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FINAL REPORT

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FOSSIL-FUEL PROCESSING TECHNICAL/PROFESSIONAL SERVICES

MASTER

CRUDE-OIL VS. COAL-OIL PROCESSING COMPARISON STUDY

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UOP SDC

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DOE Task No. 9
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1.0 SUMMARY

The economic viability of upgrading coal derived liquids to acceptable motor fuel and heating fuel end products is presently of great interest in this country. This study evaluates three refinery schemes that have been developed for the processing of H-Coal liquid obtained from the Hydrocarbon Research Incorporated's (HRI) coal liquefaction process. The refinery processing employed for the naphtha and lighter components of the H-Coal liquid is essentially the same for all three schemes. It is in the processing of the H-Coal distillate product that refinery variations occur, and these differences are outlined below:

1. Hydrotreating of the middle coal distillate to produce a No. 2 fuel oil equivalent product
2. Hydrocracking of the total coal distillate to produce more gasoline and higher quality distillate fuel
3. Hydrotreating of the light coal distillate to a No. 2 fuel oil equivalent, and hydrogenating the heavy coal distillate to upgrade feedstock to a fluid catalytic cracking (FCC) unit

To provide a perspective of the value of coal liquid relative to petroleum, a parallel set of petroleum refinery schemes, processing a 65/35 Light/Heavy Arabian crude oil blend, was developed. A summary of the major petroleum refinery scheme variations is given below:

1. Reduced crude desulfurization with FCC processing of the desulfurized VGO

2. Reduced crude desulfurization with hydrocracking of the desulfurized VGO
3. Solvent demetallization of the vacuum pitch with desulfurization and FCC processing of VGO and demetallized oil
4. Solvent demetallization of the vacuum pitch with hydrocracking of the VGO and demetallized oil

Various gasoline to distillate ratios were set as parameters in developing the best possible processing scheme for the petroleum and coal liquid refineries.

Linear programming (LP) techniques were used to select the optimal schemes at various product ratios. The LP helped provide preliminary yields, material balances, capital cost, utility and operating cost information to determine the profitability of each scheme at the specified gasoline/distillate ratio. The refinery economics are based on typical product prices and a 30% cost of capital for the total refinery investment. The 30% cost of capital allows for a 15% after-tax discounted cash flow rate of return on investment.

Applying the same product prices to all cases and subtracting operating costs and the capital change, a comparative feedstock value is calculated. This method places the various refinery schemes on a common basis and gives an appraisal of the relative value of the H-Coal liquid charge stock, based on new refinery facilities.

Individual "base case" process unit investment costs were scaled from detailed estimates which reflect U.S. East Coast construction to UOP standards. The LP model has the capability of adjusting the investment costs exponentially for variations in capacity from the base case.

The study results are specific only to the parameters established in the study basis. Variations in the study basis, such as charge stock type, product prices, or product requirements could have a significant impact of the study results.

The more conventional petroleum refining schemes, based on separation of a high-sulfur residue and desulfurization of FCC feedstock, were penalized by rigorous constraints on fuel oil sulfur content and refinery emissions and by the processing of only sour imported crude oil. The constraints were imposed to provide consistency between coal and petroleum derived products, and the crude oil is the "swing" crude representing the most likely flexibility in U.S. supply. Complex schemes, including residue hydrotreating and hydrocracking, required to meet these objectives would tend to have higher costs and result in a lower petroleum value. A different crude slate or product specifcations, or a market for a small amount of high sulfur fuel oil, would tend to shrink the differential between the petroleum and coal liquid refinery cases.

The results below show the relative product value at each product ratio for the petroleum and coal liquid refinery:

	Raw Material Value ($/Barrel)	
Gasoline/Distillate Ratio	Coal Liquid	Petroleum
0.7	20.21	17.35
1.0	20.63	17.60
1.5	20.21	17.82
2.0	20.23	17.89

If the coal liquid feed contained a significant amount of material distilling higher than the product endpoint, or if the evaluation considered processing in an existing refinery designed specifically for upgrading petroleum, the relative value of coal liquid would be substantially lower.

2.0 INTRODUCTION

In response to the ever increasing difficulty in securing this country's energy needs, many energy sources previously considered exotic or hard-to-tap are now being given careful scrutiny. One such area is the conversion of coal to synthetic hydrocarbon liquids suitable as substitutes for conventional petroleum products, such as gasoline or heating fuel oils.

At the present time, considerable effort and money are being expended to evaluate and develop feasible coal liquefaction schemes. Whichever scheme is finally selected as the most viable, the coal liquid products obtained from that scheme will still have to undergo considerable upgrading before they are considered environmentally and functionally acceptable end products. The upgrading required would be performed in a coal liquids refinery, applying technology used in many petroleum refineries, but with processing capabilities specifically designed to handle coal liquids.

2.1 Purpose of Study and Objectives

This study evaluates the processing requirements and economics of upgrading a coal liquefaction product to acceptable motor fuel and heating fuel. In order to get a more complete perspective of the significance of the processing characteristics anticipated in refining coal liquid, a parallel case incorporating a new, high-conversion refinery for imported petroleum has been developed for comparison. The side-by-side analysis of these two refineries enables us to evaluate more effectively the economic and processing differences between refining coal liquid and crude oil, as outlined in the study objectives listed below:

1. Develop various feasible coal liquid refining schemes to produce the following product slate:

   a. LPG

   b. Unleaded Gasoline

   c. 0.2 wt-% Sulfur No. 2 Fuel Oil

d. 0.7 wt-% Sulfur No. 6 Fuel Oil

2. Evaluate the schemes developed above at gasoline-to-distillate fuel oil ratios varying from 0.7 to 2.0

3. Develop modern high conversion import petroleum refinery schemes as in No. 1 above

4. Evaluate the petroleum schemes at the same gasoline-to-distillate fuel oil ratios as in the coal oil schemes

5. Determine the optimum refining schemes at each of the gasoline-to-distillate fuel oil ratios considered

6. Examine the differences in operating requirements and investments between the optimum coal liquid and petroleum refinery cases

7. Indicate any unusual processing characteristics or product properties anticipated in the processing of coal liquid

The coal liquid analysis on which this study is based was derived from Hydrocrabon Research Incorporated's (HRI) H-Coal process, while the petroleum is a 65/35 volumetric blend of Light/Heavy Arabian crude oils. *It must be cautioned that the applicability of the results of this report to the refining of coal liquids obtained from other processes has not been considered and would require separate studies.*

These upgrading studies, together with studies of coal liquefaction costs, can be applied to assess the incremental cost of securing domestic supplies of liquid hydrocarbon fuels.

3.0 BASIS OF STUDY

The basis of study was established in a Department of Energy contract for the study of coal liquid upgrading awarded to UOP/SDC (UOP/System Development Corporation consortium).

3.1.1 Coal Oil Source

The charge to the upgrading complex is C_4-470°C distillate obtained from the atmospheric column overhead and bottoms products of the HRI H-Coal process.

3.1.2 Goal Liquid Properties

Coal liquid properties were obtained from the following sources:

1. November 28, 1978, Mobil Research and Development Corporation analyses of HRI H-Coal syncrude performed under DOE contract EF-77-C-01-2676.
2. HRI analyses of H-Coal process PDU, Run 5.
3. UOP laboratory analyses of H-Coal distillate samples obtained from HRI.

The Mobil R&D data were the primary source of coal liquid charge stock information for this study, with the other information available being used primarily in a supplementary capacity. The usage of these data is discussed below.

3.2.1 Crude Oil Source

A 65/35 volume ratio blend of Light/Heavy Arabian crude oils.

3.2.2 Crude Oil Properties

A summary of the composition of the crude oil blend is:

	BPSD	API Gravity	Wt. Sulfur
Light Arabian Crude	65,000	34.4	1.69
Heavy Arabian Crude	35,000	28.2	2.84
Blend	100,000	32.2	2.09

Two sets of crude oil fractions were established to provide the processing charge stocks necessary to achieve the product ratios set for the study.

True Boiling Point Cuts, °C	Vol-%	Wt-%	BPSD	Sp.Gr.
C_4-	2.40	1.57	2,400	---
C_5-65	5.11	3.83	5,110	0.6477
65-161	13.86	11.75	13,860	0.7328
-or-				
65-193	19.72	17.03	19,720	0.7467
161-260	17.04	15.74	17,040	0.7982
-or-				
193-260	11.30	10.56	11,300	0.8076
260-352	14.61	14.42	14,610	0.8535
352-565	29.40	31.12	29,400	0.9153
565+	17.93	21.47	17,930	1.035

(Complete petroleum and coal liquid properties are presented in Appendix A.)

3.3.1 Product Slates

The product slate for both the petroleum and coal liquid refineries will be restricted to make only:

1. LPG
2. Unleaded Gasoline
3. No. 2 Fuel Oil
4. No. 6 Fuel Oil

3.3.2 Product Specifications

The products listed above will be constrained to conform to the following requirements:

1. Unleaded gasoline will be produced in both regular and premium grades with the (Research + Motor octanes)/2 rating for these two grades being 88.0 and 92.5, respectively. The ratio of unleaded regular to unleaded premium gasoline has been set at 70:30 as per literature predictions for the early 1980's

2. A maximum of 0.2 wt-% Sulfur is permitted in No. 2 Fuel Oil

3. A maximum of 0.7 wt-% Sulfur is permitted in No. 6 Fuel Oil

4. All refinery products are to meet ASTM specifications as provided in Appendix B

5. In the refinery LP optimizations, some of the product specifications for gasoline and fuel oil are blended using indices. Those product specifications using blending indices are:

   a. Reid Vapor Pressure

   b. Flash Point

   c. Pour Point

   d. Viscosity Index

   These indices are also provided in Appendix B.

6. The gasoline reid vapor pressure specification has been set at an average maximum of 10.5 psig, as the ASTM accepted RVP specifications are not normally utilized in commercial applications.

3.3.3 Product Requirements

The product slates considered in this study are based on gasoline to middle distillate ratios ranging from 0.7 to 2.0; LPG and No. 6 fuel oil are sold on an "as produced" basis.

3.4.1 Operating Assumptions

Refinery capacity for both cases is 100,000 barrels per calendar day with an on stream efficiency of 330 stream days per year for both refineries. Only power and water are purchased. Other utilities such as fuel and steam, are produced internally in the refinery.

3.4.2 Refinery Construction and Configurations

1. Refinery construction has arbitrarily been chosen as the Philadelphia area based on its proximity to the Eastern coal fields and upon its status as a major refining area.

2. Estimated erected costs are on a U.S. East Coast basis

3. Refinery configurations for both the crude and coal oil refineries are based on hypothetical new unit construction emphasizing high conversion processing in an environmentally acceptable context

3.5 ECONOMIC BASIS

3.5.1 Investment Costs

Investment costs are based on:

1. 2nd quarter, 1979 costs

2. Philadelphia area location

3. Adding an estimating allowance of about 10% of total on plot and off plot investment to account for additional cost items such as interest during construction

3.5.2 Working Capital Includes

1. Value of 30 days of feed and 15 days of product inventories in storage

2. Estimated value of spare catalyst and spare parts

3. Estimated value of accounts receivable less accounts payable for 30 days

3.5.3 Capital Charge Factor

1. In Tables 13 through 20, processing costs are based on a capital charge equal to 30% of cost investment (including working capital and initial catalyst costs) per year. The capital charge factor is approximately equivalent to a 15% after-tax discounted cash flow rate of return on investment based on:

   a. 51% income tax (based on 46% Federal and 9.5% State)

   b. Four-year construction period

   c. Investment payments equal to 10%, 15%, 25% and 50% of total investment during the four years of construction

   d. 50% of design capacity during first year of operation, 100% in second year and thereafter

   e. 10% investment tax credit

   f. Sum of the years' digits depreciation: 13-year tax life of refining equipment

   g. 20-year project life

3.5.4 Operating Costs

1. Maintenance cost estimated at 2% of onsite plus offsite investment per year

2. Taxes and insurance estimated at 2½% of onsite plus offsite investment per year

3. Labor charged at an average rate of $9.50/hr with a 35% allowance for fringe benefits and a 25% allowance for supervision

4. Utilities charges for purchased utilities at:

   a. Power at 3.5¢/kWh

   b. Raw water at 25¢/cubic meter

3.5.5 Product Prices

The product prices assigned are somewhat arbitrary, but are based upon typical current prices:

	Price ($/Barrel)
LPG	18.00
Unleaded Regular Gasoline	29.50
Unleaded Premium Gasoline	30.50
0.2 wt-% Sulfur, No. 2 Fuel Oil	26.50
0.7 wt-% Sulfur, No. 6 Fuel Oil	21.00
Sulfur ($/MT)	50.00

3.5.6 Charge Stock Values

Charge stock values are determined independently for each scheme. They are derived by deducting the operating costs, general and administrative costs, and a 30% capital charge on total investment from the gross refinery sales.

3.5.7 General Refinery Costs

General and administrative costs have been fixed at five million dollars per year in this study.

3.5.8 Overall Study Economics Basis and Limitations

This study is based upon current 1979 construction costs and product prices. No attempt has been made to provide for cost or price escalation.

The refinery offsites costs are based on refinery complexity factors and do not reflect a comprehensive offsites evaluation.

The capital charge factor used for economic evaluation is based upon that derived in a study by Chevron Research Company under DOE Contract EF-76-C-01-2315, which appears consistent with the basis of this study.

4.0 DISCUSSION

4.1 Charge Stock Properties

A detailed summary of the crude oil and coal liquid properties used in this study is provided as Appendix A. In this section are discussed those properties which were significant with respect to both process selection and processing severity.

The 65/35 Light/Heavy Arabian blend used in this study represents a typical imported crude oil. Its properties are generally considered intermediate for some imported crudes presently available. The coal liquid on the other hand bears little resemblance to any crude oil now being processed. The comparison provided below should serve to amplify the preceding statement by highlighting important coal liquid properties:

4.1.1 Charge Stock Comparison - Based on Wt-% of Feedstock

Light Gases (C_4-)

Crude Oil — 1.6 wt-% consisting primarily of propane and butanes.

Coal Oil — 5.6 wt-% consisting only of butanes. All lighter than butane product has been assumed to be consumed in the H-Coal process.

Light Naphtha

Crude Oil — 3.8 wt-% with a (R+M)/2 of 65. Because of its low octane and easy processibility, upgrading of the light naphtha stream is an effective method of achieving improved gasoline pool octane.

Coal Oil — 4.2 wt-% with a (R+M)/2 of 85 due to relatively high level of naphthenes and olefins present. Olefin level is about 10%.

Heavy Naphtha

Crude Oil

(65-160°C cut = 11.8 wt-%, 65-193°C cut = 17 wt-%) The contaminant levels of the heavy naphtha are low and do not present any processing difficulties. This Arabian naphtha blend is considered to have a relatively low cyclic content (naphthenes 20-25 vol-%, aromatics 10-15 vol-%), but nevertheless is readily upgradable to a high octane gasoline blending component by modern catalytic reforming.

Coal Liquid

(60-176°C cut = 30 wt-%, 60-205°C cut = 40 wt-%) As is illustrated in Appendix A, the coal liquid naphtha contains large quantities of nitrogen (500-3000 ppm depending on cut selection), and a very high level of oxygen in the form of phenols and cresols. The oxygen level varies from 5000 ppm for the 60-176°C cut to 30,000 ppm for the 60-205°C cut. Examination of Appendix A indicates that the majority of the oxygen falls in the 170-210°C range where most phenols and cresols are concentrated. A significant amount of diolefins are present in the coal naphtha (about 2 wt-%). These diolefins must be given careful consideration in naphtha fractionation or upgrading.

The coal liquid naphtha has proven to be a very rich feedstock in terms of high octane blending components (aromatics) and aromatic precursors (naphthenes). This is illustrated by the paraffin/ naphthene/aromatics "PNA" levels observed for the hydrotreated coal naphthas listed below:

	60-176°C Cut	60-205°C Cut
Paraffins, Vol-%	20.0	14.5
Naphthenes, Vol-%	50.8	38.1
Aromatics, Vol-%	29.2	47.4

The aromaticity of the naphtha is further reflected by its low hydrogen content, 11.6 wt-%, versus about 14.5 wt-% for a typical crude oil naphtha. The sulfur level of the coal liquid naphtha is about 0.2 wt-%, which is not limiting in setting hydrotreating conditions relative to other contaminants.

Distillate

Crude Oil

(193-350°C cut = 25 wt%) The sulfur level of the crude distillate is 0.95 wt-%; otherwise contaminant levels are low and pose no processing problems. The hydrogen content of this material is about 13 wt-% which is indicative of its suitability as a transportation fuel.

Coal Liquid

(204-343°C cut = 45.2 wt-%) The quantity of the coal liquid middle distillate produced is substantially greater than that produced from crude oil. This is to be expected as all the products recovered from coal liquefaction for processing are distillable as opposed to the large quantity of residue present in the Arabian crude oil blend.

The coal liquid distillate is a highly aromatic material with high levels of contaminants. The aromatics concentration is reflected by a 15 °API gravity and a hydrogen concentration of only 9.7 wt-%. The oxygen level of 1.0 wt-% is present mostly in the form of phenolic compounds. Nitrogen is present in a concentration of 4000 wt-ppm and causes the H-Coal liquid to be quite unstable in storage with rapid gum formation occurring. This high nitrogen level will pose the most difficult obstacle in upgrading the coal distillate. Olefins are also present in the distillate in a concentration of about 6 wt-%. The combination of olefins and nitrogen compounds such as pyridines results in a bromine number of 50 for this material. Although some ash has been found in some of the analyses used as the basis for this study, it is felt that proper fractionation in a commercial coal liquefaction facility would preclude the presence of ash in the middle distillate. As in the coal oil naphtha, the distillate sulfur level of 0.2 wt-% is quite low.

Residue

Crude Oil

(350°C+ residue = 52.6 wt-%) The largest segment of the Light/Heavy Arabian crude oil blend is the 350°C+ residue. Of this residue, about 59% (31.1 wt-% of crude) is estimated to be distillable by vacuum fractionation in this study. The processing of the reduced crude is the primary factor in the selection of the various crude oil refinery schemes under consideration. While the vacuum gas oil (VGO) is comparatively easy to process because of its low metals levels, upgrading of the residue by direct catalytic processing would be very costly. Alternative schemes are considered for processing vacuum residue to gasoline and distillate fuel oil blending components, while minimizing the costs required to reduce the sulfur, metals and nitrogen contents of the residual product to allowable levels.

Coal Liquid

(343-417°C cut = 4.9 wt-%) This heavy distillate cut represents only a small fraction of the total H-Coal liquid. All heavier liquids produced in the H-Coal process are either recycled or utilized for hydrogen production in the liquefaction process. The heavy distillate has an API gravity of 1.3°, and is almost completely composed of aromatic structures as verified by the low 8.0 wt-% hydrogen content. The nitrogen level of 1.3 wt-% along with a high oxygen concentration of 1.3

wt-% and a 1200 ppm ash level indicate that this material is both unstable and difficult to process.

4.1.2 General H-Coal Liquid Characteristics

The following comments can be made concerning the properties of the H-Coal liquid:

1. A high olefin content of about 6-8 wt-% is present throughout the coal liquid
2. The product sulfur level is consistently low throughout the whole distillation range
3. Oxygen compounds are almost non-existent in the light naphtha, increase to almost 10 wt-% oxygen in the 175-205°C naphtha, and then drop to about 1 wt-% oxygen through the middle and heavy distillate range
4. The coal liquid hydrogen content averages more than 3 wt-% lower than that for equivalent crude oil cuts
5. The significant bromine number is primarily due to the nitrogen compounds present, such as pyridines, and secondarily due to olefins
6. Consistent diene values are not available for the coal liquid, but the test data do indicate that special consideration will have to be given to process design because of the presence of dienes
7. A summary of the H-Coal liquid product quantities is given below:

Cut	BPSD	Wt-%
C_4-C_6	13,870	9.80
C_6-204°C Naphtha	41,250	40.14
204-343°C Distillate	40,890	45.19
343-471°C Distillate	3,990	4.87
	100,000	100.00

4.2 PROCESS FLOW SCHEMES

4.2.1 Petroleum Refinery Processing Schemes

The criteria used for scheme selection of the petroleum refinery were based upon a grass roots facility processing 100,000 barrels per calendar day (BPCD) of imported crude oil to yield a high proportion of light products. The feedstock consists of a 65/35 Light/Heavy Arabian crude oil blend. The rationale for process unit selection was derived primarily from the hypothetical refinery product slate which required the following:

1. High conversion processes
2. Low sulfur fuel oil products
3. Unleaded gasoline production
4. Gasoline to distillate fuel oil product ratios ranging from 0.7 to 2.0

A more comprehensive discussion of refinery product slates is provided later in this report. Also, generalized process descriptions for the various refinery schemes are presented in Appendix D.

In light of the product requirements listed above, four primary processing schemes were developed. The processing of all atmospheric gas oil (AGO) and lighter crude oil fractions was essentially the same in all of these schemes, and typical of processing observed in a present day refinery. The schemes are differentiated by the method of upgrading atmospheric residue to lighter, low-contaminant products, and the refinery scheme nomenclature is based on these crude oil residue processing options. The four primary petroleum oil refinery schemes are depicted in Figures 1-4 and are listed below:

<u>Scheme I</u> Reduced crude desulfurization with fluid catalytic cracking (FCC) processing of the desulfurized VGO

<u>Scheme II</u> Reduced crude desulfurization with hydrocracking of the desulfurized VGO

Scheme III   Solvent demetallization of the vacuum pitch with desulfurization and FCC processing of VGO and deasphalted oil

Scheme IV   Solvent demetallization of the vacuum pitch with hydrocracking of the VGO and deasphalted oil

Central to the requirement of selecting an optimal scheme or schemes is the need to examine product slates having gasoline/distillate fuel oil ratios ranging from 0.7 to 2.0. The most attractive scheme at one product ratio may no longer be optimal when an alternative product ratio is the target.

The processing of the lighter than 350°C crude oil fractions is essentially the same in all the schemes with only minor variations.

C_4- Light Gases

All the light petroleum gases together with saturated gases from most of the refinery's process units are sent to a gas recovery section. In this section, H_2S gas is removed from the light hydrocarbons via amine absorption. The H_2S is converted in a sulfur recovery plant to yield sulfur as a product. C_2- and lighter gases are separated from propane and butanes by fractionation and utilized as fuel gas. Propane/butane is treated for mercaptan removal and recovered as LPG, with the butanes being used for gasoline blending.

C_5/C_6 Light Straight Run Naphtha

Total naphtha from atmospheric crude distillation and stabilization is hydrotreated and fractionated to yield light and heavy naphtha cuts. Because of the rather stringent octane demands on a refinery producing unleaded gasoline, the light naphtha is readily upgraded via isomerization to avoid more difficult octane upgrading in other parts of the refinery.

Heavy Naphtha

The hydrotreated heavy naphtha is catalytically reformed to produce a high octane gasoline blending component. In the schemes where hydrocracking is employed, the heavy hydrocracked naphtha is combined with the heavy straight run naphtha as reformer charge stock. Two alternative heavy naphtha cuts, a C_7-160°C and a C_7-193°C, are considered in order to facilitate product slate variation.

Kerosine

Kerosine cuts of 160-260°C and 193-260°C are examined. These cuts are either treated for mercaptan removal or hydrotreated if product sulfur blending proves difficult. The kerosine is either blended directly to No. 2 fuel oil or is used as a cutter stock for refinery or No. 6 fuel oil.

Atmospheric Gas Oil (AGO)

The 260-350°C AGO is hydrotreated and blended to fuel oil. It is also used as a charge to a fluid catalytic cracker (FCC), where maximum conversion to gasoline is desired.

Significant variations in the crude oil processing schemes are present primarily in the processing of the 350+°C crude oil residue. These variations of the four schemes and their strengths and weaknesses are discussed.

Scheme I. The 350+°C atmospheric column residue is desulfurized to greater than 90 wt-% sulfur removal. Such desulfurization is very costly, but it does enable the refinery to meet all the sulfur requirements set for both the product No. 2 and No. 6 fuel oils as well as the internal refinery fuel oil without further sulfur removal steps. Residue desulfurization decreases the quantity of vacuum residue present and improves the quality of the vacuum gas oil in terms of processibility. The VGO recovered by vacuum distillation of the desulfurized product is further processed either in a FCC or a hydrocracker to obtain the desired product slate.

FCC products must be further treated or processed before being blended into finished products. The FCC produces a significant quantity of C_3/C_4 olefins which are treated and either alkylated with isobutane or polymerized to make high octane gasoline blending components. The FCC gasoline is treated before blending, while the FCC middle distillate (light cycle oil, "LCO") is used in fuel oil blending. The aromatic nature of the LCO makes it a poor diesel fuel component; however, it is quite acceptable in a heating fuel. The FCC slurry product having a high boiling point and aromatic content is suitable for use only as refinery fuel or in heavy fuel oil.

Scheme II. In this scheme the 350+°C atmospheric residue is again desulfurized; however, the VGO is sent to a hydrocracker. The hydrocracker design can provide for great flexibility in product slate. Hydrocracking does produce high quality distillate, but at the expense of increased capital investment and operating costs as compared with FCC processing of the same charge stock.

In contrast to the FCC products, the hydrocracker products are high in hydrogen content. The hydrocracker light naphtha is blended directly to gasoline, and the heavy naphtha makes a desirable catalytic reformer feed because of its relatively high liquid yield and hydrogen yield upon reforming. The hydrocracker distillates make desirable diesel or No. 2 fuel oil blending components.

The upgraded vacuum residue produced in both Schemes I and II (sulfur content 0.8-1.0 wt-%) is blended with lighter distillate components to meet No. 6 fuel oil specifications.

Scheme III. The 350+°C atmospheric residue is charged to a vacuum distillation column from which 350-565°C vacuum gas oil and 565+°C vacuum residue are recovered. The vacuum residue is processed in a solvent demetallizing unit to produce a heavy, high sulfur demetallized oil and a heavy, high contaminant pitch. This pitch is of such poor quality that it can be disposed of in only a limited number of ways such as:

1. It can be processed in a partial oxidation unit to produce refinery hydrogen. This is very expensive, but it does convert the high level of sulfur in the pitch (6.5 wt-%) to H_2S

that can be accommodated readily in conventional processes.

2. The pitch can be blended with a distillate (to make it pumpable) and burned as refinery fuel. In this case the flue gas would be very high in SO_x and therefore would be treated for SO_x removal. This option is also expensive, but it does have the advantage of also being able to treat NO_x in the flue gas at a relatively small additional cost.

3. Blending of the pitch with hydrotreated distillates to produce No. 6 fuel oil. This is the least acceptable option because the quantity of valuable distillate required would be about 10 times that of the pitch being blended.

The demetallized oil produced from solvent deasphalting is combined with the vacuum gas oil, desulfurized at high severity and then processed in an FCC unit as in Scheme I.

Scheme IV. This scheme is similar to Scheme III except that the blended demetallized oil and vacuum gas oil are sent directly to a hydrocracker to produce desired products.

Because of the stringent sulfur specifications set for the No. 2 and No. 6 fuel oils, a substantial amount of hydrogen is required in these processing schemes. In the cases involving hydrocracking this demand is even further amplified. To meet the hydrogen requirements of these refinery schemes hydrogen production facilities are required. The H_2 plant feedstocks are light refinery gas streams or light naphtha if a steam reforming plant is installed. If a high contaminant pitch must be disposed of, then a partial oxidation plant can be utilized.

4.2.2 Coal Liquid Refinery Processing Schemes

The coal liquid processing schemes developed are actually less complex than those developed for the petroleum refinery. This is because the coal liquid contains much less residual material than the crude oil. However, processing to obtain acceptable products is severe and somewhat unique due to the nature of the contaminants present in the coal liquid. The properties of the coal liquid which have the greatest impact on its processing are as follows:

1. Distillation - Examination of Appendix A shows that most of the C_4-471°C coal liquid falls either in the naphtha or middle distillate range.
2. Nitrogen - The high nitrogen concentration present through most of the coal liquid range necessitates severe hydrotreating to produce:
   a. Acceptable catalytic reformer charge stock
   b. Stable fuel oil
3. Low Hydrogen Concentration - Large quantities of hydrogen are required if cracking of the coal distillate is desired.
4. Diolefins - Heaters and heat exchange must be given careful consideration to avoid coking of the tubes.
5. Oxygen - The oxygen concentration is very high, especially in the heavy naphtha/light distillate range. Almost all of the oxygen must be removed from the naphtha before it is processed in a catalytic reformer.

Based on the coal liquid characteristics and the product slates desired, three basic processing schemes were developed. The processing of the naphtha and lighter coal liquid is the same in all schemes. The variations examined are in the processing of the coal liquid distillate. These three schemes are depicted in Figures 5, 6 and 7, and the functional characteristics are summarized below:

1. Scheme A (Figure 5) - Hydrotreating of the middle coal distillate to produce a No. 2 fuel oil product.
2. Scheme B (Figure 6) - Hydrocracking of the total coal distillate to produce more gasoline and higher quality distillate fuel than in Scheme A.
3. Scheme C (Figure 7) - Hydrotreating of the light coal distillate to a No. 2 fuel equivalent, and hydrogenating of the

heavy coal distillate to produce a "crackable" charge stock which is processed in an FCC unit.

A key element in coal liquid processing is the low-temperature, low-pressure fractionation employed. This fractionation scheme was developed because of the diolefins present in the raw coal liquid and the necessity to fractionate this material at low temperature (no higher than about 230°C) to avoid coke formation in the heater tubes. As a consequence all fractionation and first stage process heating on diolefin containing streams will be by steam heat exchange. The fractionation stages shown in Figures 5-7 are performed at successively lower pressures as lighter components are removed in previous stages. However, because of temperature requirements, mild hydrotreating to remove diolefins before the light/middle distillate splitter may be required. Alternatively, more severe hydrotreating of this feedstock, sufficient to produce stable fuel oil, may be utilized, but this alternative offers no real benefit for the following reasons.

1. If 260+ or 288+°C heavy distillate is recovered from this fractionator for FCC processing, this stream must undergo additional severe hydrotreating to yield a suitable FCC charge stock.

2. If a 343+°C heavy distillate is taken as a bottoms product from the fractionator, it will have been excessively hydrotreated since this stream is only suitable as refinery fuel or as a hydrogen source in a partial oxidation unit.

In all the liquid oil refinery cases, a C_4-C_6 cut is taken overhead in the primary (C_4-471°C) fractionator. The fractionator bottoms (C_6-471°C cut) is sent to a naphtha/distillate splitter. The fractionator overhead is hydrotreated to remove any nitrogen and sulfur compounds which may be present and to saturate olefins. The hydrotreated product is either used as feedstock for hydrogen production, blended directly to gasoline, or split into C_4 and C_5/C_6 fractions to facilitate LPG production and more flexible gasoline blending. Although not depicted in Figures 5, 6 and 7, light gas treating and processing is included in all refinery schemes.

The primary fractionator bottoms (C_6-471°C cut) is charged to a naphtha/distillate splitter where a variable naphtha cut is taken overhead and a variable distillate cut is recovered as bottoms. The cuts taken from this splitter were varied to provide flexibility in achieving a wider range of gasoline/distillate product ratio. The cuts considered are as follows:

Naphtha	Distillate
C_6-177°C	177-471°C
C_6-193°C	193-471°C
C_6-204°C	204-471°C

A further reason for providing the 177-204°C swing cut between naphtha and distillate was the extremely high oxygen level in this cut (about 10 wt-%). This oxygen level (present primarily as phenolic compounds) must be reduced to very low levels via hydrotreating; otherwise water formation might be so rapid in a catalytic reformer as to impair the performance of the reformer, even if recycle gas drying were utilized. The naphtha hydrotreating required in a coal liquid refinery is much more severe than that normally used in a petroleum refinery due to the high nitrogen and oxygen content of the coal naphtha. A two stage hydrotreater is required to process the coal naphtha. The first stage operates at low severity and is used for diolefin saturation. The second stage operates at severe conditions and is used for denitrification, desulfurization and oxygen removal.

Chemically the coal naphtha is easily reformable to high octane levels because of the large quantity, high octane aromatics and aromatic precursors in the coal naphtha. However, due to the relatively large quantities of indanes, tetralins, decalins and phenolic compounds present, continuous catalyst regeneration may be advisable.

It is the distillate processing downstream of the naphtha/distillate splitter which provides the differences between the three processing schemes selected. The processing involved specific to these schemes is as follows:

<u>Scheme A.</u> The distillate obtained from the naphtha/distillate

splitter is charged to a middle distillate/heavy distillate splitter. A 343°C endpoint middle distillate cut is recovered overhead from this splitter. This distillate is severely hydrotreated to produce a storage stable No. 2 fuel oil type product. At least 90% denitrification is required to produce a stable middle distillate.

The 343-471°C heavy distillate is used as refinery fuel or as feed to a partial oxidation plant for hydrogen production. Because of its high level of contaminants, the heavy distillate will probably have to be inhibited for use as refinery fuel oil.

Scheme B. The total naphtha/distillate splitter bottoms is charged to a hydrocracker. Hydrocracking severity is set by the gasoline/No. 2 fuel oil ratio desired. The hydrogen consumption required to hydrocrack coal distillate is very high and can vary from 2.5-5.0 wt-% of charge depending on the level of conversion required. Although oxygen removal and denitrification requires significant amounts of hydrogen, most of the hydrogen utilized is consumed in saturating aromatic compounds and hydrocracking.

Scheme C. The naphtha/distillate splitter bottoms is further fractionated into light and heavy distillate cuts. The light distillate is hydrotreated to produce a stable fuel oil. The heavy distillate is severely hydrotreated to produce an upgraded distillate having a hydrogen content of 11 wt-%, which is suitable as an FCC feedstock. Laboratory tests have indicated that the unprocessed heavy distillate is much too refractory because of its low hydrogen content to achieve acceptable conversion to desirable products in an FCC unit; however, hydrogen saturation to about 11 wt-% does produce a crackable FCC charge stock. The expected conversion of the upgraded FCC feed is about 65 vol-% versus 30 vol-% for the raw distillate. To provide for product flexibility the following cuts were considered for Scheme C:

	Light Distillate	Heavy Distillate
High Gasoline Cases	204-260°C	260-471°C
Low Gasoline Cases	176-288°C	288-471°C

Alternative Coal Liquid Processing

Although not considered in this study, a realistic H-Coal liquid processing scenario may be the hydrotreating of the total coal liquid in the H-Coal plant as it is produced. Because of the storage instability of the H-Coal liquid product, the immediate hydrotreating of this material before refinery processing may prove necessary. While the relative instability of the coal liquid may not be a problem if the H-Coal plant and refinery are in close proximity, there may be problems with polymer formation if the coal liquid must be pumped through a long pipeline or if kept in storage for a significant length of time before processing.

Though the inclusion of a total coal liquid hydrotreater reduces subsequent downstream hydrotreating, the net effect will probably be slightly increased capital and processing costs; the processing conditions set for hydrotreating of the total H-Coal liquid cannot practically meet the hydrotreating and product requirements of the various coal liquid fractions simultaneously.

4.3 MATERIAL BALANCES AND PRODUCT PROPERTIES

4.3.1 Petroleum Oil Refinery

Overall material balances for all the petroleum refinery schemes are presented in Tables 1-4. Also included in these tables are the throughputs for each process unit associated with each processing scheme. The actual LP output including material balances and product properties for the various schemes is shown in Appendix C. The following products may be produced, with gasoline and No. 2 fuel oil being required products:

1. LPG
2. Unleaded Regular Gasoline
3. Unleaded Premium Gasoline
4. No. 2 Fuel Oil

5. No. 6 Fuel Oil

6. Sulfur

These products are produced in the ratios specified in the basis of this study (gasoline/No. 2 fuel oil ratios of 0.7, 1.0, 1.5 and 2.0) and meeting U.S. ASTM product requirements as listed below:

1. LPG is not specified directly by any blending properties, but recovered on an "as produced" basis. LPG should meet product requirements as set by proper refinery design.

2.

Gasoline	Octane (R+M)/2	Vapor Pressure (kPa)
Unleaded Regular Gasoline	88.0	48-93
Unleaded Premium Gasoline	92.5	48-93

Distillation, °C	Unleaded Regular Gasoline	Unleaded Premium Gasoline
Sulfur, Wt-%	0.1	0.1
Minimum @ 55	10%	10%
Maximum @ 77	50%	50%
Minimum @ 113	50%	50%
Minimum @ 185	90%	90%
Maximum Endpoint	225°C	225°C

3. No. 2 Fuel Oil

Maximum Sulfur, Wt-%	0.2
Minimum Flash Point, °C	38
Maximum Pour Point, °C	-7
Distillation, °C	
Maximum @ 282	90%
Minimum @ 338	90%
Maximum Viscosity cSt @ 38°C	3.6
Minimum Viscosity cSt @ 38°C	2.0

4. No. 6 Fuel Oil

Maximum Sulfur, Wt-%	0.7
Maximum Viscosity cSt @ 50°C	638

Projection of the gasoline market for the 1980's indicates that two grades of unleaded gasoline will be produced. The (R+M)/2 octane ratings for the two grades of unleaded gasoline should remain at about 88 and 92.5, with a regular/premium unleaded gasoline ratio of roughly 70/30.

In order to meet emissions regulations, the refinery fuel oil must meet the same 0.7 wt-% sulfur specification as the No. 6 fuel oil product, otherwise flue gas scrubbing must be employed to reduce stack emissions to an acceptable level.

Analysis of Scheme III at the various gasoline/distillate fuel oil ratios examined in this study indicates that Scheme III is not feasible due to excessive quantities of high metals, high sulfur pitch produced in this scheme. The pitch cannot be completely consumed either as fuel or for hydrogen production via partial oxidation. Blending of this pitch to produce acceptable No. 6 fuel oil would require such large quantities of distillate blending components as to greatly decrease the quantity of gasoline and No. 2 fuel oil produced at a given product ratio.

In Scheme IV all the high contaminant pitch produced is consumed within the refinery, because of the greater refinery fuel and hydrogen requirements than those in Scheme III.

4.3.2 Coal Liquid Refinery

The product slates for the three coal liquid refinery schemes have been set to meet the same specifications as those for the petroleum refinery. Also, the product ratio targets were identical. The overall coal liquid refinery product slates and unit throughputs are detailed in Tables 5-8. The actual LP output for the coal liquid refinery schemes studied is presented in Appendix C. Comparison of the coal liquid to similar petroleum product slates shows that a slightly greater quantity of gasoline and distillate products is produced. It is apparent that this is because no No. 6 fuel oil is produced in the coal liquid refinery, since H-Coal liquid contains little heavy material.

Aside from possible problems due to diolefins, fractionation of the various coal liquid cuts is fundamentally the same as fractionating petroleum oils. Emphasis in the remainder of this section will be devoted to the other primary coal liquid processing units.

C_4-C_6 Hydrotreater

This hydrotreater performs the following functions:

1. Saturates the olefins present in the C_4-C_6 cut (A large part of this feed may be used in hydrogen production.)
2. Removes any nitrogen or oxygen compounds, which could contribute to gasoline instability
3. Removes mercaptan and other sulfur, also a hydrogen plant catalyst poison

Naphtha Hydrotreater

This is a two stage hydrotreater with the first stage being a small, low severity, low temperature reactor for hydrogenating diolefins. The second stage is a relatively severe high pressure reactor, which is required for the difficult task of reducing the nitrogen level in the catalytic reformer charge from about 3000 ppm to 0.5 ppm. The oxygen content of the naphtha is also reduced to a level which will not promote excessive water formation in the catalytic reformer beyond that which can be controlled by a recycle gas drier.

The naphtha feedstock to the hydrotreater is so aromatic that the severe hydrogenation performed in the hydrotreater causes significant saturation of aromatics in the naphtha. This saturation causes the naphtha API gravity to increase as follows:

Naphtha Cut	Raw Naphtha API	Hydrotreated Naphtha API
C_6-177°C	44.5	48.3
C_6-193°C	36.8	44.4
C_6-204°C	35.0	43.4

Because of the naphtha saturation, there is also a significant shift in feed naphtha versus product naphtha chemical properties as shown below:

	Naphtha Fractions, °C					
	Raw Naphtha			Hydrotreated Naphtha		
Composition	C_6-177	C_6-193	C_6-204	C_6-177	C_6-193	C_6-204
Paraffins, vol-%	14.7	11.3	10.4	20.0	15.6	14.5
Olefins, vol-%	5.0	5.3	5.0	---	---	---
Naphthenes, vol-%	47.6	37.4	34.3	50.8	40.6	38.1
Aromatics, vol-%	32.7	46.0	50.3	29.2	43.8	47.4

Catalytic Reformer

As seen in the previous section, the naphtha feed to the catalytic reformer is extremely rich in high octane components and precursors (aromatics and naphthenes). Because of the characteristics of the naphtha (hydrotreated naphtha octane is about 85 RON), relatively mild reforming conditions can be used to make high octane products. Since the reaction conditions are very mild, high reformate liquid yields can be obtained:

	Reformate Yields, LV-%	
Hydrotreated Naphtha	94 RON	100 RON
C_6-177°C	93.8	92.8
C_6-193°C	95.5	93.7
C_6-204°C	96.2	94.3

Contrary to expectations implied by the high naphtha naphthene content, reformer hydrogen production is lower than that for a typical crude oil naphtha being upgraded to the same octane. Since reformate octane is achieved by only partial dehydrogenation of the naphtha, other hydrogen producing reactions are essentially not initiated. To achieve high hydrogen production this feedstock would probably have to be pushed to the range of 104-105 RON clear. Approximated hydrogen yields are given below:

Hydrotreated Naphtha	Hydrogen Yields, SCFB 94 RON	100 RON
C_6-177°C	1035	1120
C_6-193°C	700	880
C_6-204°C	610	780

These hydrogen yields are at variance with hydrogen yields obtained by others for H-Coal naphtha in theoretical and laboratory evaluations. This variance is primarily due to rather large differences in chemical composition observed in the samples taken from different periods of the H-Coal process development unit operation. (It must be cautioned that the catalytic reforming yield estimates presented here are theoretical and have not been verified by pilot plant testing.)

Distillate Hydrotreating

Upgrading of the coal liquid distillate to obtain a storage stable fuel requires very severe hydrotreating because of the raw distillate, high nitrogen concentration (4000 ppm). Other contaminants such as sulfur (0.1 wt-%) which is already acceptably low, and oxygen (1.0 wt-%) present no processing problems and are readily removed at the processing conditions required for denitrification. Although insufficient data are available, if diolefins are present to an extent greater than about 1.0 wt-% an additional low severity, low temperature hydrotreating section will be required for diolefin removal.

Denitrification of at least 90% is required of the H-Coal distillate in order to obtain a stable fuel oil. At this severity there is a significant saturation and cracking of the aromatic compounds, with a resultant product volumetric expansion and production of naphtha range and lighter components. Table 9 provides representative yield estimates for the upgrading to stable fuel oil of the distillate cuts considered in this study.

Hydrocracking

Because of the high distillate nitrogen content, a two-stage hydrocracking operation is required. The first stage removes most of the feed sulfur, oxygen and nitrogen, as well as saturates olefins. The first stage effluent is stripped of NH_3, H_2S, H_2O and light gases and is then charged to the second stage where most of the hydrocracking takes place. The charge to the hydrocracker is the 204-471°C effluent from the naphtha/distillate splitter column. Unconverted 343+°C material effluent from the hydrocracker second stage is recycled to extinction.

The processing conditions and catalyst selection for hydrocracking of H-Coal distillate result in reaction equilibria that favor more polyaromatic cracking with less individual aromatic saturation than hydrotreating the same material. This processing characteristic results in a greater net volumetric product benefit than is realized for coal distillate hydrotreating. Product yields and properties for various hydrocracking conversion levels are presented in Table 10.

FCC

In order to have an acceptable level of conversion in the FCC unit, the FCC charge stock must be severely hydrotreated. The heavy distillate cuts (260-471 and 288-471°C) selected as FCC charge are very refractory because of their aromatics concentration, and have a hydrogen content of only 8.7 and 8.3 wt-%, respectively. Conversion in the FCC of the untreated heavy distillate is only about 30 vol-% with poor overall product quality. Laboratory work has shown that severe hydrogenation of the heavy distillate to a hydrogen concentration of about 11.0 wt-% makes it suitable as a FCC feedstock. FCC conversion realized at this hydrogen content is about 60-65 vol-% to gasoline and lighter products. FCC feedstock hydrotreating yields and properties are included in Table 9. The FCC feedstock hydrogenation yields shown in this table indicate much less product hydrocracking, but more product hydrogen saturation. This fact is due to the selection of hydrogenation processing conditions that are significantly different from those employed for H-Coal distillate hydrotreating for fuel oil production.

A summary of the FCC yields and product properties is given below:

FCC Yields

Products	260-471°C Wt-%	260-471°C Vol-%	288-471°C Wt-%	288-471°C Vol-%
C_4-	12.65	---	13.5	---
C_5-400 Gasoline	36.70	43.3	38.20	45.8
Light Cycle Oil	38.15	36.3	31.45	29.9
Clarified Oil	5.40	4.5	9.50	8.0
Coke	7.10		7.30	
	100.00		100.00	

Product Properties

	260-471°C API	260-471°C RON	260-471°C MON	288-471°C API	288-471°C RON	288-471°C MON
Gasoline	47	95	83	47	95	83
Light Cycle Oil	12.5			10		
Clarified Oil	-6			-6		

Partial Oxidation

The partial oxidation option for hydrogen production has been incorporated as a means of disposing of the undesirable and poor quality 343-471°C heavy distillate stream recovered in the hydrotreating scheme. Only about 4.0 vol-% of the H-Coal liquid falls in this range, and it is best disposed of either as a refinery fuel or as a hydrogen source.

4.3.3 Processing Conclusions

While the processing schemes are not the same as in a crude oil refinery, upgrading of coal liquids to acceptable end products is certainly technically feasible. It must be pointed out, however, that the coal oil refinery products properties can be significantly different from crude oil refinery properties. For example, the No. 2 fuel oil is still very aromatic and would not meet heating oil gravity requirements without extensive further hydrogenation. This is even more true if the coal distillate is desired for use as a transportation fuel. Probably the coal oil distillate can best be blended with crude oil No. 2 fuel oil stocks after it has been hydrotreated, or it can be used directly, possibly with modified burner systems.

The raw untreated 190-340°C coal liquid distillate has an API gravity of about 15.0 and a aromatic and polar unsaturated compound content of over 85 wt-%. After hydrotreating for 90% denitrification and stabilization to produce a stable No. 2 fuel oil type material, the API gravity has been increased to about 23-24 while the aromatic concentration has been decreased to about 40-50 wt-%.

Even though the hydrotreated No. 2 fuel oil has a high energy content per unit volume, the question of its suitability for a wide variety of heating fuel uses would require extensive testing to be resolved. In general, highly aromatic fuels tend to burn with increased smoke and soot formation, and would have to be debited in value or hydrotreated for aromatics saturation to be considered as a replacement for petroleum derived No. 2 fuel oil. However, it is felt that the hydrotreated distillate could be used directly by many industrial users with little, if any, burner modifications. Therefore, in this study no debit has been assigned to the value of the coal liquid derived No. 2 fuel oil, even though it might have a limited market due to it properties.

Further upgrading of the already hydrotreated distillate to produce an acceptable 30 API product would require additional severe hydrotreating with a hydrogen consumption in the range of 1200 SCFB at an added processing cost of about $3/barrel. It would be directionally more attractive to hydrocrack the H-Coal distillate to gasoline than to try to hydrotreat it to meet No. 2 fuel oil gravity specifications for the following reasons:

1. Addition of sufficient hydrogen to the H-Coal distillate to produce a 30 API No. 2 fuel oil would consume a large quantity of energy relative to the improvement in product quality as compared to a partially upgraded and stable 23-24 API product.

2. Rather than saturate the aromatic structures present in the coal liquid distillate, it would be more desirable to crack the polyaromatics present in the distillate to mono-aromatics through the use of a selective hydrocracking catalyst. Hydrocracking would thus recover most of the cyclic structures present in the coal liquid without a greatly increased hydrogen consumption (about 3500 SCFB for complete conversion of the distillate to naphtha and lighter product vs. about 2800 SCFB required for saturation of the raw distillate to a 30 API product). The hydrocracked naphtha would be very valuable as a catalytic reformer chargestock because it is very rich in high octane precursors and components. The reformer product can then be used as a high octane blending component at a value significantly above that for No. 2 fuel oil.

Although the gasoline products produced in this study have been blended to meet unleaded regular and premium specifications, examination of the hydrotreated straight run coal liquid naphtha and of the hydrocracked coal liquid naphtha indicates that these naphthas could possibly be used to even greater benefit than just meeting the coal liquid refinery gasoline requirements. These naphthas are very rich in naphthenes and aromatics and, therefore, are excellent feedstocks for the production of a high octane gasoline blending component.

	193°C EP Hydrotreated SR Naphtha	190°C EP Hydrocracked Naphtha
Paraffins, vol-%	16.5	8
Naphthenes, vol-%	51.5	53
Aromatics, vol-%	32.0	39

These naphthas can be catalytically reformed to a high octane (105 RON clear) and would be very valuable as a gasoline blending component in other gasoline pools which are octane deficient. Also, a high severity of reforming is required to recover the large amount of hydrogen available in the naphthenes of the coal liquid naphthas.

4.4 ECONOMIC EVALUATION

4.4.1 Direct Operating Requirements

The crude oil refinery direct operating requirements (utilities, catalyst, chemicals and labor) are greater than those observed for the coal oil refinery. Utilities consumption for the crude oil refinery schemes is presented in Table 11, while those for the coal oil refinery schemes is presented in Table 12. Utility production and costs are as in the "Basis" of this study. All the components of direct operating requirements are included as operating costs in the overall economic comparison which will be discussed later in this section.

4.4.2 Investment Requirements

Capital costs are based upon 2nd quarter, 1979, East Coast erection. The capital costs provided for the process units in this study are curve type costs. These costs have been derived from data based on previous construction of process units similar to those incorporated in the study, and have been escalated to reflect current costs.

The costs developed for the refinery offsites are generalized and are based on relative refinery complexity. The relationships between refinery complexity and offsites costs are as outlined by W. L. Nelson.[1]

4.4.3 Economic Comparison

The overall refinery economics for the crude oil refinery schemes studied and the gasoline/distillate ratios considered, are presented in Tables 13-16. The coal oil refinery economics are presented in Tables 17-20. The refinery economics are based on the product

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(1) See Bibliography

prices set in the Basis of the study and upon a 30% cost of capital for the total refinery investment. The 30% cost of capital reflects a 15% after tax discounted cash flow rate of return on investment. Applying the same product prices to all cases, and subtracting operating costs and the capital charge, a comparative feedstock value is calculated. This method places the various refinery schemes on a common basis and gives an appraisal of the relative value of the H-Coal oil liquid charge stock based on new refining facilities. The cost of processing coal liquid in an existing refinery designed for petroleum would be relatively higher than one designed exclusively for processing coal liquid, since the processing facilities would not be used efficiently for upgrading coal liquids.

Referring to the flow schemes described earlier, optimum process unit throughputs, processing severities, and product blends were established by LP modelling for a range of gasoline/distillate ratios.

Petroleum Refinery

Of all of the petroleum refinery cases the Scheme IV option (including vacuum fractionation, solvent demetallizing, and hydrocracking) proved to be the most attractive because of a combination of higher revenue and lower operating costs than Schemes I and II. The economic comparison of these schemes is presented in Tables 13-16. The method of comparison provides a relative maximum acceptable charge stock value to achieve a 15% discounted cash flow rate of return. The relative charge stock values for the various schemes and product ratios for the petroleum refinery are given below:

	Charge Stock Value, $/Barrel		
Gasoline/Distillate Ratio	Scheme I	Scheme II	Scheme IV
0.7		16.91	17.35
1.0	17.44	17.38	17.60
1.5	17.78	17.58	17.82
2.0	17.89	17.62	17.89

The above charge stock values illustrate the relative attractiveness of each case.

Coal Liquid Refinery

The economic results for the coal liquid refinery cases are shown in Tables 17-20. The three basic schemes studied are summarized below at the four gasoline/distillate ratios investigated:

0.85 Gasoline/Distillate Ratio

None of the schemes could yield a 0.7 gasoline to distillate ratio, because of excessive gasoline production. Only the distillate hydrotreating scheme could achieve a ratio as low as 0.85.

1.0 Gasoline/Distillate Ratio

Both the distillate hydrotreating and hydrocracking schemes were feasible in this product range, but the hydrotreating scheme proved superior due to its significantly lower capital and operating costs. Not enough distillate can be produced in the FCC scheme to obtain this gasoline/distillate ratio.

1.5 Gasoline/Distillate Ratio

All three schemes were feasible in this range. The distillate hydrocracking scheme was somewhat superior to the hydrotreating scheme, due to significantly greater product revenues which more than offset the increased capital requirements. The FCC case is not viable because the heavy coal liquid distillate is not a suitable FCC feedstock and results in relatively low product values and high capital costs.

2.0 Gasoline/Distillate Ratio

The distillate hydrocracking scheme is clearly superior at this ratio, because the hydrotreating scheme cannot produce sufficient gasoline to meet the ratio. The FCC scheme is uneconomic as noted above due to relatively low product revenues and high capital costs.

A summary of the charge stock values for each scheme and product ratio is given below:

	Charge Stock Value, $/Barrel		
Gasoline to Distillate Ratio	Scheme A	Scheme B	Scheme C
0.85	20.21		
1.0	20.63	19.66	
1.5	19.91	20.21	18.53
2.0		20.23	18.93

A review of the study results shows that for the given parameters the processing of coal liquid is less expensive than the processing of petroleum, because of the reduced refinery complexity and investment required. In other words, a refiner can afford to pay about $2.00/ barrel more for the coal liquid than a Light/Heavy Arabian crude oil blend and still realize the same rate of return on investment for a new refinery specific to the feedstock processed.

The small proportion of coal liquid feedstock distilling higher than the product endpoint also is significant with respect to the relative value of petroleum and coal liquid. If the proportion of such residue increases, the relative value of coal liquid would decrease due to higher processing cost to convert heavy oil to lighter products.

It should again be pointed out that the results of this study are based upon "curve-type" numbers and generalized correlations. However, the study results should provide a good indication of the relative processing requirements, costs, and product slates associated with each scheme investigated.

This study has quantified the differences in values of a typical crude oil and the H-Coal liquid product, based on respective grass roots refining facilities. These techniques can be applied to other coal liquid feeds or crude oil processing schemes to establish the relative attractiveness of the many possible options for increasing the availability of premium hydrocarbon fuels. An evaluation is underway of coal liquids processing in a "typical" existing petroleum refinery, and results will be reported in the future.

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TABLE 1

Petroleum Refinery
Product Slates and Process Units
0.7 Gasoline/Distillate Ratio

Products, BPCD	RCD + FCC Scheme 1	RCD + HC Scheme II	Deasph. + FCC Scheme III	Deasph. + HC Scheme IV
LPG	Not feasible due to excessive production of gasoline	4,321	Not feasible due to production of excessive pitch	8,024
Unleaded Premium		10,731		10,916
Unleaded Regular		25,038		25,470
No. 2 Fuel Oil		51,099		51,978
No. 6 Fuel Oil		5,087		-
Sulfur (MTD)		(274)		(274)
		96,276		96,388
Process Unit Throughputs, BPCD (MTD)				
Crude		100,000		100,000
Vacuum		-		47,429
Solvent Demetallization		-		17,952
Reduced Crude Desulf.		47,339		-
Hydrocracker		34,907		39,637
FCC		-		-
Naphtha Hydrotreater		36,043		34,092
Catalytic Reformer		30,888		28,980
Kerosine Treater		11,304		17,047
Isomerization		5,117		5,117
HF Alkylation(1)		-		-
Polymerization(1)		-		-
FCC Gasoline Treater		-		-
AGO Hydrotreater		14,606		14,606
Flue Gas SO_x Removal (SO_x, MTD)		-		(105)
Saturate Gas Recovery		(1,519)		(1,661)
Sulfur Plant		(274)		(274)
Hydrogen Plant		(121)		-
Partial Oxidation		-		(108)

(1) Based on C_5+ product

TABLE 2

Petroleum Refinery
Product Slates and Process Units
1.0 Gasoline/Distillate Ratio

Products, BPCD	Scheme I	Scheme II	Scheme III	Scheme IV
LPG	750	6,726		9,693
Unleaded Premium	12,369	12,888	Not feasible	13,151
Unleaded Regular	28,860	30,072	due to	30,686
No. 2 Fuel Oil	41,229	42,960	production	43,838
No. 6 Fuel Oil	13,863	5,114	of excessive	-
Sulfur (MTD)	(264)	(274)	pitch	(274)
	97,071	97,760		97,368
Process Unit Throughputs, BPCD (MTD)				
Crude	100,000	100,000		100,000
Vacuum	-	-		47,429
Solvent Demetallization	-	-		17,952
Reduced Crude Desulf.	47,429	47,339		-
Hydrocracker	-	34,905		39,637
FCC	34,673	-		-
Naphtha Hydrotreater	20,279	40,474		39,651
Catalytic Reformer	15,138	35,318		34,537
Kerosine Treater	17,047	11,304		17,047
Isomerization	5,117	5,117		5,117
HF Alkylation(1)	3,381	-		-
Polymerization(1)	2,258	-		-
FCC Gasoline Treater	17,525	-		-
AGO Hydrotreater	14,606	14,606		14,606
Flue Gas SO_x Removal (SO_x, MTD)	-	-		101
Saturate Gas Recovery	(1,161)	(1,765)		(1,904)
Sulfur Plant	(264)	(274)		(274)
Hydrogen Plant	(60)	(124)		-
Partial Oxidation	-	-		(116)

(1) Based on C_5+ Product

TABLE 3

Petroleum Refinery
Product Slates and Process Units
1.5 Gasoline/Distillate Ratio

Products, BPCD	Scheme I	Scheme II	Scheme III	Scheme IV
LPG	972	8,568		9,844
Unleaded Premium	15,072	15,355	Not feasible due	15,749
Unleaded Regular	35,167	35,828	to production	36,747
No. 2 Fuel Oil	33,492	34,122	of excessive	34,997
No. 6 Fuel Oil	13,078	4,549	pitch	-
Sulfur (MTD)	(267)	(274)		(271)
	97,781	98,422		97,337
Process Unit Throughputs, BPCD (MTD)				
Crude	100,000	100,000		100,000
Vacuum	-	-		47,429
Solvent Demetallization	-	-		17,952
Reduced Crude Desulf.	47,429	47,339		-
Hydrocracker	-	34,909		39,562
FCC	37,703	-		-
Naphtha Hydrotreater	20,291	46,329		47,523
Catalytic Reformer	15,177	41,281		42,407
Kerosine Treater	17,047	11,304		11,304
Isomerization	5,117	5,117		5,117
HF Alkylation(1)	4,461	-		-
Polymerization(1)	2,788	-		-
FCC Gasoline Treater	24,626	-		-
AGO Hydrotreater	14,606	14,606		14,606
Flue Gas SO_x Removal (SO_x, MTD)	-	-		(106)
Saturate Gas Recovery	(1,180)	(2,026)		(2,026)
Sulfur Plant	(267)	(274)		(271)
Hydrogen Plant	(61)	(132)		
Partial Oxidation	-	-		(105)

(1) Based on C_5+ product

TABLE 4

Petroleum Refinery
Product Slates and Process Units
2.0 Gasoline/Distillate Ratio

Products, BPCD	Scheme I	Scheme II	Scheme III	Scheme IV
LPG	1,194	9,553		10,966
Unleaded Premium	16,752	16,732	Not feasible due	17,314
Unleaded Regular	39,088	39,041	to production	40,399
No. 2 Fuel Oil	27,920	27,887	of excessive	28,856
No. 6 Fuel Oil	13,138	5,551	pitch	368
Sulfur (MTD)	(268)	(274)		(272)
	98,092	98,764		97,903
Process Unit Throughputs, BPCD (MTD)				
Crude	100,000	100,000		100,000
Vacuum	-	-		47,339
Solvent Demetallization	-	-		17,918
Reduced Crude Desulf.	47,429	47,339		-
Hydrocracker	-	34,614		39,562
FCC	44,578	-		-
Naphtha Hydrotreater	20,309	49,740		51,394
Catalytic Reformer	15,192	44,623		46,271
Kerosine Treater	17,047	11,304		11,304
Isomerization	5,117	5,117		5,117
HF Alkylation(1)	5,178	-		-
Polymerization(1)	3,268	-		-
FCC Gasoline Treater	29,027	-		-
AGO Hydrotreater	14,606	14,606		14,606
Flue Gas SO_x Removal (SO_x, MTD)	-	-		(127)
Saturate Gas Recovery	(1,212)	(2,159)		(2,190)
Sulfur Plant	(268)	(274)		(272)
Hydrogen Plant	(62)	(132)		(49)
Partial Oxidation	-	-		(64)

(1) Based on C_5+ product

TABLE 5

Coal Liquid Refinery
Product Slates and Process Units
Minimum Gasoline/Distillate Ratio

	Scheme A*	Scheme B	Scheme C
Products, BPCD			
LPG	2,992		
Unleaded Regular	28,826	Not	Not
Unleaded Premium	12,372	feasible	feasible
No. 2 Fuel Oil	48,517	below 1.0	at 1.0
Sulfur (MTD)	(16)	ratio	ratio or
	92,707		below
Process Unit Throughputs, BPCD (MTD)			
Primary Fractionator	100,000		
C_4/LSR Splitter	1,976		
Naphtha/Lt. Distillate Splitter	86,136		
Lt. Distillate/Hvy. Dist. Splitter	51,465		
C_4-C_6 Hydrotreater	12,330		
Naphtha Hydrotreater	30,862		
Catalytic Reformer	28,694		
Distillate Hydrotreater	51,465		
FCC Feed Hydrotreater	-		
FCC	-		
Hydrocracker	-		
HF Alkylation**	-		
Hydrogen Plant	(163)		
Partial Oxidation	-		
Sulfur Plant	(16)		
Saturate Gas Recovery	(311)		
FCC Gasoline Treater	-		

* Approximate gasoline/distillate ratio = .85
** Based on alkylate product

TABLE 6

Coal Liquid Refinery
Product Slates and Process Units
1.0 Gasoline/Distillate Ratio

	Scheme A	Scheme B	Scheme C
Products, BPCD			
LPG	1,859	3,975	Not feasible at 1.0 gasoline/distillate ratio
Unleaded Regular	32,025	32,068	
Unleaded Premium	13,744	13,763	
No. 2 Fuel Oil	45,815	45,877	
Sulfur (MTD)	(17)	(19)	
	93,443	95,683	
Process Unit Throughputs, BPCD (MTD)			
Primary Fractionator	98,756	93,971	
C_4/LSR Splitter	-	-	
Naphtha/Lt. Dist. Splitter	84,090	80,944	
Lt. Dist./Hvy. Dist. Split.	51,685	-	
C_4-C_6 Hydrotreater	13,698	13,035	
Naphtha Hydrotreater	32,407	29,001	
Catalytic Reformer	30,161	26,964	
Distillate Hydrotreater	47,971	9,775	
FCC Feed Hydrotreater	-	-	
FCC	-	-	
Hydrocracker	-	42,171	
HF Alkylation**	-	-	
Hydrogen Plant	(157)	(154)	
Partial Oxidation	-	-	
Sulfur Plant	(17)	(19)	
Saturate Gas Recovery	(319)	(461)	
FCC Gasoline Treater	-	-	

** Based on alkylate product

TABLE 7

Coal Liquid Refinery
Product Slates and Process Units
1.5 Gasoline/Distillate Ratio

	Scheme A	Scheme B	Scheme C
Products, BPCD			
LPG	1,979	4,174	5,631
Unleaded Regular	38,587	38,907	37,041
Unleaded Premium	16,561	16,698	15,897
No. 2 Fuel Oil	36,802	37,107	35,327
Sulfur (MTD)	(19)	(19)	(19)
	93,929	96,886	93,896
Process Unit Throughputs, BPCD (MTD)			
Primary Fractionator	100,000	100,000	97,920
C_4/LSR Splitter	615	-	13,691
Naphtha/Lt. Dist. Splitter	86,136	86,136	84,345
Lt. Dist./Hvy. Dist. Split.	45,171	-	53,647
C_4-C_6 Hydrotreater	13,871	13,871	13,582
Naphtha Hydrotreater	40,955	30,862	30,703
Catalytic Reformer	38,315	28,694	28,553
Distillate Hydrotreater	41,208	3,339	31,130
FCC Feed Hydrotreater	-	-	22,517
FCC	-	-	24,339
Hydrocracker	-	46,435	-
HF Alkylation**	-	-	2,877
Hydrogen Plant	(46)	(171)	(187)
Partial Oxidation	(118)	-	-
Sulfur Plant	(19)	(19)	(19)
Saturate Gas Recovery	(295)	(569)	(626)
FCC Gasoline Treater	-	-	10,531
LPG Treater			3,856

** Based on Alkylate Product

TABLE 8

Coal Liquid Refinery
Product Slates and Process Units
2.0 Gasoline/Distillate Ratio

	Scheme A	Scheme B	Scheme C
Products, BPCD			
LPG	Not feasible	5,596	3,802
Unleaded Regular	at high	43,002	42,145
Unleaded Premium	gasoline/	18,456	18,088
No. 2 Fuel Oil	distillate	30,760	30,147
Sulfur (MTD)	ratios	(19)	(19)
		97,814	94,182
Process Unit Throughputs, BPCD (MTD)			
Primary Fractionator		100,000	100,000
C_4/LSR Splitter		-	11,288
Naphtha/Lt. Dist. Splitter		86,136	86,136
Lt. Dist./Hvy. Dist. Splitter		-	47,253
C_4-C_6 Hydrotreater		13,871	13,871
Naphtha Hydrotreater		30,862	38,880
Catalytic Reformer		28,694	36,254
Distillate Hydrotreater		3,075	28,741
FCC Feed Hydrotreater		-	18,512
FCC		-	20,006
Hydrocracker		46,494	-
HF Alkylation**		-	2,453
Hydrogen Plant		(207)	(191)
Partial Oxidation		-	-
Sulfur Plant		(19)	(19)
Saturate Gas Recovery		(745)	(534)
FCC Gasoline Treater		-	8,796
LPG Treater			3,265

** Based on Alkylate Product

TABLE 9

Yields and Properties of Hydrotreated H-Coal Distillate
(Yields based on Charge)

			Product Yields, LV%						
Cut, °C	Charge API	C_4-	C_5/C_6	C_6+	C_7-177°C	C_7-204°C	177+°C	204+°C	Total C_5+ Liquid
177-204	12.2	N/A	1.01	101.40					102.41
177-343	16.5	N/A	3.17		6.33		94.24		103.74
204-260	19.0	N/A	1.46			2.90		101.55	105.91
204-288	16.5	N/A	1.59			3.12		101.39	106.10
204-343	15.1	N/A	1.60			3.15		101.91	106.66
204-471	13.6	N/A	1.78			3.50		101.33	106.61
260-471*	9.1	N/A	0.15			0.26		108.09	108.35
288-471*	6.9	N/A	0.18			0.33		108.01	108.34

Cut, °C	Product,	API	H_2 Consumption, SCFB
177-204	(C_6+)	28.1	910
177-343	(177+°C)	23.3	1650
204-260	(204+°C)	26.6	1600
204-288	"	24.1	1650
204-343	"	23.4	1650
204-471	"	22.0	1690
260-471*	"	19.9	1740
288-471*	"	17.4	2070

Cut, °C	Chargestock Aromatics(1)	Chargestock PNA(2)	Product	Product Aromatics(1)	Product PNA(2)
177-204	96 wt.%	0	C_6+	90 wt.%	0
177-343	87		177+°C		
204-260	85		204+°C		
204-288	85		204+°C		
204-343	85		204+°C		
204-471	86	400 wt. ppm	204+°C		

* FCC Chargestock

(1) Includes polar type aromatic compounds (aniline, pyridine, etc.).
(2) PNA - specifically carcinogenic poly-nuclear aromatics.

TABLE 10

H-Coal Distillate Hydrocracking Yields and Product Properties (Yields based on charge)

Chargestock Properties	Cut, °C	API	Nitrogen, ppm	Sulfur, wt.%	Oxygen, wt.%
	204-471	13.6	5,000	0.11	1.0

Conversion Cases Based Vol.% C_5+ Naphtha Produced	C_4-	Product Yields, LV% C_5/C_6	C_7/190°C	190-343°C	Total C_5+
20%	N/A	6.01	13.99	85.40	105.40
40%	N/A	11.82	28.18	65.06	105.06
60%	N/A	17.64	42.36	44.68	104.68

Conversion Cases	Total C_5+ Wt.%	H_2 Consumption, SCFB
20%	95.94	1650
40%	93.14	2120
60%	90.32	2590

C_5/C_6	Product Properties API	RON-0
All Conversions	63	93

C_7/190°C	API	RON-0	Paraffins, vol.%	Naphthenes, vol.%	Aromat., vol.%
All Conversions	40.8	90	8	53	39

190-343°C	API	Pour Pt., °C
All Conversions	24.3	-42

TABLE 11

Petroleum Refinery

Petroleum Refinery Utilities

0.7 Gasoline/Distillate Cases	Scheme I	Scheme II	Scheme III	Scheme IV
Power, KWH	Infeasible due to excessive production of gasoline	31,100	Infeasible due to excessive pitch production	30,650
High Pressure Steam, MT/hr.		13		0
Medium Pressure Steam, MT/hr.		-23		-101
Low Pressure Steam, MT/hr.		38		79
Fuels, 10^6 KCal/hr.		480		480
Boiler Feedwater, MT/hr.		105		87
Cooling Water, M^3/hr.		2,640		4,820
1.0 Gasoline/Distillate Cases				
Power, KWH	26,600	31,000	Infeasible due to excessive pitch production	32,400
High Pressure Steam, MT/hr.	9	1.3		0
Medium Pressure Steam, MT/hr.	-22	-24		-4.2
Low Pressure Steam, MT/hr.	38	42		80
Fuels, 10^6 KCal/hr.	300	470		500
Boiler Feedwater, MT/hr.	94	102		91
Cooling Water, M^3/hr.	3,880	2,870		5,280
1.5 Gasoline/Distillate Cases				
Power, KWH	27,250	32,600	Infeasible due to excessive pitch production	22,000
High Pressure Steam, MT/hr.	0	12.6		0
Medium Pressure Steam, MT/hr.	-23	-24		-106
Low Pressure Steam, MT/hr.	44	52		105
Fuels, 10^6 KCal./hr.	310	500		560
Boiler Feedwater, MT/hr.	96	104		88
Cooling Water, M^3/hr.	4,410	3,070		3,250
2.0 Gasoline/Distillate Cases				
Power, KWH	27,500	33,300	Infeasible due to excessive pitch production	27,600
High Pressure Steam, MT/hr.	0	0		0
Medium Pressure Steam, MT/hr.	-20	-24		-2.2
Low Pressure Steam, MT/hr.	47	58		95
Fuels, 10^6 KCal/hr.	320	530		560
Boiler Feedwater, MT/hr.	99	107		89
Cooling Water, M^3/hr.	4,750	3,260		4,420

TABLE 12

Coal Liquid Refinery Utilities

Minimum Gasoline/Distillate Cases	Scheme A*	Scheme B	Scheme C
Power, KWH	22,700	-	-
High Pressure Steam, MT/hr.	86	-	-
Medium Pressure Steam, MT/hr.	1.5	-	-
Low Pressure Steam, MT/hr.	10	-	-
Fuels, 10^6 KCal/hr.	380	-	-
Boiler Feed Water, MT/hr.	65	-	-
Cooling Water, M^3/hr.	1,320	-	-
1.0 Gasoline/Distillate Cases			
Power, KWH	22,100	33,400	-
High Pressure Steam, MT/hr.	90	62	-
Medium Pressure Steam, MT/hr.	1.5	-28	-
Low Pressure Steam, MT/hr.	10	24	-
Fuels, 10^6 KCal/hr.	380	350	-
Boiler Feed Water, MT/hr.	63	69	-
Cooling Water, M^3/hr.	1,300	1,920	-
1.5 Gasoline/Distillate Cases			
Power, KWH	34,700	34,900	25,800
High Pressure Steam, MT/hr.	110	48	62
Medium Pressure Steam, MT/hr.	2	-31	-11
Low Pressure Steam, MT/hr.	11	28	23
Fuels, 10^6 KCal/hr.	360	360	360
Boiler Feed Water, MT/hr.	74	72	91
Cooling Water, M^3/hr.	4,090	2,050	2,580
2.0 Gasoline/Distillate Cases			
Power, KWH	-	38,100	24,300
High Pressure Steam, MT/hr.	-	34	88
Medium Pressure Steam, MT/hr.	-	-31	-10
Low Pressure Steam, MT/hr.	-	37	36
Fuels, 10^6 KCal/hr.	-	380	360
Boiler Feed Water, MT/hr.	-	85	87
Cooling Water, M^3/hr.	-	2,270	2,770

* .85 Gasoline/Distillate Ratio

TABLE 13
Petroleum Refinery
Economic Comparison
0.7 Gasoline/Distillate Ratio

	$MM/Year	
	Scheme II	Scheme IV
Total Investment		
Plant Investment	342.0	323.0
Allowance for Offsite Investment	400.0	420.0
Working Capital	110.0	110.0
Initial Catalyst & Chemicals	10.5	7.7
Additional Costs @ 10% of Plant + Offsites	74.2	74.3
	936.7	935.0
Sales	956.5	957.7
Operating Costs		
Labor, Catalyst, Chemicals, and Utility Purchase Costs	19.8	5.7
Maintenance, Property Taxes & Insurance @ 4-1/2% of Plant & Offsite Investment	33.4	33.4
	53.2	39.1
General & Administrative Costs	5.0	5.0
Capital Charges @ 30% of Total Investment	281.1	280.5
Raw Materials Value	617.2	633.1
Raw Materials Value ($/Barrel)	16.91	17.35

TABLE 14
Petroleum Refinery
Economic Comparison
1.0 Gasoline/Distillate Ratio

	$MM/Year		
	Scheme I	Scheme II	Scheme IV
Total Investment			
Plant Investment	303.0	341.0	335.0
Allowance for Offsite Investment	400.0	400.0	420.0
Working Capital	110.0	110.0	110.0
Initial Catalyst & Chemicals	5.2	9.0	4.3
Additional Costs @ 10% of Plant + Offsites	70.3	74.1	75.5
	888.5	934.1	944.8
Sales	958.8	972.4	971.3
Operating Costs			
Labor, Catalyst, Chemicals, and Utility Purchase Costs	19.1	19.6	6.1
Maintenance, Property Taxes & Insurance @ 4-1/2% of Plant & Offsite Investment	31.7	33.3	34.0
	50.8	52.9	40.1
General & Administrative Costs	5.0	5.0	5.0
Capital Charges @ 30% of Total Investment	266.6	280.2	283.4
Raw Materials Value	636.4	634.3	642.5
Raw Materials Value ($/Barrel)	17.44	17.38	17.60

TABLE 15
Petroleum Refinery
Economic Comparison
1.5 Gasoline/Distillate Ratio

	$MM/Year		
	Scheme I	Scheme II	Scheme IV
Total Investment			
Plant Investment	314.0	352.0	336.0
Allowance for Offsite Investment	400.0	400.0	420.0
Working Capital	110.0	110.0	110.0
Initial Catalyst & Chemicals	5.5	9.4	4.4
Additional Costs @ 10% of Plant + Offsites	71.4	75.2	75.6
	900.9	946.6	946.0
Sales	976.3	984.5	979.2
Operating Costs			
Labor, Catalyst, Chemicals, and Utility Purchase Costs	19.9	19.8	6.0
Maintenance, Property Taxes & Insurance @ 4-1/2% of Plant & Offsite Investment	32.1	33.9	34.0
	52.0	53.7	40.0
General & Administrative Costs	5.0	5.0	5.0
Capital Charges @ 30% of Total Investment	270.3	284.0	283.8
Raw Materials Value	649.0	641.8	650.4
Raw Materials Value ($/Barrel)	17.78	17.58	17.82

TABLE 16
Petroleum Refinery
Economic Comparison
2.0 Gasoline/Distillate Ratio

	$MM/Year		
	Scheme I	Scheme II	Scheme IV
Total Investment			
Plant Investment	322.0	359.0	350.0
Allowance for Offsite Investment	400.0	400.0	420.0
Working Capital	110.0	110.0	110.0
Initial Catalyst & Chemicals	5.6	9.8	4.9
Additional Costs @ 10% of Plant + Offsites	72.2	75.9	77.0
	909.8	954.7	961.9
Sales	983.9	988.4	986.8
Operating Costs			
Labor, Catalyst, Chemicals, and Utility Purchase Costs	20.6	19.5	5.7
Maintenance, Property Taxes & Insurance @ 4-1/2% of Plant & Offsite Investment	32.5	34.2	34.7
	53.1	53.7	40.4
General & Administrative Costs	5.0	5.0	5.0
Capital Charges @ 30% of Total Investment	272.9	286.4	288.6
Raw Materials Value	652.9	643.3	652.8
Raw Materials Value ($/Barrel)	17.89	17.62	17.89

TABLE 17

Coal Liquid Refinery
Economic Comparison
0.85 Gasoline/Distillate Ratio

	$MM/Year Scheme A
Total Investment	
Plant Investment	160.0
Allowance for Offsite Investment	220.0
Working Capital	120.0
Initial Catalyst & Chemicals	7.5
Addit. Costs @ 10% of Plant + Offsites	38.0
	545.5
Sales	937.9
Operating Costs	
Labor, Catalyst, Chemicals and Utility Purchase Costs	14.5
Maintenance, Property Taxes and Insurance @ 4 1/2% of Plant and Offsite Investment	17.1
	31.6
General & Administrative	5.0
Capital Charges @ 30% of Total Investment	163.7
Raw Materials Value	737.6
Raw Materials Value ($/Barrel)	20.21

TABLE 18

Coal Liquid Refinery
Economic Comparison
1.0 Gasoline/Distillate Ratio

	$MM/Year	
	Scheme A	Scheme B
Total Investment		
Plant Investment	162.0	233.0
Allowance for Offsite Investment	220.0	280.0
Working Capital	120.0	120.0
Initial Catalyst & Chemicals	7.7	8.1
Addit. Costs @ 10% of Plant + Offsites	38.2	51.3
	547.9	692.4
Sales	953.8	969.5
Operating Costs		
Labor, Catalyst, Chemicals and Utility Purchase Costs	14.2	16.2
Maintenance, Property Taxes and Insurance @ 4 1/2% of Plant and Offsite Investment	17.2	23.1
	31.4	39.3
General & Administrative	5.0	5.0
Capital Charges @ 30% of Total Investment	164.4	207.7
Raw Materials Value	753.0	717.5
Raw Materials Value ($/Barrel)	20.63	19.66

TABLE 19

Coal Liquid Refinery
Economic Comparison
1.5 Gasoline/Distillate Ratio

	$MM/Year		
	Scheme A	Scheme B	Scheme C
Total Investment			
Plant Investment	210.0	239.0	268.0
Allowance for Offsite Investment	270.0	280.0	320.0
Working Capital	120.0	120.0	120.0
Initial Catalyst & Chemicals	8.9	8.0	3.8
Addit. Costs @ 10% of Plant + Offsites	48.0	51.9	58.8
	656.9	698.9	770.6
Sales	969.5	992.3	955.9
Operating Costs			
Labor, Catalyst, Chemicals and Utility Purchase Costs	19.1	16.4	16.7
Maintenance, Property Taxes and Insurance @ 4 1/2% of Plant and Offsite Investment	21.6	23.4	26.5
	40.7	39.8	43.2
General & Administrative	5.0	5.0	5.0
Capital Charges @ 30% of Total Investment	197.1	209.7	231.2
Raw Materials Value	726.7	737.8	676.5
Raw Materials Value ($/Barrel)	19.91	20.21	18.53

TABLE 20

Coal Liquid Refinery
Economic Comparison
2.0 Gasoline/Distillate Ratio

	$MM/Year	
	Scheme B	Scheme C
Total Investment		
Plant Investment	251.0	264.0
Allowance for Offsite Investment	295.0	330.0
Working Capital	120.0	120.0
Initial Catalyst & Chemicals	8.6	4.6
Addit. Costs @ 10% of Plant + Offsites	54.6	59.4
	729.2	778.0
Sales	1,004.1	972.8
Operating Costs		
Labor, Catalyst, Chemicals and Utility Purchase Costs	17.2	16.7
Maintenance, Property Taxes and Insurance @ 4 1/2% of Plant and Offsite Investment	24.6	26.7
	41.8	43.4
General & Administrative	5.0	5.0
Capital Charges @ 30% of Total Investment	218.8	233.4
Raw Materials Value	738.5	691.0
Raw Materials Value ($/Barrel)	20.23	18.93

FIGURE 1
(SCHEME I)

CRUDE OIL REFINERY

REDUCED CRUDE DESULFURIZATION & FCC

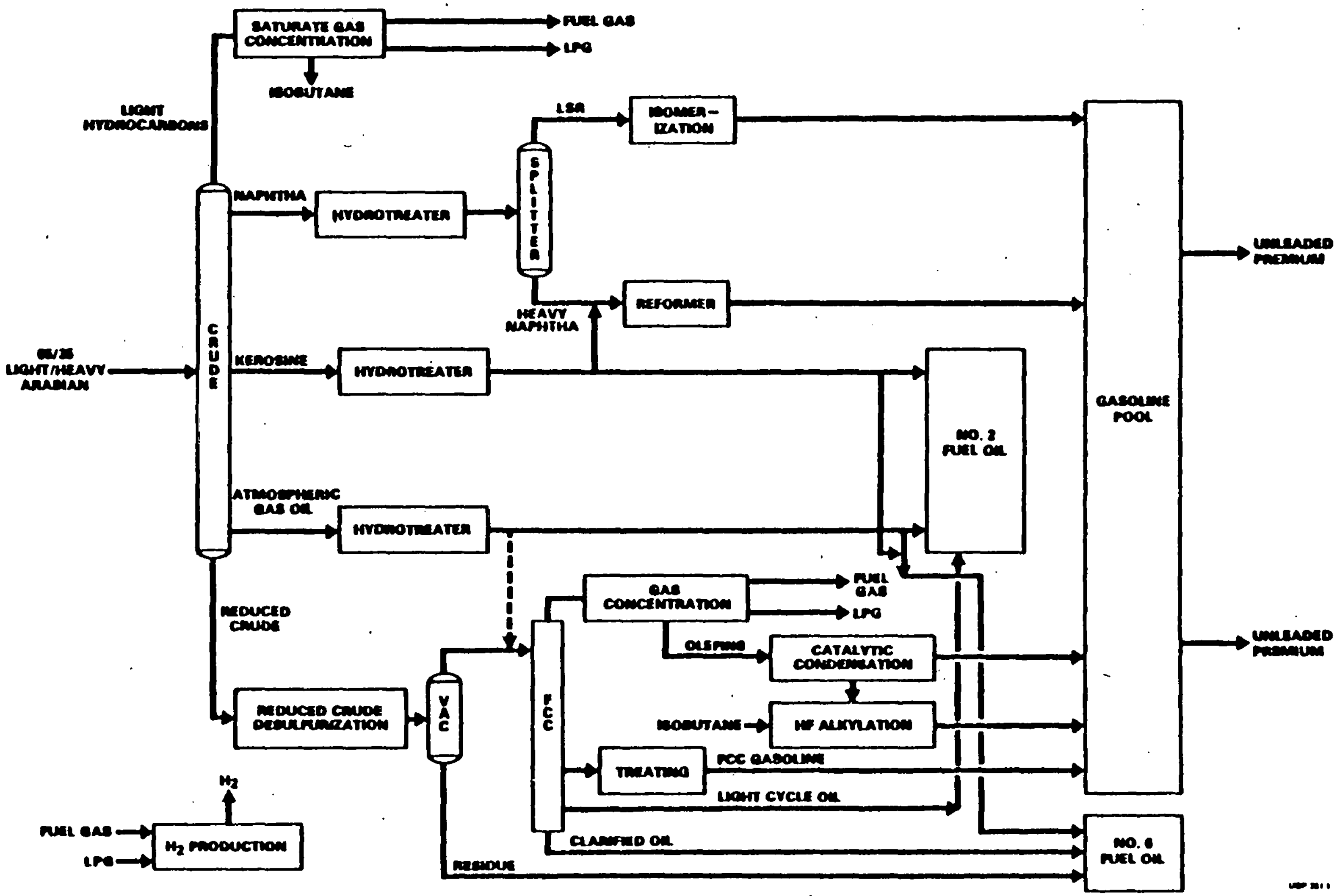

FIGURE 2
(SCHEME II)

CRUDE OIL REFINERY

REDUCED CRUDE DESULFURIZATION & HYDROCRACKING

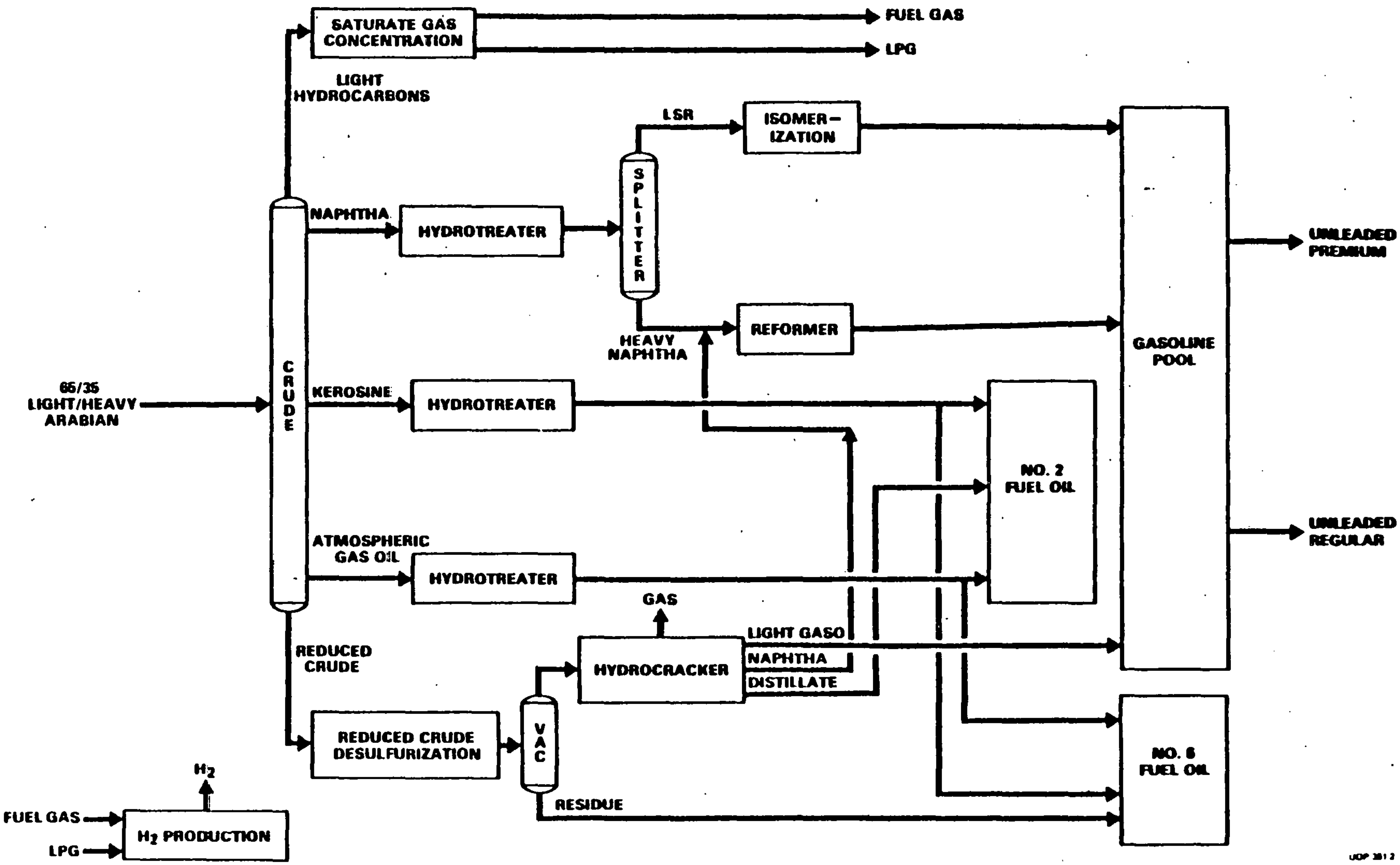

FIGURE 3
(SCHEME III)

CRUDE OIL REFINERY

FCC & VACUUM RESIDUE DEMETALLIZATION

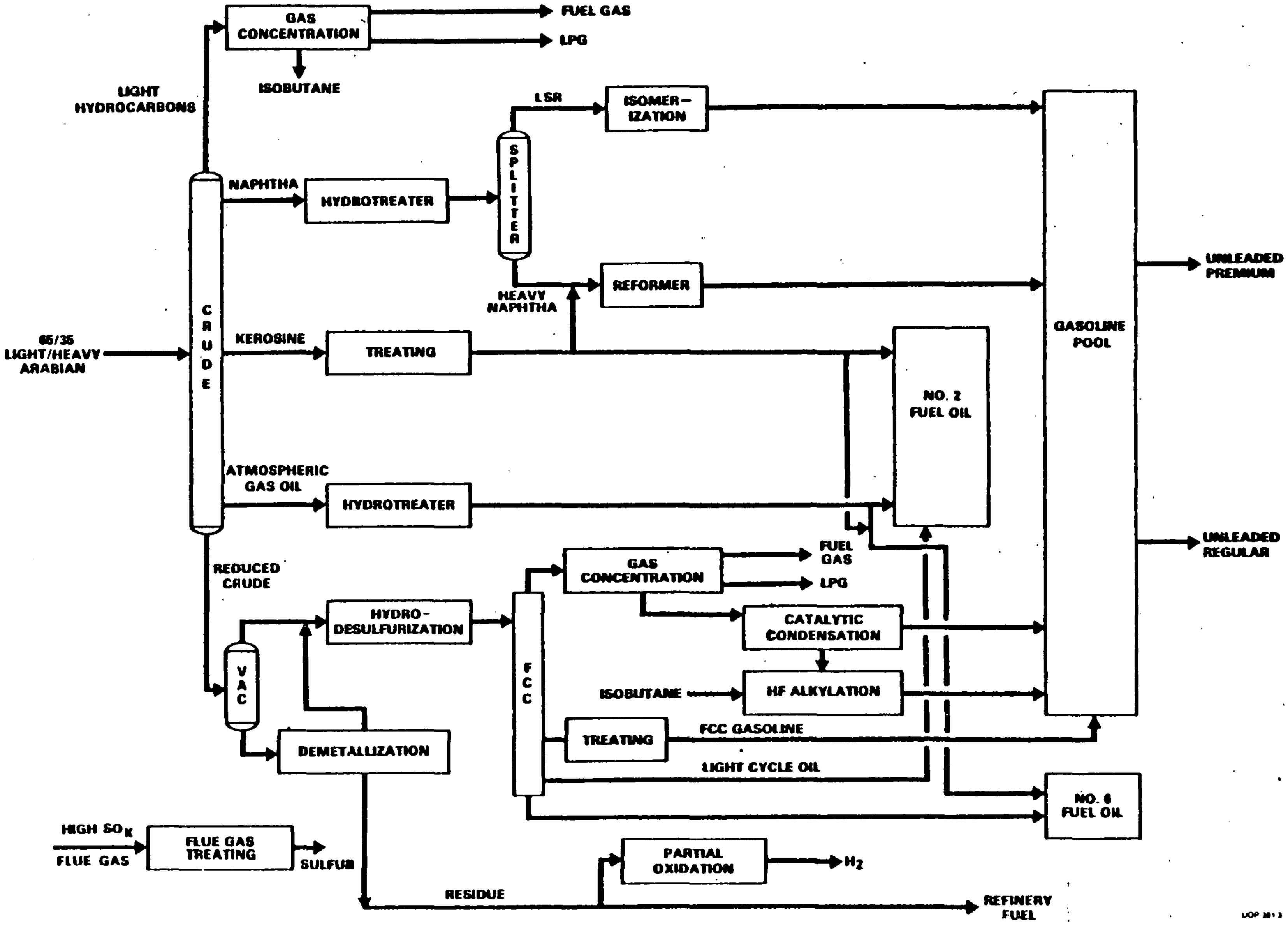

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FIGURE 4
(SCHEME IV)

CRUDE OIL REFINERY

HYDROCRACKING & VACUUM RESIDUE DEMETALLIZATION

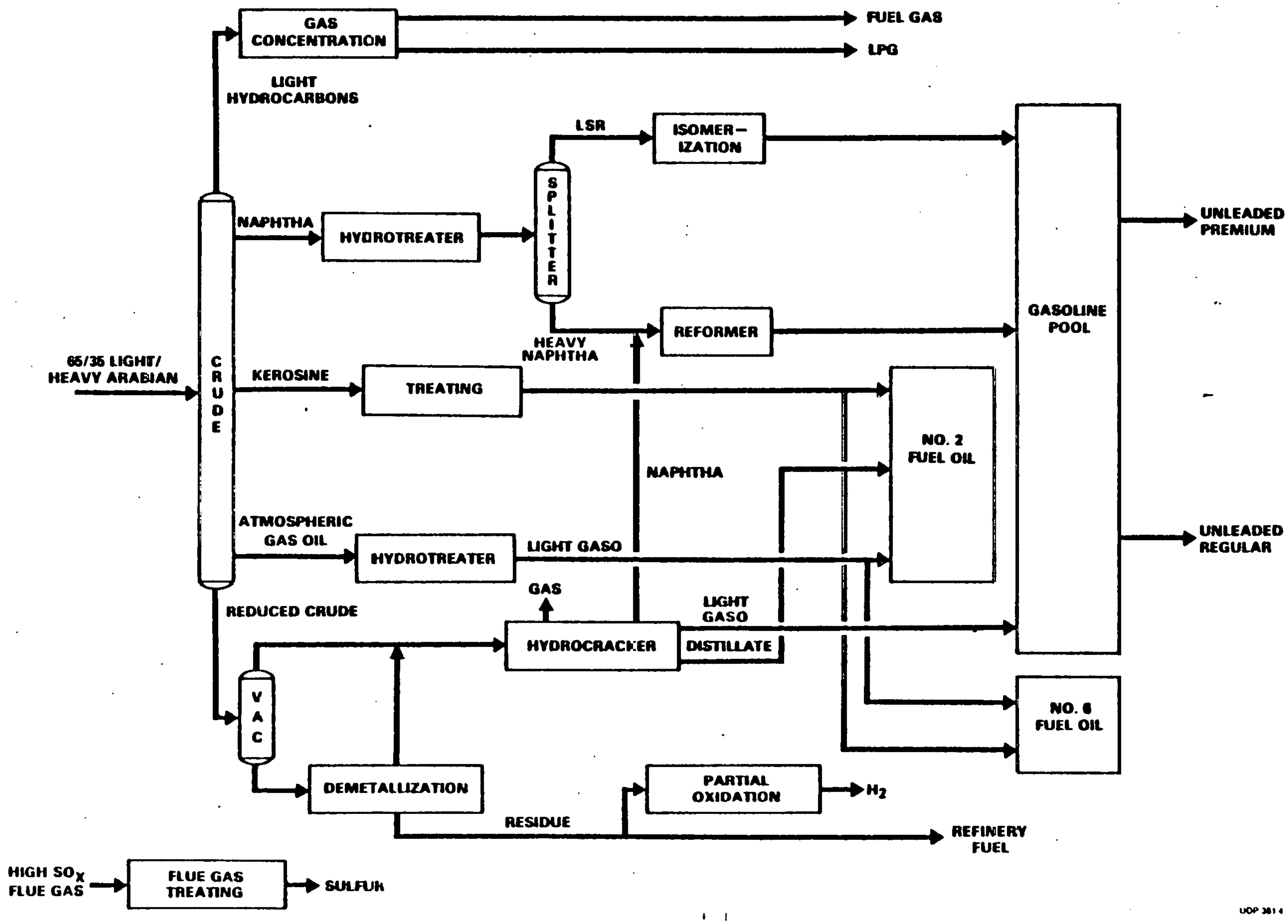

FIGURE 5
(SCHEME A)

COAL OIL REFINERY

HYDROTREATING CASES WITH LOW TEMPERATURE FRACTIONATION

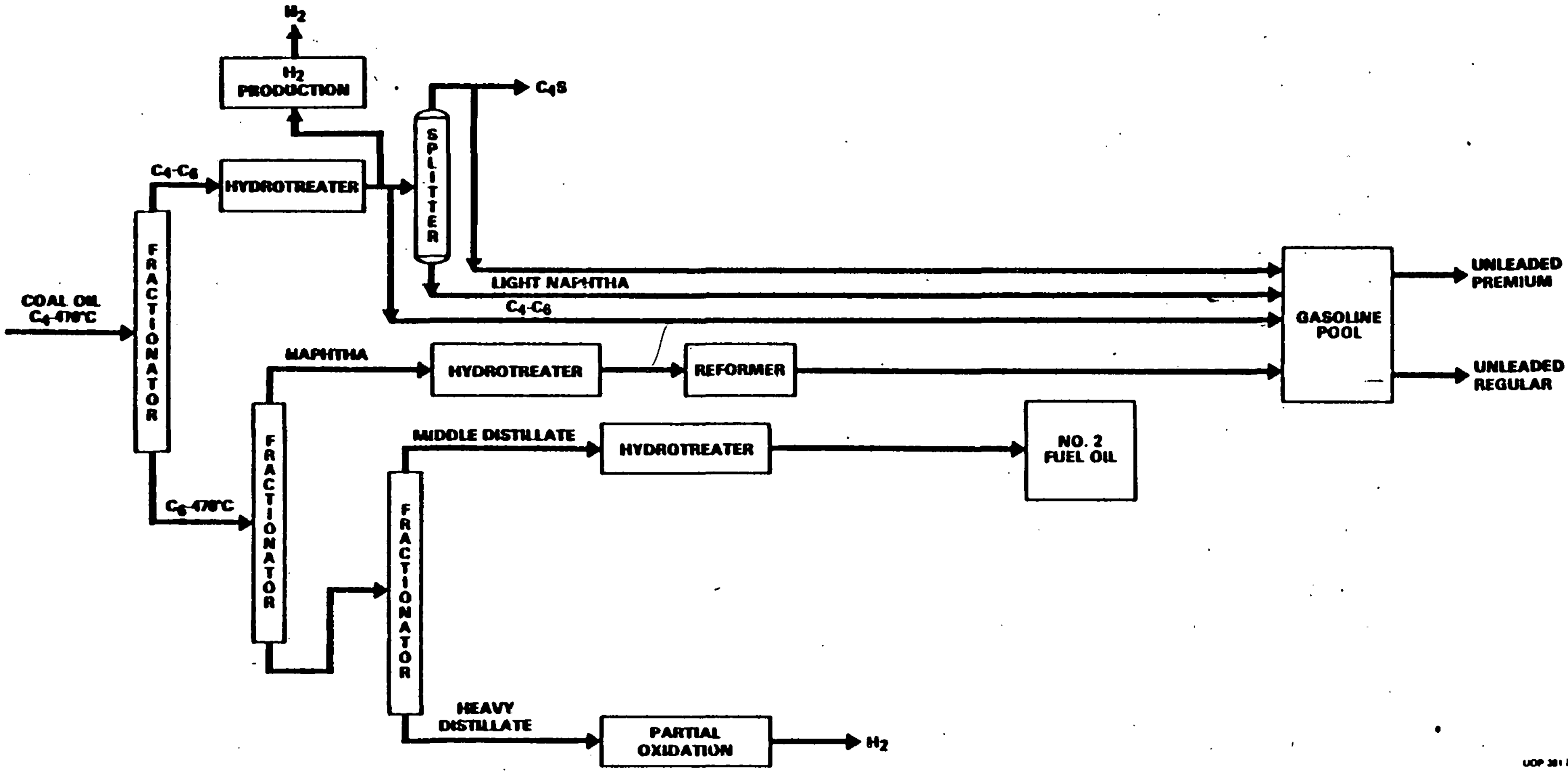

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FIGURE 6
(SCHEME B)

COAL OIL REFINERY

HYDROCRACKING CASES WITH LOW TEMPERATURE FRACTIONATION

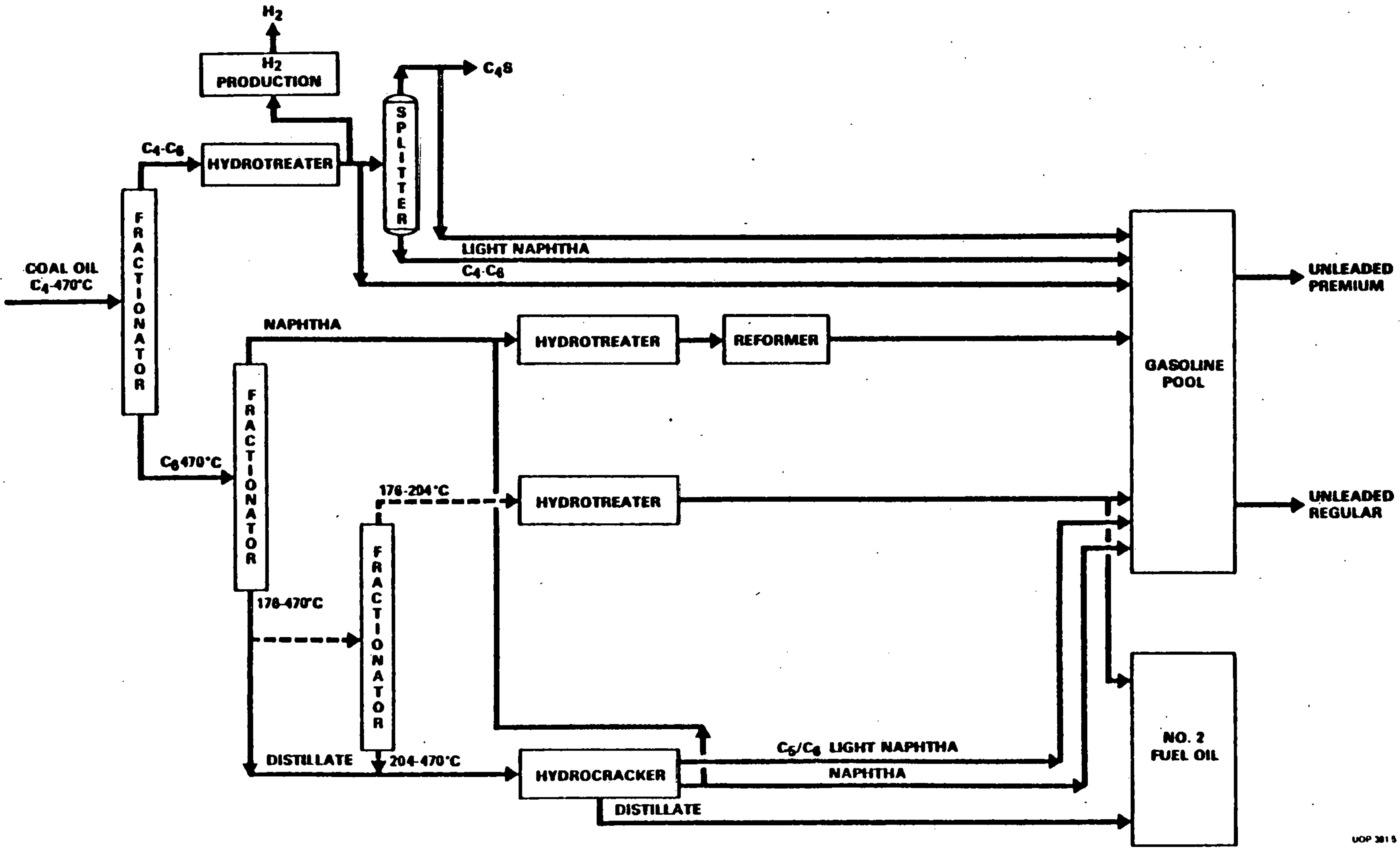
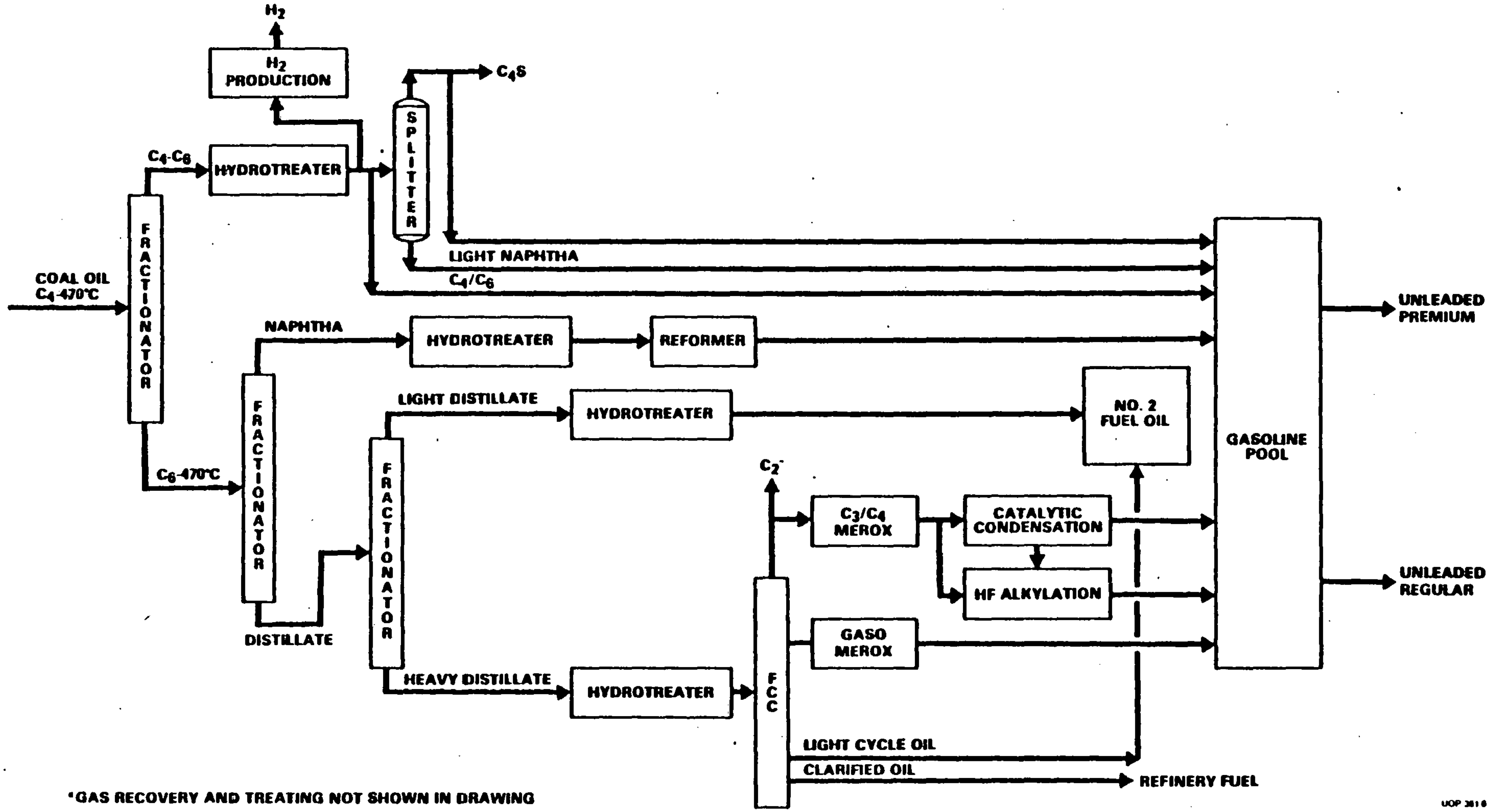


FIGURE 7
(SCHEME C)

COAL OIL REFINERY

FCC CASES WITH LOW TEMPERATURE FRACTIONATION



*GAS RECOVERY AND TREATING NOT SHOWN IN DRAWING

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FIGURE 8

H-COAL LIQUID DISTILLATION

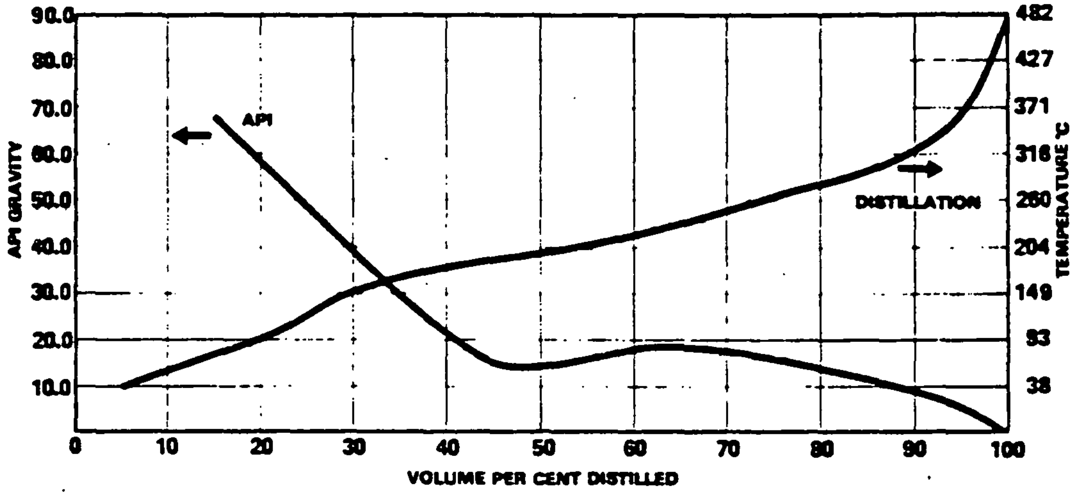
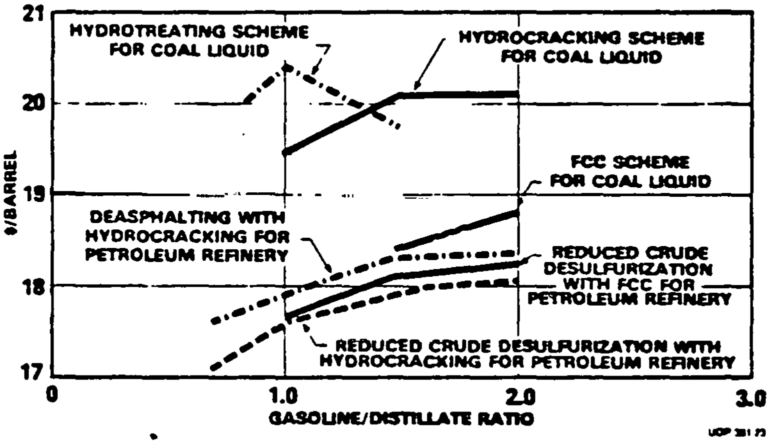


FIGURE 9

ECONOMIC COMPARISON FOR PETROLEUM vs. COAL LIQUID REFINERY



APPENDICES A AND B

APPENDIX A
65/35 Lt./Hvy. Arabian Petroleum Blend

Component	Vol.%	Wt.%	Sp.Gravity	Sulfur,Wt.%	RON-0	MON-0	Paraffin, LV%	Naphthene,LV%	Aromatic,LV%
C_3-	.70	.40	.4878	.0003	89.0	91.6	100.0		
C_4	1.74	1.17	.5807	.007	66.0	65.1	91.7	6.1	2.1
C_4-66°C	5.11	3.83	.6477	.02	38.0	35.9	67.9	20.7	11.4
66°C-160°C	13.86	11.75	.7328	.04	33.0	31.0	65.0	20.7	14.3
(66°C-193°C)	(19.72)	(17.03)	.7467	.23	15.2	13.8	57.7	21.0	17.7
160°C-260°C	17.04	15.74	.7982	.31	12.1	11.0	57.4	21.2	15.9
(193°C-260°C)	(11.30)	(10.56)	.8076	1.4					
260°C-352°C	14.61	14.42	.8535	2.5					
352°C-566°C	29.40	31.12	.9153	5.0					
Residue	17.93	21.47	1.0350						
Total		100.00	.8645	2.09					

Component	Con Carbon, Wt.%	Viscosity Index @ 50°C	N_2, Wt.%	Smoke Point	Pour Point, °C
160°C-260°C		5.5			-53.0
193°C-260°C		7.7			-44.9
260°C-352°C	.002	17.0	.01	5.15	-10.8
352°C-566°C	.66	29.5	.08		29.7
Residue	22.85	48.2	.31		48.9
Total	5.1		.091		

APPENDIX A
Coal Liquid Properties

	C_4-471°C	C_6-471°C	C_4/C_6	C_6/177°C	C_6/193°C	C_6/204°C	177/204°C	177/193°C
Yield, vol.%	100.0	86.13	13.87	30.84	38.9	41.25	10.41	8.10
Yield, wt.%	100.0	90.2	9.8	28.39	37.5	40.14	11.75	9.11
API	30.5	23.2	97.9	44.5	36.8	35.0	12.2	11.1
Sp. Gr.	.8733	.9145	.617	0.804	0.8408	0.8498	0.9844	0.9823
Oxygen, wt.%	1.72	1.97	-	0.52	2.7	3.0	8.6	8.2
Nitrogen, wt.%	0.37	0.40	-	0.052	0.29	0.30	0.9	1.0
Sulfur, wt.%	0.15	0.15	0.1	0.2	0.2	0.2	0.2	0.22
CCR, wt.%	0.10	0.10	-	-	-	-	-	-
Carbon, wt.%	86.7	87.1	83.5	85.5	85.1	84.9	83.5	83.7
Hydrogen, wt.%	11.0	10.5	16.5	13.1	11.9	11.6	7.9	8.1
Bromine Number	41.7	43.1	26	11.0	11.0	12.0	11.0	11.0
Ash, ppm	67	63.0	-	1	1	1	1	
C_7 Insol.	.10	.11	-	-	-	-	-	-

APPENDIX A
H-Coal Distillate Properties

	204/343°C	343°C+	204°C+	204/260°C	204/288°C	260°C+	288°C+
Yield, vol.%	40.89	3.99	44.88	21.88	32.58	23.0	12.3
Yield, wt.%	45.19	4.87	50.06	23.55	35.66	26.51	14.4
API	15.1	1.3	13.6	19	16.5	9.1	6.9
Sp. Gr.	.9652	1.0655	.9742	.9402	.9561	1.0067	1.0225
Oxygen, wt.%	1.0	1.3	1.0	1.0	1.0	1.05	1.1
Nitrogen, wt.%	0.4	1.3	.5	.38	.39	.55	.8
Sulfur, wt.%	0.1	0.2	.1	.1	.1	.12	.14
CCR, wt.%	<.1	2.0	.2	<.1	<.1	.4	.7
Carbon, wt.%	88.8	89.2	88.8	87.9	88.4	89.0	89.1
Hydrogen, wt.%	9.7	8.0	9.6	10.1	9.8	8.7	8.3
Bromine Number	50		50	about the same			
Ni + V, wt. ppm		(small - other metals also present)					
Ash, ppm	10	1200	120	1	2	220	340
C_7 Insol.	0	2.0	.2	0	0	.4	.7

APPENDIX B

ASTM Specifications

Grade of Fuel Oil	Sulfur Wt. %	Flash Pt. °C	Pour Pt. °C	Distillation Temperature °C		Kinematic Visc. cSt @ 38°C		Kinematic Visc. cSt @ 50°C	API
				90%	Pt.				
	(Max.)	(Min.)	(Max.)	(Min.)	(Max.)	(Min.)	(Max.)	(Max.)	(Min.)
No. 2	0.2	38	-7	282.0	338.0	2.0			30
No. 6	0.7	60.0						638	

	Distillation Temperature, °C					RVP	Sulfur
	10% Max.	50% Min.	Pt. Max.	90% Max.	End Point Max.	Lb. Max.	Wt.% Max.
Gasoline	55	77	113	185	225	13.5	0.10