. This particular scope required G/C to explore alternative approaches to utilizing the off-gas and thereby studying a once-through Fischer-Tropsch concept. The scope also required that the plant produce all liquid products and useful byproducts, e.g., cogenerated electricity.

Projects of this type are unique in that the end results are seldom as predictable as straightforward process plant design efforts. The results of this project can be best described as serendipitous i.e., discoveries which manifested themselves in the selection of a molten carbonate fuel cell (MCFC) for power production. By combining the MCFC with the once-through slurry Fischer-Tropsch (SFT) reactor, a synergism was developed that used the available product gas effectively and increased the level of power production significantly.

By conducting a thorough process selection phase at the beginning of the project, it was possible to enter the plant design phase without an iterative step. Being conceptual in scope, the plant costing and economic analysis include contingencies. Sufficient sensitivities are present to provide indicators of the need to conduct research in selected design and process areas.

2.0 TECHNICAL APPROACH

The production of diesel fuel and gasoline by indirect liquefaction of coal requires the integration of gasification, cleanup, liquefaction, and refining processes. In the concept studies here, the Fischer-Tropsch reactor is operated in a once-through mode that has a tail gas consisting of unreacted H_2 and CO along with CO_2, C_1, C_2 and other non-concondensed vapors. We considered the option to either burn them in a combustion turbine/combined cycle, or to feed them to a fuel cell to produce electricity. In this paper, we discuss the fuel cell option. Figure 1 is a block diagram of the plant concept.

2.1 Background

The feed coal type assumed for transport liquids production via a coal gasification/F-T synthesis route is a generic Illinois No.5 bituminous. A comparative analysis conducted early in the project indicated that high temperature cocurrent-flow, oxygen-blown entrained gasifiers such as the Shell or Texaco gasifiers were most appropriate when assessed against other commercial gasifiers or advanced gasifiers currently under development. Either Shell or Texaco can produce a syngas containing a very low fraction of methane at high thermal conversion efficiencies which is a prerequisite for all-liquids production. The Shell gasification process was selected as preferable based on its relative merits such as higher (H_2+CO) yields per unit weight of coal (on a moisture-free basis), minimum light hydrocarbons in the raw gas, higher carbon conversion potentials, good coal flexibility and other desirable design/operational characteristics. In conjunction with gasifier selection analysis, it was also determined that, for the purposes of removing sulfur and other detrimental species from the syngas, a conventional cold gas cleanup is preferable to other options such as high-temperature desulfurization and physical/chemical pre-cleaning of coal

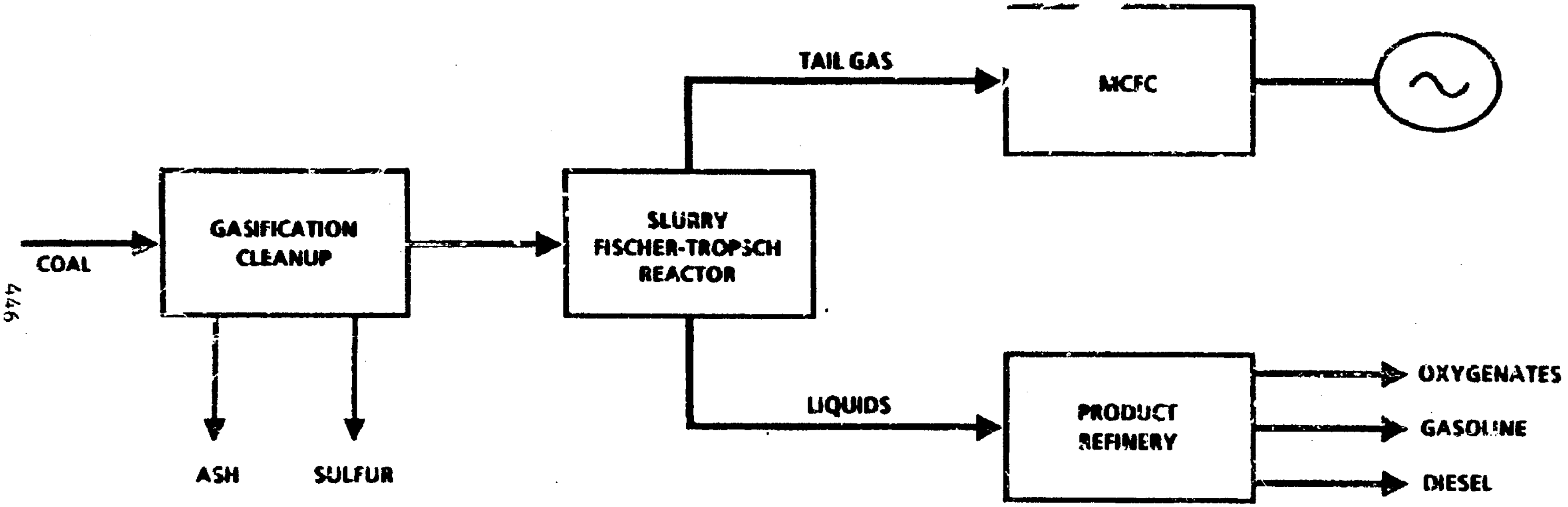


FIGURE 1
BASIC PLANT CONCEPT

Coal-derived syngas is, by nature, rich in CO and relatively deficient in H_2. Typically, the H_2/CO ratio in the coal-based syngas is about 0.5 and is best processed in a slurry-phase Fischer-Tropsch reactor of the type first demonstrated by Kolbel. A very desirable feature of the Kolbel type slurry reactor is that it is simpler in design and is capable of removing a greater quantity of exothermic heat of reaction than other reactor types. Furthermore, as shown in the Mobil slurry F-T pilot data (1, 2), its single pass conversion can be very high and, hence, it is most suited for applications when the F-T plant is configured for an once-through mode of operation. The conventional Arge fixed bed and/or Synthol circulating fluid bed reactors commercialized at the SASOL plants are deemed less applicable to handle the Shell coal gas, which has a H_2/CO ratio as low as 0.426. Moreover, the slurry F-T reactors tends to reject its major waste byproduct in the form of CO_2 (rather than H_2O vapor as is the case with Arge or Synthol when fed with a syngas having a H_2/CO ratio of about 2). The CO_2 so generated in the slurry reactors can be separated and recovered for an interesting synergistic power generation application downstream of the SFT area (see below).

The slurry F-T reactor can be run in either low-wax or high-wax modes of operation (1, 2). When the SFT plant is operated in an once-through mode for all-liquid production as in the present study, the high-wax mode is clearly the preferred choice. In the high-wax mode of operation, the reactor wax yield (C_{26}^+) increases substantially to 40-60 wt.% while the undesirable C_1C_2 yield is reduced to 2-6 wt.%. The yield of condensable oxygenates that are coproduced is also low at about 5 wt.%. The reactor wax, along with other middle fractions, can be further processed and upgraded in a downstream refinery to maximize the overall production of gasoline/diesel. In this study we selected a refining scheme similar to an UOP study (3) which includes a proprietary wax hydrocracking scheme and other more conventional fractionation, hydrotreating, and isomerization processes.

The Mobil pilot high-wax data is fitted into a modified Schulz-Flory equation so that the distribution of raw products could be described in a compact parametric format. The relative distribution so determined is illustrated in Figure 2 for oxygenates (5.05 wt.%), waxy fraction (F2, 49.24 wt.%), and the balance of hydrocarbons (F1, 45.71 wt.%). The raw gas exiting the once-through slurry F-T reactors will also contain about 15 wt.% of unconverted ($CO+H_2$) in the feed. After passing through various separation units for C_3^+ recovery and separation of the bulk of CO_2, the tail gas containing unreacted ($CO+H_2$) plus non-condensable C_1C_2 still carries a sizable fraction of the energy in the coal. This gaseous stream could be utilized in a combined-cycle power plant for power generation at about 36-39 % efficiency. In the present study, however, this tail gas along with the recovered CO_2 stream is more effectively and synergistically utilized in a molten carbonate fuel cell (MCFC) power plant to generate electric power electrochemically at more than 50% conversion efficiency.

The MCFC plant converts chemical energy directly to electricity The principle of MCFC operation is that, when fuel gas containing ($CO+H_2$) is inputted to the anode side of the MCFC, the following reactions occur:

$$H_2 + CO_3^{=} \dashrightarrow H_2O + CO_2 + 2e^-$$

$$CO + H_2O \dashrightarrow CO_2 + H_2$$

The cathode of the MCFC requires both CO_2 and O_2 (from the air) for the following reaction to occur:

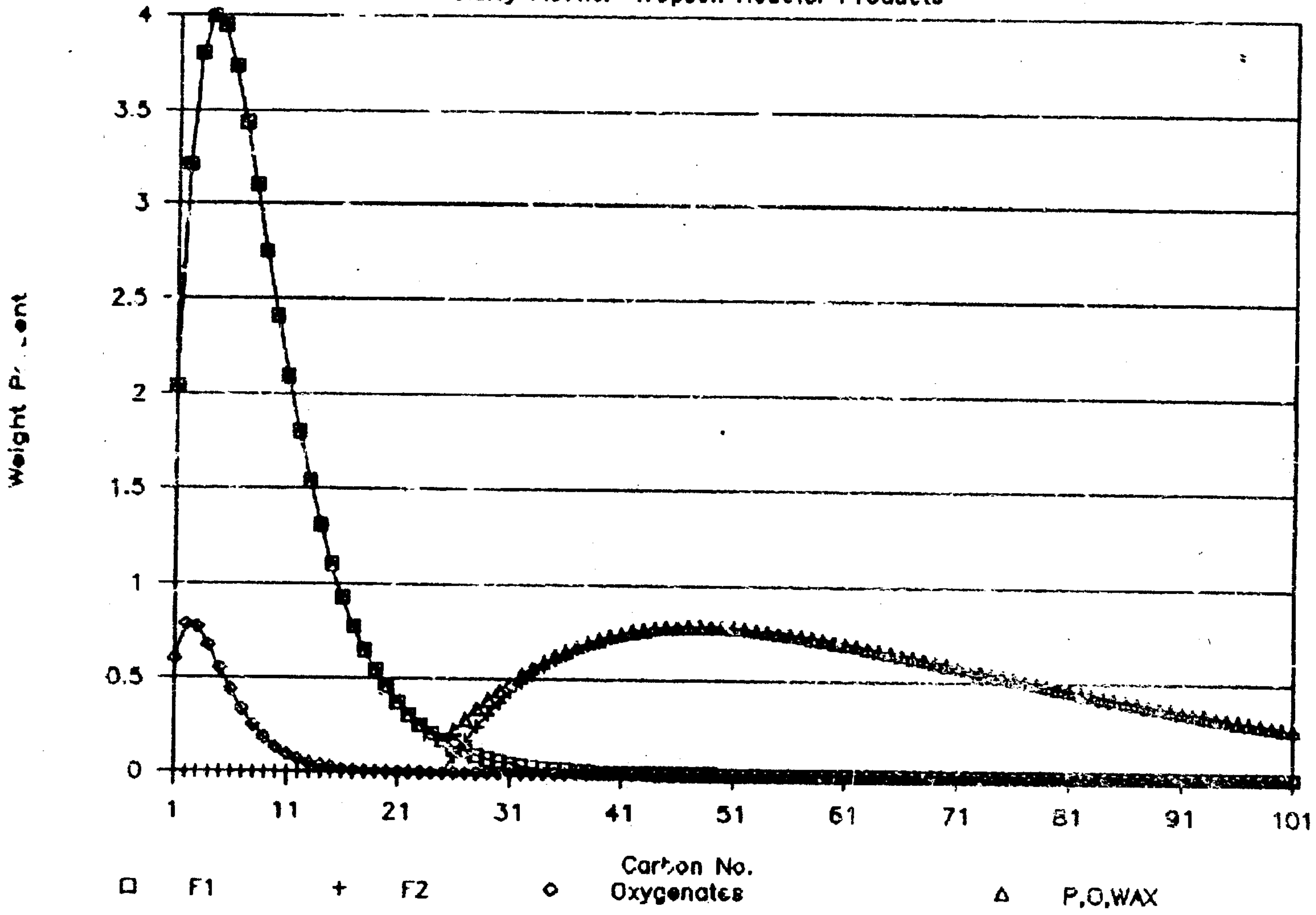
$$0.5\ O_2 + CO_2 + 2\ e^- \dashrightarrow CO_3^{=}$$

In essence, CO_2 inputted to the cathode side carries the necessary oxygen in an ionized form which is transferred to the anode side for the anode "combustion" reactions to proceed. Thus, the "waste byproduct" CO_2 from the SFT area is utilized as the necessary "input" in the MCFC plant for further conversion of the unreacted chemicals The

FIGURE 2

MODIFIED SHULTZ-FLORY DISTRIBUTION

Slurry Fischer-Tropsch Reactor Products



electricity thus generated is in the form of direct current which, however, can be converted readily to AC in an inverter at about 96% efficiency.

The MCFC technology is still in an early stage of development but it is expected to operate at about 1,100-1,300 F and 150+ psia. Its projected performance characteristics (the voltage vs. DC current density as a function of fuel utilization) can be estimated from various references (4, 5) as shown in Figure 3. The operating point selected for present study is: $V = 0.799$ Volt, $i = 220\ mA/cm^2$, and $U = 85\%$ fuel ($CO+H_2$) utilization at $P = 180$ psia and $T = 1,200$ F. The final exhaust gas from the MCFC anode has very low heating value of only about 55 Btu/scf and is best disposed of in a furnace as fuel gas.

In summary, the syngas produced in the gasification plant is utilized in a form of grand cascade: first as the feedstock in the SFT plant for gasoline/diesel production, the unreacted fraction of which is then processed in the MCFC plant for electro-chemical power generation followed by expansion of the anode/cathode exhaust gases in turbo-machinaries for mechanical work recovery and, finally, combustion of the remaining poor quality fuel gas (along with other waste fuel gases generated in the entire complex) in a furnace in support of a central Rankin-cyle steam power plant for additional electric power generation. In this way, the energy in the coal is best utilized and hence make all the technologies involved more economically feasible.

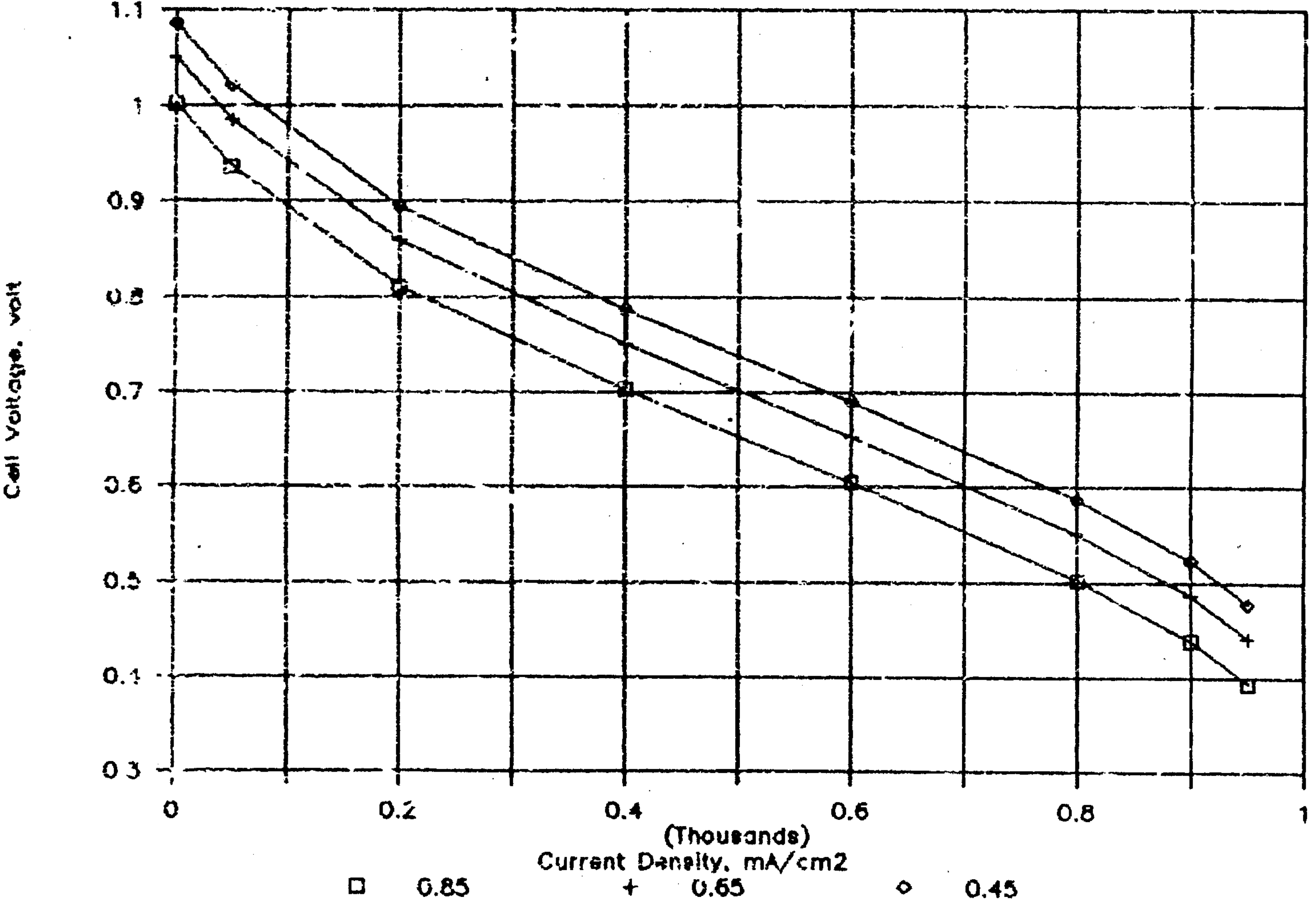
2.2 Plant Design

A baseline conceptual design was developed based on criteria established by the DOE and on the results of the process screening studies with the objective to maximize both liquids and electric power. The plant consumes 30,000 tons per day of dry Illinois No. 5 coal which is gasified in the Shell dry feed entrained flow gasifier. Gas cleanup is conventional and removes sulfur to the low level required of the Slurry Fischer-Tropsch (SFT) catalyst. The SFT operates in a high wax mode without recycle and produces a broad spectrum of products which are

FIGURE 3

MCFC PERFORMANCE CHARACTERISTICS

(Parameter = Fuel Util Fraction)



further refined to produce diesel fuel and gasoline. The tail gases containing H_2, CO, and CO_2 are reacted in a molten carbonate fuel cell to produce about one third of the electric power. Steam generated from waste heat throughout the plant is used to produce the balance of electric power.

Whether the plant is built as a single entity or is designed in multiple trains at different locations, the magnitude still requires that the site be located in proximity to the coal mine. It is assumed that the mine mouth site will have adequate water supply, and provisions for landfilling of ash from the gasifier.

PLANT DESCRIPTION

This section describes the baseline plant, designed in a conceptual scale. Figure 4 is a process block flow diagram of the plant, and Table 1 presents the overall performance summary for the plant.

The following process flow diagrams have been developed for this plant design:

Figure 5	Coal Gasification PFD
Figure 6	Gas Cleanup PFD
Figure 7	Fischer Tropsch Reactor PFD
Figure 8	Molten Carbonate Fuel Cell PFD
Figure 9	Fischer Tropsch Product Refinery PFD

Figure 10 is an overall plant steam balance, and Table 2 presents an energy balance for the plant.

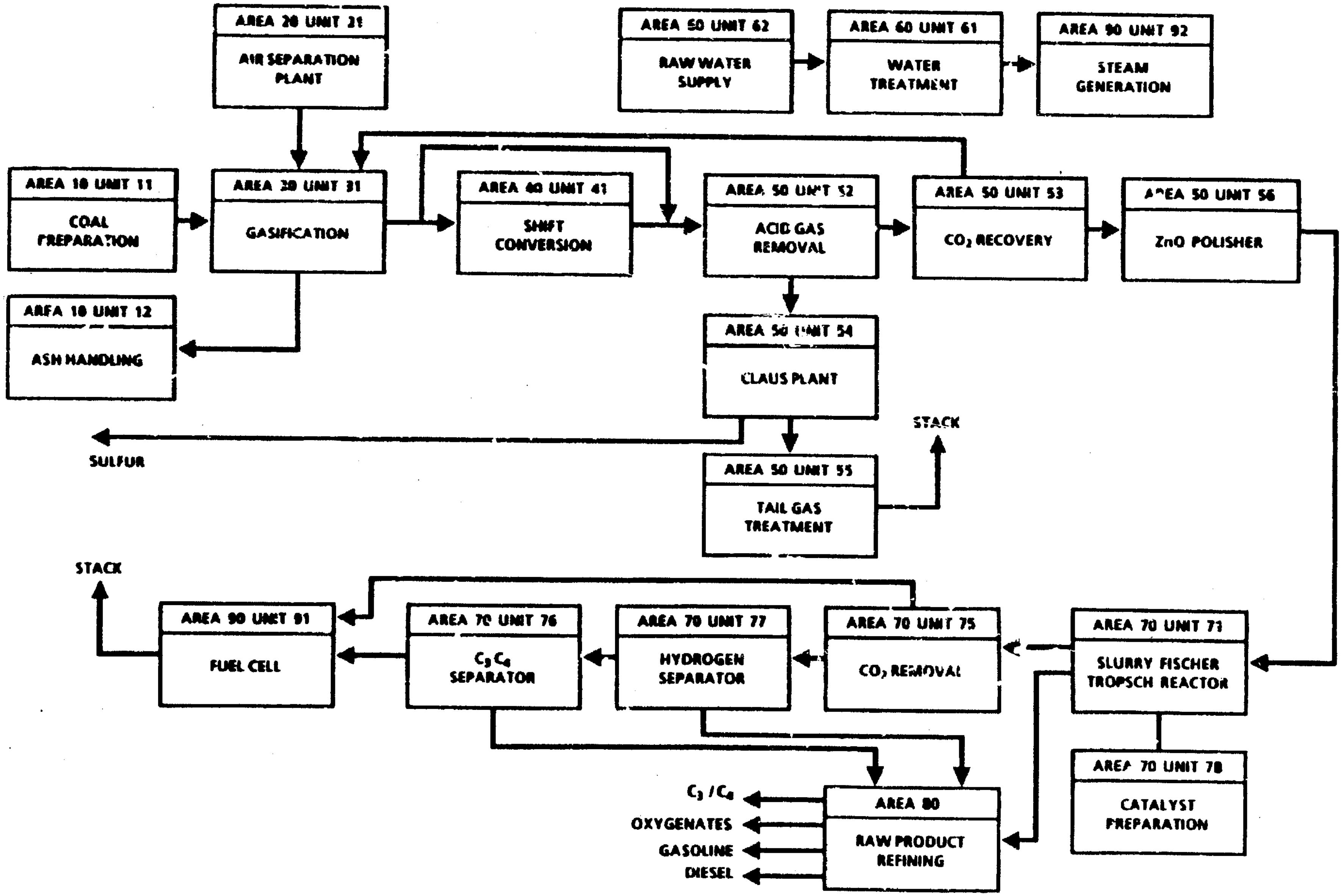


FIGURE 4
OVERALL PLANT BLOCK FLOW DIAGRAM

TABLE 1

Performance Summary

MATERIAL		
Coal Feed Rate (Dry)	2,500,000	lb/hr
Oxidant Feed (99.8% Oxygen)	2,000,305	lb/hr
Clean Gas to SFT	4,222,820	lb/hr
Sulfur Recovery	71,683	lb/hr
Slag/Char Production (dry)	400,254	lb/hr
POWER PRODUCTION		
Gross Steam Turbine Power	1,101,802	kw
Net Steam Turbine Power	1,055,526	kw
Gross Fuel Cell Power	611,871	kw
Net Fuel Cell Power	492,448	kw
Process Plant Power Needs	458,464	kw
Net Power Production	1,089,510	kw
LIQUID PRODUCTS		
LPG Production	0	lb/hr
Oxygenates Production	40,311	lb/hr
Gasoline Production	239,564	lb/hr
Diesel Production	380,613	lb/hr
OVERALL PLANT EFFICIENCY (HHV)	56.43	percent

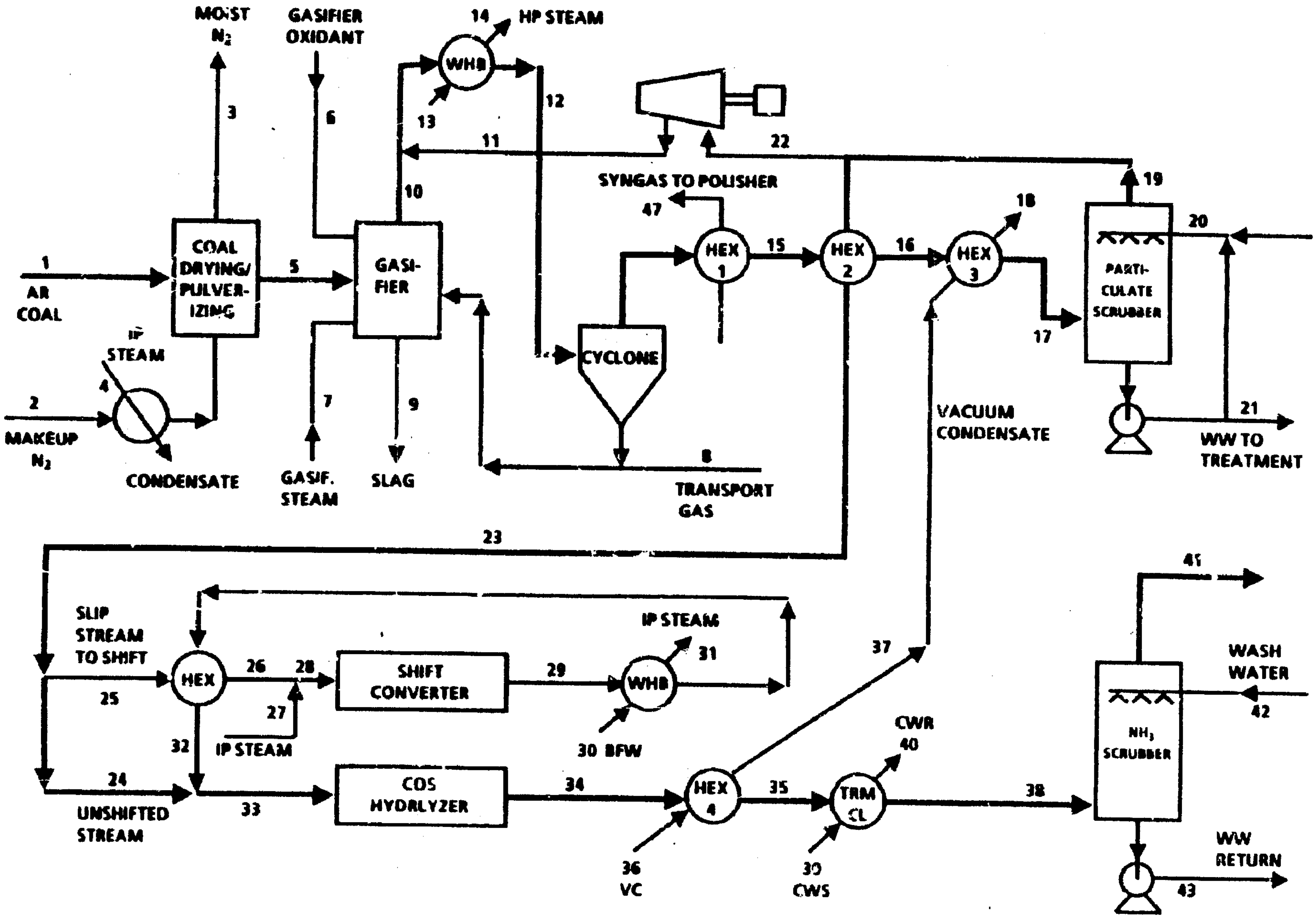


FIGURE 5
GASIFICATION PROCESS FLOW DIAGRAM

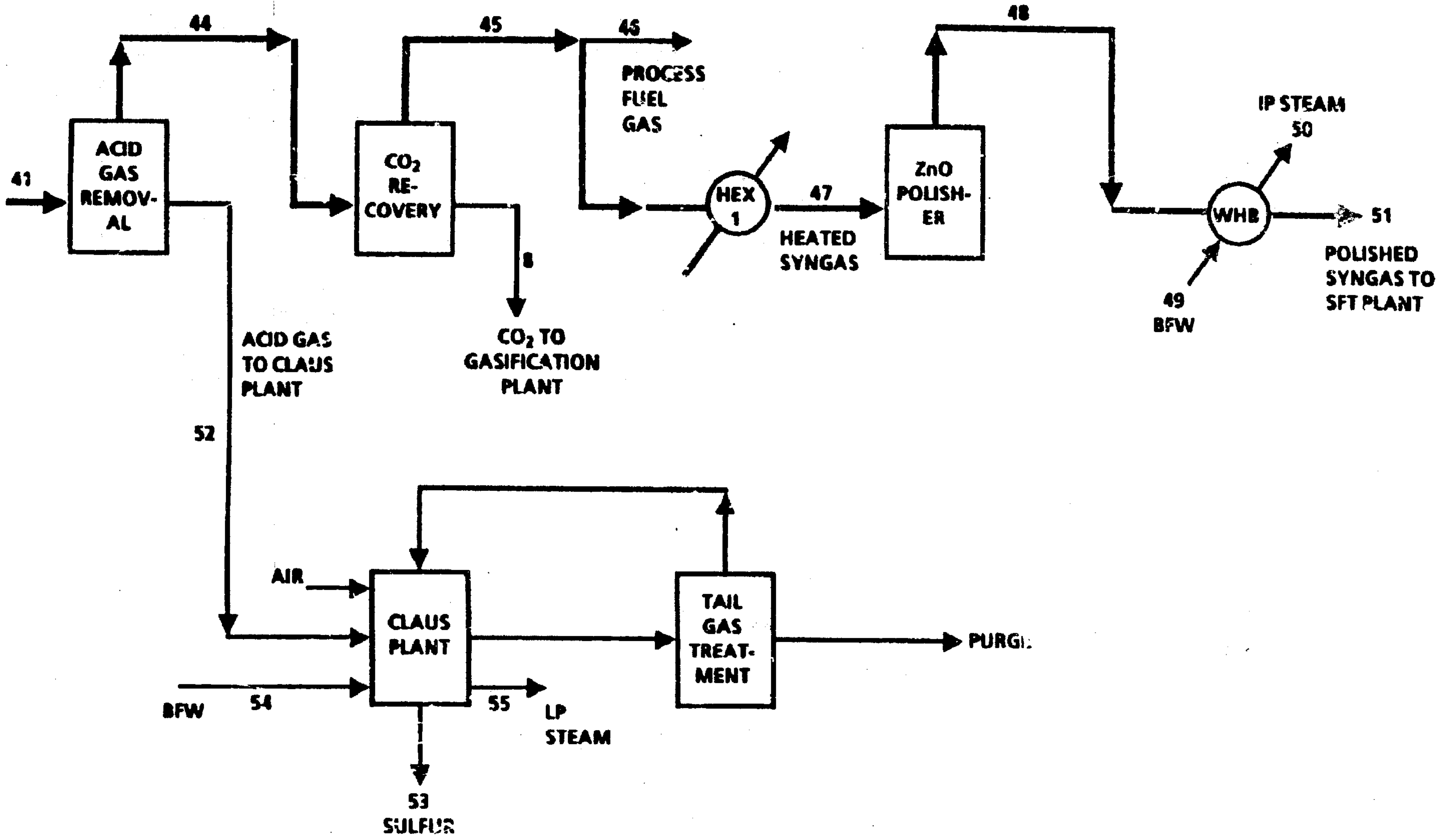


FIGURE 6
GAS CLEANUP PROCESS FLOW DIAGRAM

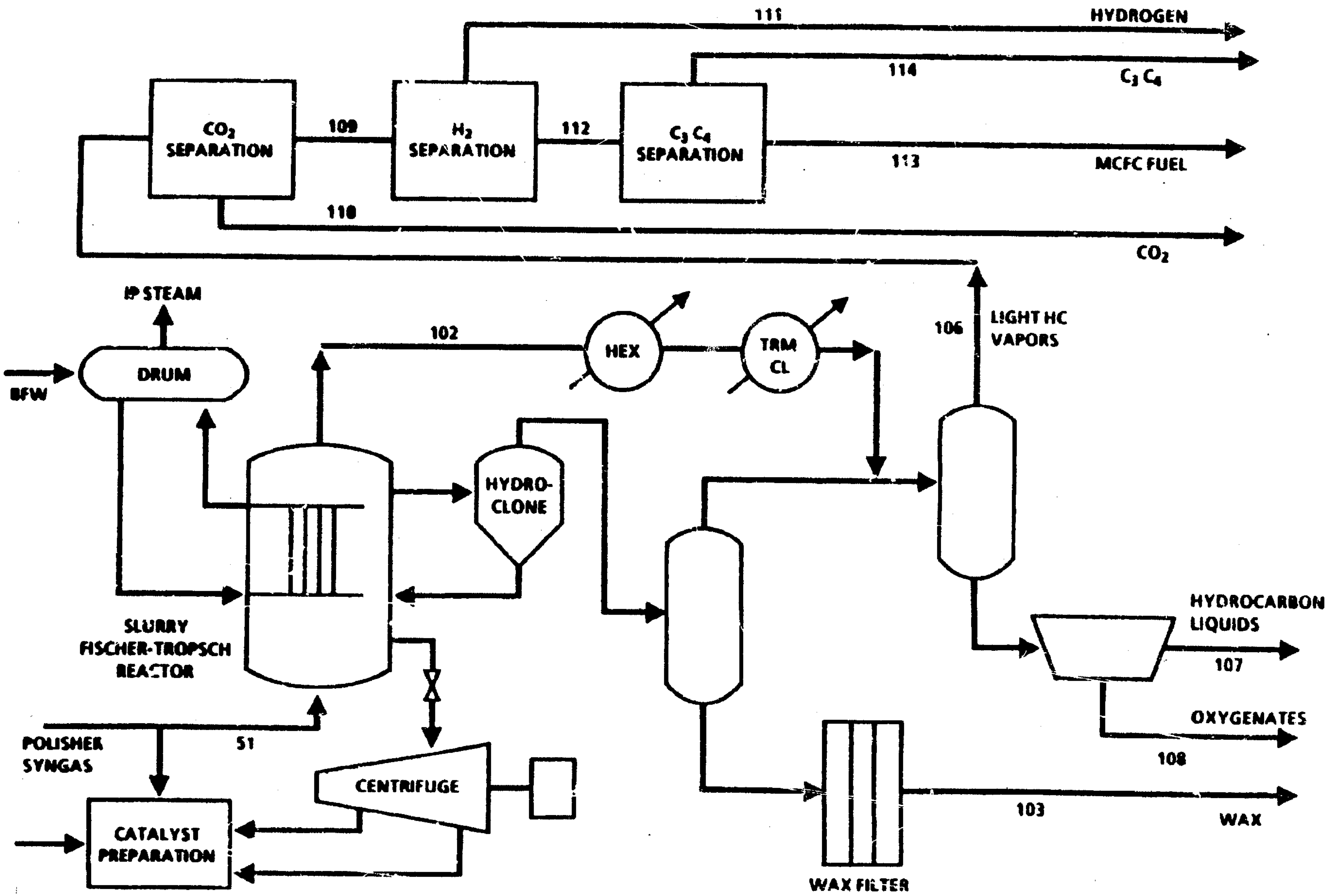


FIGURE 7
FISCHER-TROPSCH REACTOR PROCESS FLOW DIAGRAM

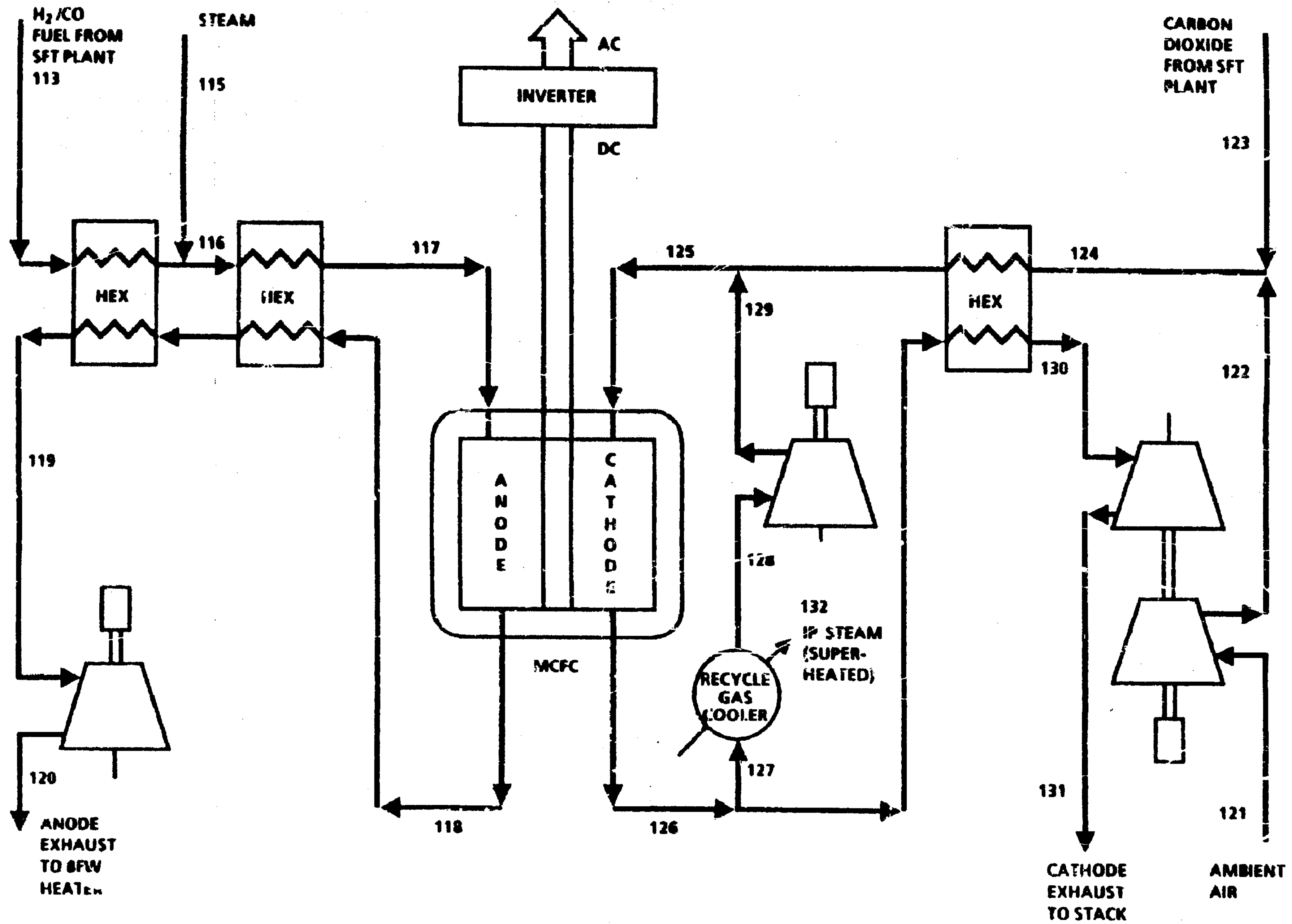


FIGURE 8
MOLTEN CARBONATE FUEL CELL PROCESS FLOW DIAGRAM

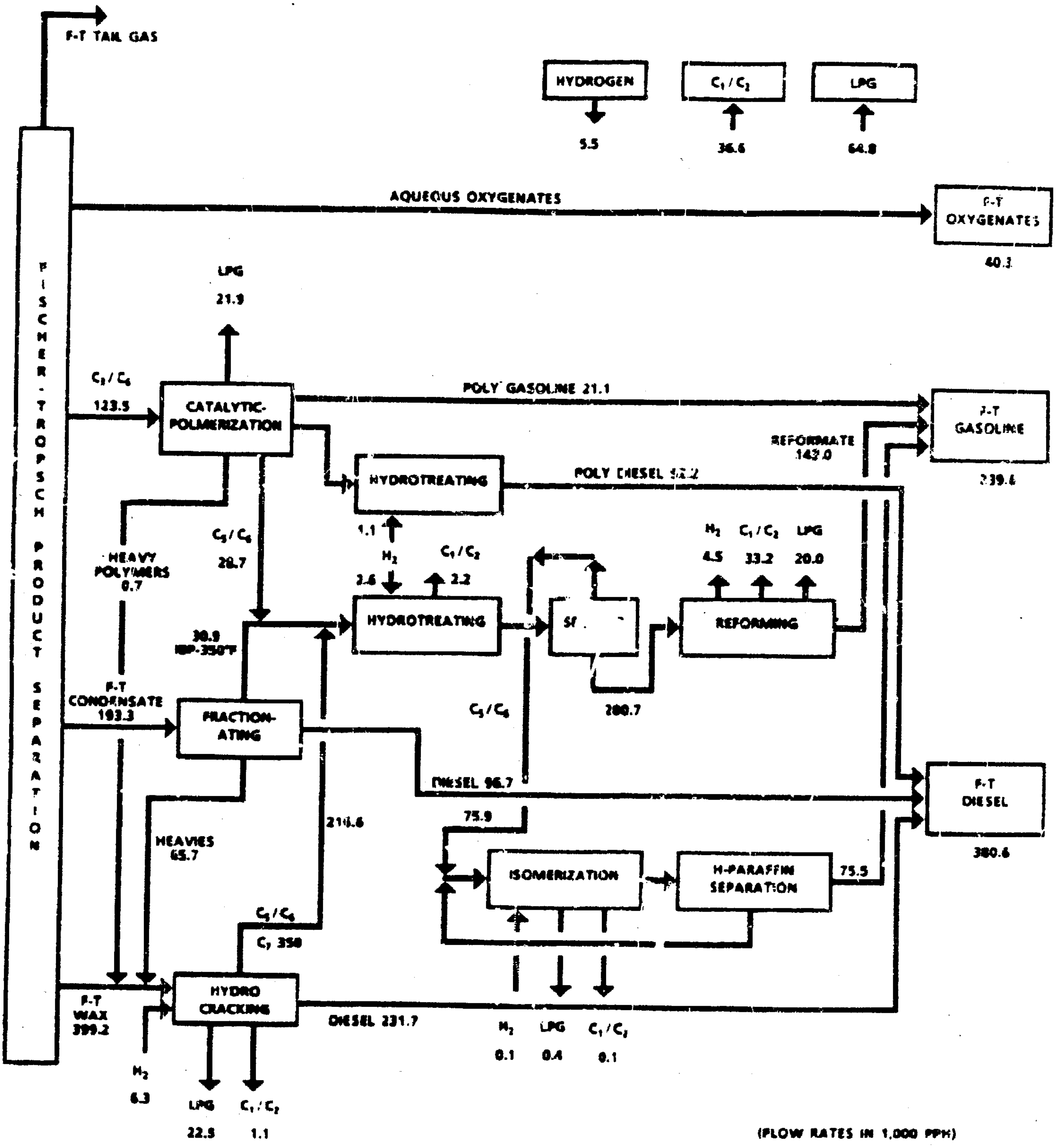


FIGURE 9
SFT PRODUCT UPGRADING AND REFINING
PROCESS FLOW DIAGRAM

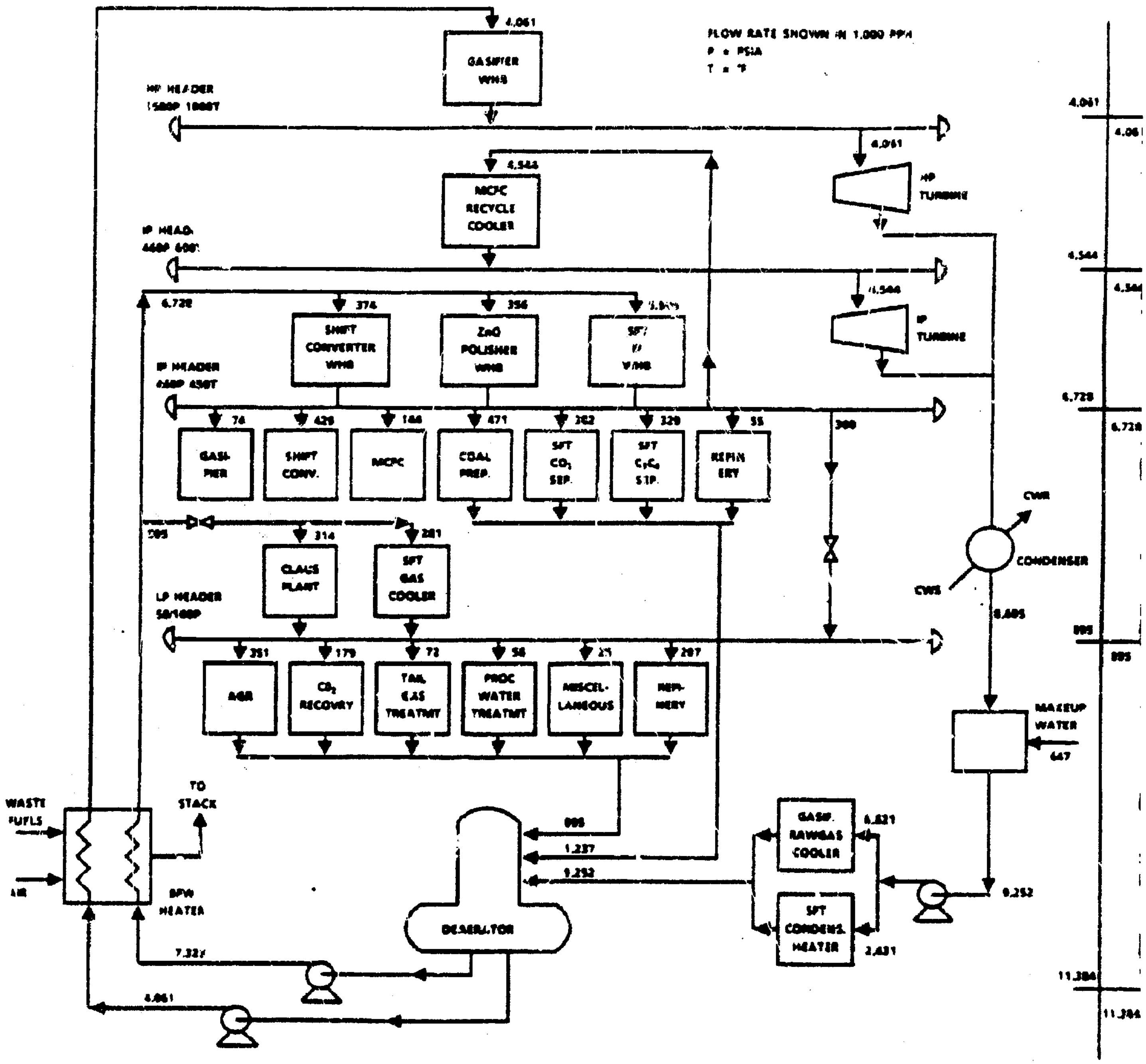


FIGURE 10
OVERALL PLANT STEAM BALANCE

TABLE 2

Shell Coal Gasification/SFT/Refinery Plant
Overall Thermal Efficiency

	pph	MWe	10^9 Btu/h HHV	Btu, %
INPUT				
Coal (dry)	2,500,000		30.95	100.00
OUTPUT				
Transportation Liquids				
Diesel	380,613		7.726	24.95
Gasoline	239,564		4.943	15.96
Oxygenates	40,311		0.581	1.88
LPG (C_3C_4)	0		0.00	0.00
Total Liquids	660,488		13.250	42.79
Sulfur (By-product)	71,683		0.509	1.64
Electricity				
Steam Plant Gross		1,102		
Net		1,056		
MCFC Plant Gross		612		
Net		492		
Process Plant Needs		458		
Overall Power Systems		1,090	3.718	12.00
TOTAL, OUTPUT	732,171	1,090	17.477	56.43

Coal Gasification and Gas Cleanup

33,304 tons per day of Illinois No. 5 coal are received and processed on site. Coal is transported directly to the gasification plant by conveyors and stored in a thirty day surge pile. To be used by the Shell gasifier, coal is pulverized and dried under a nitrogen blanket to 80%-200 mesh and 5% moisture. 460 psia steam provides the heat for drying.

The Shell gasifier is a pressurized dry feed process operating at 2,550°F and 400 psia. Dry pulverized coal mixed with steam and oxygen is fed into the gasifier with a screw feeder. Above the combustion zone, the gas is quenched with recycled fuel gas to solidify the entrained fly ash particles. Slag produced during gasification falls to the gasifier bottom, and is water quenched and discharged for landfill disposal. The hot product gas passes through a waste heat boiler in which 1,600 psia steam is produced. Upon exiting the boiler at 750°F, the gas passes through a hot cyclone. Flyash containing unreacted carbon from the cyclone is returned to the gasifier. The gas goes through a series of heat exchangers and is finally cooled in a wet particulate scrubber. A portion of the gas exiting the scrubber is recompressed and recycled to the gasifier as a quench for the hot raw gas. The remainder of the scrubbed gas is reheated to 334°F in preparation for CO Shift and COS hydrolysis.

A slip stream is taken from the fuel gas for CO shift conversion. Using a converter exit temperature of 862°F, sufficient shift reaction occurs to result in a H_2/CO ratio of 1.139. Upon remixing this stream with the main gas stream, the resulting H_2/CO ratio is 0.67 which is the design requirement for the Slurry Fischer-Tropsch (SFT) synthesis gas. The mixed gas stream at 375°F passes through a COS hydrolyzer bed in which 95 percent of the COS is converted to H_2S. The gas is then cooled to 108°F, and passes through an ammonia scrubber before entering the acid gas removal process. A Selexol process is used to remove H_2S and some CO_2 from the gas.

99.9 percent of the H_2S is removed from the gas stream, leaving a sulfur concentration of 10 ppm. Acid gas produced by regenerating the Selexol solution has a higher heating value of 199 Btu/scf which is adequate for Claus plant combustion. 860 tons per day of sulfur are produced by the Claus plant, and a SCOT plant treats the Claus tail gas. Additional CO_2 is recovered from the main synthesis gas stream for use as a carrier gas in the gasification plant. At this point, 0.5 percent of the synthesis gas is extracted for all fuel gas use throughout the plant. The clean synthesis gas is heated to 690°F and passes through a fixed bed of ZnO which polishes the gas to essentially a zero sulfur content. Upon leaving the polisher, the gas is cooled to 500°F at 250 psia for entry to the SFT reactor.

Fischer-Tropsch Synthesis

The SFT operates in a once-through mode to produce a combination of raw products. These are then separated into diesel fuel and gasoline fractions along with oxygenerated hydrocarbon liquids and liquified petroleum gas. The remaining tail gas is used as fuel for a molten carbon fuel cell to generate power for export.

The clean synthesis gas passes through the bottom of the SFT reactor and bubbles upward through the slurry catalyst. Under these conditions, 85 percent of the H_2 and CO is reacted in a single pass to produce a distribution of hydrocarbon and oxygenate products. Temperature of the exothermic reactor is controlled by in-bed boiler tubes producing nearly 6 million pounds per hour of intermediate pressure steam.

The wax product produced in the SFT reactor is continuously withdrawn to maintain the slurry level. The catalyst is separated from the slurry in a hydroclone, and returned to the SFT. The wax product is filtered through a vertical leaf filter, and fed to the products refinery. The remaining product of the SFT is in vapor form and leaves the top of the SFT reactor at 500°F. Low pressure steam is generated from a waste heat boiler, and the vapor stream is trim-cooled to 100°F. Hydrocarbons above C_4 and oxygenates are condensed and separated from the vapor. The liquids are further gravity separated into C_5^+ hydrocarbons and oxygenates and are sent to the products refinery.

The remaining non-condensable vapors go through a series of processes which adjust the gas composition for use as anode feed to the molten carbonate fuel cell (MCFC). The vapors initially pass through a CO_2 separation process in which 90 percent of the CO_2, because of its high partial pressure, is absorbed from the gas phase.

The rich absorber solvent is heated to 325°F under pressure to drive off the CO_2 at 300°F and 190 psia. The regenerated lean solvent is then cooled to 110°F and circulated back to the absorption unit. The CO_2 goes to the MCFC cathode inlet. The CO_2-lean gas passes to a hydrogen separation process. This is a membrane separation which removes a quanity of the hydrogen from the fuel gas stream to meet the hydrogen requirements of the product refinery. After hydrogen separation, the fuel gas is sent to a hydrocarbon absorber to remove the C3/C4 LPG. This is accomplished by an oil absorbent which is then regenerated in a steam-heated stabilizer. The regenerated oil is then cooled to 140°F for recirculation back to the absorber unit. The remaining fuel gas then passes to the MCFC anode inlet.

Power Generation

There are two sources of electric power generation for this plant. About one third of the total power is generated by a molten carbonate fuel cell (MCFC) that uses the gaseous products from the Fischer-Tropsch reactor. The balance of power is from a combination of expansion turbine and steam turbines which utilize available sources throughout the plant.

The product gas stream from the SFT reactor is preheated by the anode exhaust to 500°F whereby stoichiometric steam is added to prevent carbon deposition and also to promote the CO shift reaction. The gas is then further heated to 1,100°F for input to the anode. The anode utilizes 85% of the fuel gas (H_2 + CO) in a single pass. The exhaust gas at 1,200°F passes through the preheater and through an expansion whereby pressure is reduced from 174 to 16 psia, and temperature from 757°F to 355°F. The gas from the expander exhaust goes to the BFW heater for ultimate combustion.

The bulk of heat produced by the MCFC is removed by the cathode flowing gas. The cathode oxygen supply is provided by a turbine air compressor mixed with the CO_2 extracted from the SFT product gas at 187 psia. The cathode feed gas is then pre-heated to 1,100°F. Since the cathode gas supply is insufficient to cool the MCFC, a cathode gas recycle loop is necessary. Gas leaving the cathode at 1,300°F is split in a ratio of 3.4 to 1. The higher flow passes through an intermediate pressure superheater and is recompressed for mixing with the incoming CO_2/air stream. The balance of the cathode exhaust passes through the pre-heater in which it is cooled to 528°F. It then goes through a turboexpander and exhausts to the atmosphere at 140°F.

Fischer-Tropsch Product Upgrading and Refining

The flow scheme was developed using both the experience gained at SASOL and the results of a recent UOP study (3) of upgrading F-T products. A hydrocracker is used to upgrade Fischer-Tropsch wax, heavy condensate and heavy polymers into transportation fuels.

The F-T condensate light fraction is hydrotreated along with C_5/C_6 from the catalytic polymerization unit and the lights from the hydrocracker. The primary purpose of the hydrotreater is to saturate the olefins and to remove any oxygenates which may be present. The hydrotreated stream is split into two fractions. The C_5/C_6 fraction is used as an isomerization feedstock and the remainder is reformed. The F-T condensate heavy fraction is sent to the hydrocracker; the middle fraction is sent to diesel blending.

The light olefins (C_3/C_6) in the Fischer-Tropsch products are catalytically polymerized into liquid products. The poly-gasoline is sent to gasoline blending. The poly-diesel is hydrotreated to improve the cetane number and sent to diesel blending. The oxygenates are recovered from the F-T aqueous stream primarily by distillation. Saleable products such as ethanol, propanol, acetone and MEK can be reclaimed. The net hydrogen needed for the refinery was recovered from the F-T tail gas. Excess fuel gas and LPG's produced in the refinery and not used internally was used for fuel in the power generation areas to produce additional electricity.

3.0 ECONOMIC ANALYSIS

Analysis of the coal to transportation fuels consisted of evaluating the various costs associated with constructing and operating the complete facility. This encompasses all of the major systems beginning with the receipt of coal through the production and storage of the product liquid fuels and production of electric power. Both of the areas of cost that contribute to determination of the annual revenue requirement were examined under separate headings. These areas consist of the facility capital cost and operation & maintenance costs, including the cost of coal fuel. An additional area included in the analysis section is cost of product. This area examined the results of the cost evaluations and expresses those results in terms of the price of liquid fuels and the associated price for generated electric power that are required to match the annual revenue requirement.

The capital cost of the complete facility was developed through a process of separately evaluating the capital cost of each of the major systems that contribute to the production of the liquid fuel and electric power products. A combination of referenced reports and in-house Gilbert data were factored to both size and 1989 dollar adjustments to obtain base erected costs. In cases where the process units employ multiple trains of equipment, relatively high scaling exponents, i.e. 0.85, were utilized. In cases where the process units are significantly less modular, exponents on the order of 0.7 were utilized. Additional adjustments were made under guidelines of EPRI TAG (P-4463-SR).

As part of the capital costs, both a project and process contingency were included. The project contingency was evaluated at a consistent rate of 15 percent. This contingency was included to address the project uncertainty and the cost of any additional equipment that could result from a more detailed design. Some of the process technology used in the various systems is still in the development stage. In an effort to address the uncertainty associated with the continuing process of development in terms of equipment design, performance and cost, the process contingency was incorporated. In this case the process contingency factors were established for each appropriate plant section. The basis for determining the factors was the criteria established in the current EPRI TAG (P-4463-SR).

Combining the first year operating and maintenance costs with the consumables, by-product credit for sulfur and fuel cost along with the carrying charge resulted in the first year revenue requirement. For the coal to liquid fuel facility, this requirement was evaluated at $1.216 billion.

The principal objective of the evaluation was to determine the first year value of the liquid fuel product. In order to determine this value, the annual (first year) revenue requirement was first established. In addition, the value for the other facility products was determined in order to recognize their contribution to the annual revenue. In the case of the generated electric power, the value is dependent on the rate per kilowatt-hour that can be expected. For the base case evaluation, a rate of 5 cents per kilowatt hour was selected.

Based on the established rate of 5 cents per kilowatt-hour for generated power of 1090 MW, along with the daily rate of 318 billion Btu's of liquid fuel product, the value of the liquid fuels necessary to meet the revenue requirement is $7.53 per million Btu based on a 90 percent capacity factor. Table 3 contains a consolidated summary of the revenue requirements that is summed and identified as the TOTAL ANNUAL COST. The table also identifies the revenue basis and identifies the unit value for generated power expressed in cents/kWh and liquid fuels expressed in $\$/10^6$Btu and $/barrel. In addition, the table identifies the percent of first year revenue realized by each of the major product streams based on the indicated unit values.

Since the facility, as configured, does yield a significant revenue from the generation of electric power and since the unit value selected for that power is a strong influence on the unit value of liquid fuels, a sensitivity to the unit value of power was performed. The cents per kWh values in the sensitivity were selected to encompass the range of rates currently being realized by cogenerating facilities and rates paid by industrial customers. Figure 11, Revenue Required Selling Prices, identifies the unit value of both liquid fuels and electricity expressed in $\$/10^6$Btu. The successive pairs of values indicate the relationship to the power revenue at the various values in cents/kWh.

4.0 SIGNIFICANT ACCOMPLISHMENTS

The scope of this task provided for a thorough screening study to determine the most appropriate processes to use in the production of liquids and power from Fischer-Tropsch synthesis. One significant result of the screening effort was the utilization of the syngas in a cascading fashion. The syngas is first utilized as the feedstock in the SFT plant for gasoline/diesel production. The unreacted syngas fraction is then processed in the MCFC plant for electro-chemical power generation followed by expansion of the anode/cathode exhaust gases for mechanical work recovery and, finally, combustion of the remaining poor quality fuel gas in support of a central Rankin-cycle steam power plant for additional electric power generation. In this way, the energy in

TABLE 3

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                    CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY

 TITLE/DEFINITION
   Case:                                   CG/MCFC/Liquid Fuels Baseline
   Plant Size: Electric/Liquid Fuels         1089.8 (MW,net)  Barrel/day:     57,939
   Fuel(type):                             Illinois #5        Cost:              1.02 ($/MMBtu)
   Design/Construction:                           5 (years)   BookLife:            30 (years)
   TPC(Plant Cost) Year:                       1989 (Dec.)    TPI Year:          1990 (Jan.)
   Capacity Factor:                              90 (%)
                                                               $x1000
 CAPITAL INVESTMENT                                          ---------
   Process Capital & Facilities                              3,063,300
   Engineering(incl.C.M.,H.O.& Fee)                            306,330
   Process Contingency                                         371,460
   Project (  tingency                                         561,164
                                                            -----------
        TO'  . PLANT COST(TPC)                              $4,302,254
        TO  L CASH EXPENDED(Mixed Year $)        $3,888,948
        AF(                                        $911,431
           L PLANT INVESTMENT(TPI)                          $4,800,379

   Royalty Allowance                                            15,317
   Preproduction Costs                                         115,097
   Inventory Capital                                            39,406
   Initial Catalyst & Chemicals(w/equip.)
   Land Cost                                                     2,500
                                                            -----------
        TOTAL CAPITAL REQUIREMENT(TCR)                      $4,972,699


 OPERATING & MAINTENANCE COSTS(First Year)
   Operating Labor                                              27,349
   Maintenance Labor                                            32,723
   Maintenance Material                                         49,084
   Administrative & Support Labor                               18,021
                                                            -----------
        TOTAL OPERATION & MAINTENANCE(1st yr.)                $127,177

 CONSUMABLE OPERATING COSTS(less Fuel)
   Water                                                         1,550
   Chemicals                                                    11,974
   Other Consumables
   Waste Disposal                                               16,093
                                                            -----------
        TOTAL CONSUMABLES(1st yr.,-fuel)                       $29,617

 BY-PRODUCT CREDITS(First Year)                               ($24,013)

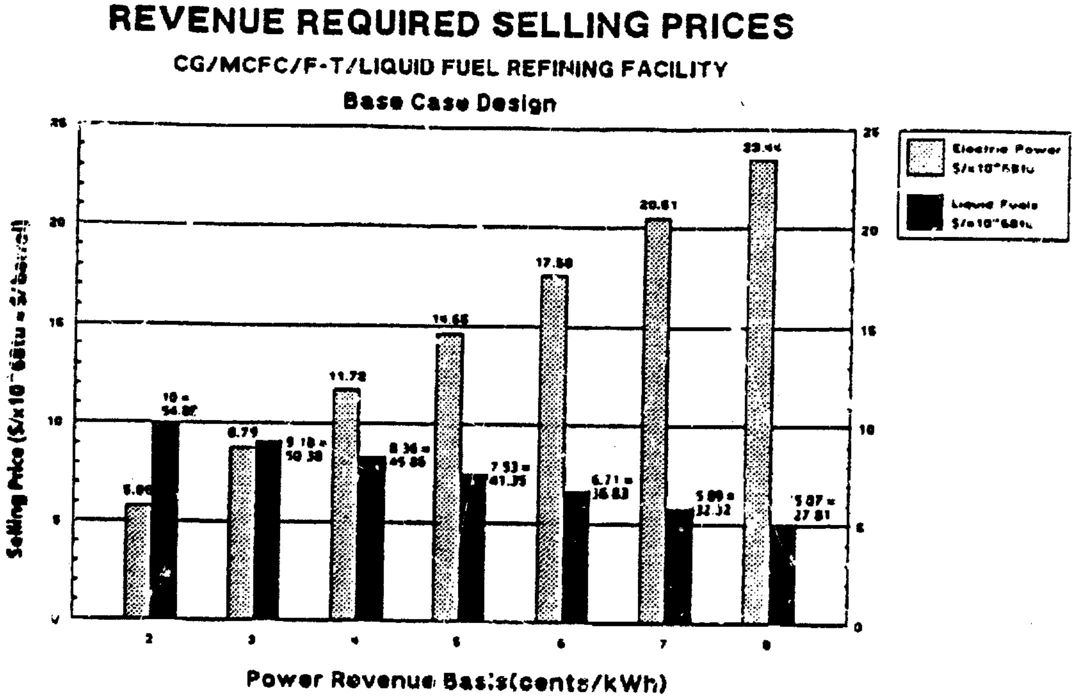
 FUEL COST(First Year)                                        $248,358

                                                                         $x1000          %
 1st YEAR OPERATION & MAINTENANCE COSTS SUMMARY                        ----------      ----
   Plant O & M                                                          $127,177        10%
   Consumables                                                           $29,617         2%
   By-product Credit                                                    ($24,013)       -2%
   Fuel                                                                 $248,358        20%
 CARRYING CHARGES(Capital)                                              $835,413        69%
                                                                      ------------    ------
        TOTAL ANNUAL COST(1st Year Basis)                             $1,216,552       100%


                                                                                    % Req'd
 REVENUE BASIS                                                                      Revenue
                                                                                  ----------
 1st YEAR VALUE OF POWER GENERATED(cents/kWh)           7.0                           35.3%

 1st YEAR LIQUID FUELS COST($/x10^6Btu-$/Barrel)      $7.53       $41.35              64.7%
```

FIGURE 11



REVENUE REQUIRED SELLING PRICES

CG/MCFC/F-T/LIQUID FUEL REFINING FACILITY

Base Case Design

the coal is best utilized and hence make all the technologies involved more economically feasible.

It was found that, by designing the plant to operate in a once-through Fischer-Tropsch mode, substantial amounts of high-value electricity are produced. By using the molten carbonate fuel cell, converison to electricity is more efficient resulting in about 12 percent of the HHV value of the coal feed ending up as electricity. Since electricity has the highest value of the products produced, this results in over a third of the revenue coming from electricity. Because of this, economics of the plant depend upon both the product mix and selling price.

It appears that prospects for SFT plant will improve in view of the leverage that electric power can provide for pricing liquids. The revenue required from the sale of liquids decreases with the rising price of electricity, and for this study the effect of pricing electricity from 2 to 8 cents per kWh is to reduce the required selling price of the liquids from $55 to $28 per barrel. However, the cost of products from the plant continues to be higher than the cost of an equivalent barrel of product produced from crude oil due to the competitive environment at this time for crude oil.

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AUTHORS INDEX