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SRC-II DEMONSTRATION PROJECT

PHASE ZERO

TASK NUMBER 3

DELIVERABLE NUMBER 9

VOL. 8 OF 9

MARKET ASSESSMENT

TRANSPORTATION FUELS FROM SRC-II UPGRADING

JULY 31, 1979

THE PITTSBURG & MIDWAY COAL MINING CO.
DENVER, COLORADO

PREPARED FOR

UNITED STATES DEPARTMENT OF ENERGY
UNDER CONTRACT
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SECTION 1
SUMMARY

Transportation fuels essentially indistinguishable from petroleum counterparts could be produced from SRC-II liquids by commercially available upgrading technology.

Conversion of SRC-II Naphtha to Gasoline

Converting SRC-II naphtha to gasoline could be accomplished at no net change in the value or cost of the SRC-II naphtha, because the added value that would result from upgrading to gasoline would be offset by the estimated cost of upgrading. Between 13,000 B/D and 22,000 B/D, including SRC-II butanes, of high octane (clear) gasoline blend stock could be produced by pretreating and reforming SRC-II naphtha. The production range would result from the boiling range of the naphtha feedstock to the conversion plant and from the coal feedstock to the SRC-II plant.

The additional cost of upgrading is estimated to be about $5/B, depending on the above feedstock factors, but as mentioned above, this is offset by the higher value of the gasoline blend stock. The additional capital investment required is estimated to be about $74 million.

The high clear (unleaded) octane of the SRC-II gasoline blend stock would result in greater yields of gasoline from crude petroleum, when blended at a petroleum refinery, because of a reduced refinery reformer severity requirement to meet the same gasoline pool octane requirement.

Gasoline output from a 150,000 B/D petroleum refinery could be increased by about 22% if the upgraded SRC-II naphtha from a 30,000 T/D SRC-II commercial conceptual plant were blended at the refinery. The petroleum refinery gasoline output would be about 105,000 B/D but with blending of SRC-II gasoline blend stock the output would increase to about 128,000 B/D.

The economics of this case and the others in this volume are based on conservative 1990 petroleum product value forecasts. These forecasts were made prior to recent oil price escalation and increasing emphasis by crude suppliers on marketing heavier crudes. The value of naphtha, in particular, has increased. This implies further evaluation of the economic potential for SRC-II naphtha would be more favorable, especially in the years beyond 1990 as petroleum becomes increasingly scarce and expensive.

Upgrading SRC-II Liquids to Maximize Transportation Fuels Production

Using all of the SRC-II naphtha and essentially all of the fuel oil range SRC-II distillate from a 30,000 B/D conceptual commercial SRC-II plant as feedstock for severe upgrading, it is estimated that the following products could be produced:

	B/D
Gasoline, including SRC-II butanes	32,000
Jet Fuel	24,000
Diesel Fuel	4,000
Total	60,000

This could be accomplished from feeding 24,000 B/D of SRC-II naphtha (C_5-400^oF) and 31,000 B/D of distillate, or a total of 55,000 B/D. The volumetric increase reflects the substantial hydrogenation required.

Capital investment required for this upgrading alternative is estimated to be about $207 million. The additional cost of upgrading is estimated to be about $7/B. Adding the $23/B estimated cost of SRC-II liquids feedstock to the cost of upgrading results in a total cost of about $0.72 per gallon.

This upgrading approach is based on data developed by Chevron in a DOE funded study of upgrading coal liquids. The data used were specifically for SRC-II liquids and were based only on available technology and catalysts, operating at higher severities than are now used commercially.

It is not possible to use coal liquids as petroleum refinery feedstocks with current petroleum refining facilities, because of the high severity pretreating needed for coal liquids. It is also not possible to average down the severity requirements at a petroleum refinery by blending coal liquids into the refinery crude charge. Coal liquids, such as SRC-II naphtha distillate, have too high a nitrogen content (a refining catalyst poison) to be processed in any significant amounts in a conventional present day petroleum refinery. The only presently available way to convert SRC-II-derived naphtha to gasoline in an existing refinery is to first pretreat the naphtha at high severity and then proceed with conventional refining.

Indirect Supply of Transportation Fuels from SRC-II Products

The products from the SRC-II process can most readily make available high quality petroleum products by displacement and substitutions. The SRC-II products from a commercial conceptual plant, including the ethylene and gasoline (from naphtha only), and their projected corresponding effect on petroleum products supply are summarized as follows.

Methane

Methane from the SRC-II process could be used directly in the existing natural gas distribution network to reduce demand for home heating oil or No. 2 fuel oil in industrial applications. This would make these middle distillate petroleum products more available.

Ethane and Propane

Ethane and propane from the SRC-II process could be used for conversion to ethylene, the principal building block of the petrochemical industry. The petroleum naphtha and gas oil feedstocks now used could be displaced by this means and, with added petroleum refining, made available as gasoline and middle distillate products (No. 2 diesel fuel, jet fuel).

Naphtha

SRC-II naphtha is an excellent gasoline reformer feedstock after it is pretreated to remove nitrogen. The pretreatment steps could be carried out with available but unused technology at competitive cost based on current estimated costs for SRC-II naphtha. Thus, pretreated and reformed SRC-II naphtha could directly augment the supply of high octane unleaded gasoline. Availability of SRC-II-derived gasoline would enable higher yields of gasoline from the oil normally charged at petroleum refineries by reducing the severity of the reforming necessary to achieve a given octane rating, thus increasing the yield.

Middle Distillate Fuel Oil

SRC-II fuel oil in the middle distillate range could potentially substitute for No. 2 fuel oil in electric utility combustion turbines. This would make available more middle distillate petroleum products, such as jet fuel, diesel fuel, and home heating oil. Also, the SRC-II middle distillate, possibly with mild hydrotreating, could be used as medium-speed diesel fuels for railroad locomotive and domestic marine applications. This is another area where SRC-II products could release an equivalent amount of higher quality petroleum-derived diesel fuel which is required for truck transportation applications.

Middle Distillate/Heavy Distillate Fuel Oil Blend

SRC-II middle and heavy distillate fuel oil blend can be used as an environmentally acceptable industrial and utility boiler fuel. The displaced No. 2 through No. 6 fuel oil could be refined into a full slate of petroleum distillate products.

INTRODUCTION

The potential applications of SRC-II products, which all have been discussed in the previous volumes, are initially the most near-term among SRC-II alternatives and hence have received the most study. Energy economics are not static and neither are those of the SRC-II process or the utility of SRC-II products in the energy market-place. As petroleum costs increase in real terms, the economics of SRC-II fuels should become increasingly competitive with petroleum.

The following reports discuss the technology and economics of the use of SRC-II liquids to produce transportation fuels indistinguishable from their petroleum counterparts. The technology for converting SRC-II products to gasoline and other transportation fuels is commercially available. Of course, no commercial experience exists in upgrading coal liquids from SRC-II to transportation fuels.

The economics in the following reports have been prepared from two points of view: Determination of the additional cost of upgrading, and a comparison of the value of the gasoline stream to a petroleum refinery for blending with petroleum-derived transportation fuels. The latter requires comparison with the cost of crude petroleum to the refinery. In order to facilitate this comparison, it is necessary to calculate the cost to the refinery of the alternative feedstocks, crude petroleum or raw SRC-II liquids. This was done by using 1990 forecast prices for products deflated to 1979, adding manufacturing costs and the same rate of return on invested capital. This procedure determines the revenue requirements and establishes the cost to the refiner of each feedstock.

Thus the economics in the following reports comparing SRC-II and crude oil values will be meaningful only for the purpose of screening alternatives. The economics prepared for comparison with the crude petroleum alternatives will be of relative interest and will not relate to marketplace prices. The economics do permit a judgement to be made, however, on the merits of converting SRC-II liquids to transportation fuels as the cost of refined petroleum-derived transportation fuels increases.

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UPGRADING SRC–II NAPHTHA TO GASOLINE

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CHEMICALS AND MINERALS DIVISION
REPORT NO. 633RK102
JULY, 1979

OVERALL SUMMARY

To aid in establishing the economics of an SRC-II conceptual commercial plant operation, the SRC-II naphtha fraction has been evaluated as a feedstock to a commercially available upgrading process to produce a high octane (unleaded) gasoline stock for blending at a petroleum refinery. This report presents the results of these calculations.

The study was done in two parts. Both parts assume that the SRC-II naphtha stream would be converted to a high octane material which is blended with other gasoline streams from a large, hypothetical petroleum refinery. In Part 1, only the SRC-II naphtha stream is upgraded, while in Part 2 a wider boiling stream including some heavier material is processed to increase throughput in an effort to take advantage of the economy of scale.

Both Parts 1 and 2 involve evaluation of a number of processing options, including alternative crude oil charge stocks to the refinery and variations in the naphtha fraction processed in the upgrading step. Three main cases were evaluated in Part 1 and one case in Part 2. All of the cases were evaluated using a proprietary Gulf computer program known as the Nonlinear Refinery-Chemicals Model. This model has the capability of optimizing process schemes by maximizing rate of return on investment. The analyses, therefore, are done on the basis of simple return on investment for comparison purposes, and no effort is made to use a discounted cash flow type of calculation.

The results suggest that SRC-II naphtha, at a calculated price indicative of a successful demonstration of the technology followed by commercialization, could be economically upgraded to a high (clear) octane number gasoline blending component using currently available technology. This conclusion was reached by determining the value of the SRC-II naphtha stream in each refinery scheme. The results are independent of the price of crude and do not reflect any real market price expectations; instead, these results provide a useful means only of comparing alternatives. The calculated values of raw SRC-II naphtha relative to the calculated value of crude oil are as follows:

SRC-II Naphtha Charge to Reforming		C_5-350°F	C_5-450°F
		Ratio: SRC-II Naphtha, \$/B	
Crude Oil	Coal Charge to SRC-II Plant	Crude Oil, \$/B	
West Texas Sour	Pittsburgh Seam	0.99	--
Kuwait	Pittsburgh Seam	1.01	1.00
Kuwait	West Virginia Powhatan	1.07	--

The value of raw SRC-II naphtha would be slightly lower than that of West Texas sour crude, which is a typical major domestic crude. However, the reserves of this crude are expected to approach depletion in the not too distant future. New crude oil is expected to become progressively heavier, higher in sulfur content, and more costly to refine. Current production of light, low-sulfur crudes in the free world is as high as 62% of the total, while known reserves are estimated to be only about 49% of the total. The value of Kuwait crude, which is considered representative of the heavier, higher-sulfur content crudes, is 97% of that of the lighter, lower sulfur West Texas crude. The value of raw SRC-II naphtha would be slightly higher than that of Kuwait crude.

The comparisons in values between projected commercial SRC-II liquid product and crude petroleum developed in this study are based on current costs and a given rate of return on capital. However, in the future, the increased costs for more risk taking exploration, on-shore as well as off-shore, higher costs for deeper drilling, and much higher costs for enhanced recovery of oil from existing fields, can only increase the price for petroleum products. At the same time, the constant dollar costs for SRC-II liquid product are expected to increase only modestly as a result of technological improvements which history has proven will take place after the launching of a new process. Therefore, the projected value of SRC-II naphtha relative to petroleum, initally at an equivalency as calculated in this study, is expected to increase significantly as supplies of petroleum decrease and costs increase.

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PART 1

1.0 SUMMARY

Economics have been calculated for hydrotreating and reforming SRC-II naphtha to a high octane number (97.0-98.0 RON clear) gasoline blending component. Raw naphtha (C_5-350°F) would be produced from a projected commercial 30,000 T/CD SRC-II plant at a rate of 11,961 B/CD when charging a Pittsburgh Seam coal considered in a previous design basis for the SRC-II plant. The raw naphtha production rate is 16,200 B/CD when charging a West Virginia Panhandle Powhatan coal in the current SRC-II plant design basis. High octane number gasoline blending stocks, including mixed SRC-II butanes, could be produced at rates of 12,903 B/CD and 21,961 B/CD, respectively. In addition, hydrotreated light distillate (350-400°F), which qualifies as a jet fuel component, would be produced at a rate of 3583 B/CD in the scheme charging naphtha at the higher rate. In the Powhatan coal case, the naphtha charge (C_5-400°F) is 23,760 B/CD.

Investments and manufacturing expenses for the naphtha upgrading plants are as follows:

Coal	Pittsburgh Seam	Powhatan
Raw SRC-II Naphtha Charge, B/CD	11,961	23,760
Total Plant Investment, $\$10^6$ (1978)	47.9	74.1
Total Mfg. Expense plus Cost of Capital $/B SRC-II Naphtha Charge	5.71	4.73

The upgraded naphtha would be blended into gasoline products from a large existing petroleum refinery for each of the following three objectives: to maintain constant production of gasoline and No. 2 fuel oil at reduced crude rate; to increase gasoline production with no change in crude rate; and to increase No. 2 fuel oil production with no change in crude rate. In one scheme, 150,000 B/CD of West Texas sour crude is charged to a refinery which employs present-day technology including coking of vacuum bottoms. In a second scheme, Kuwait crude, having a higher sulfur content, is charged at the same rate to a refinery which employs advanced technology including hydrodesulfurization (HDS) of atmospheric bottoms plus fluid catalytic cracking (FCC) of HDS bottoms.

Calculated values of the raw naphtha to the refiner for each of these cases along with the calculated crude oil prices, are as follows:

	Calculated Values for Purpose of Comparison		
Crude	West Texas	Kuwait	
Raw SRC-II Naphtha (C_5-350°F) B/CD	11,961	11,961	16,200
Crude Oil Price, Calculated, \$/B (\$/10^6 Btu)	19.67 (3.43)	19.14 (3.33)	19.14 (3.33)
Raw SRC-II Naphtha, \$/B (\$/10^6 Btu)			
For Crude Replacement	15.71 (2.88)	-	-
For Additional Gasoline	19.41 (3.56)	19.29 (3.54)	20.43 (3.78)
For Additional No. 2 Fuel Oil	17.69 (3.22)	-	-

On the basis of increasing refinery gasoline production, the value of raw SRC-II naphtha would compare favorably with the price of crude petroleum, despite the high capital cost for removing the high concentration of nitrogen. Relative values of raw SRC-II naphtha to petroleum increase as the naphtha production rate increases and as the gravity and sulfur content of petroleum increase.

The revenue requirements on which the calculated values are based were prepared from product price projections which are not up-to-date and which, therefore, do not reflect the recent upsurge in crude petroleum as well as the growing premium in price for lighter feedstocks. The refined product prices were for 1990 and were subsequently deflated by 6% per year to 1979. Crude oil prices are calculated to include 20% cost of capital before tax in the Base Case refineries without SRC-II naphtha. Naphtha upgrading and incremental refinery costs, a cost of capital of 20%, and transportation of reformate from the hypothetical SRC-II site to the hypothetical refinery are included in the naphtha values given above. The refinery economics are derived from a proprietary Gulf optimization model to give the best refinery configuration. A simple return on investment approach is used. For this reason, these economics are not done on a discounted cash flow (DCF) basis, and the selection of 20% before tax return on investment (or cost of capital) is entirely arbitrary and is clearly not reflective of current industrial requirements for return on investment. In summary then, these economics will appear low, because of low forecasted product prices and low return on investment, but have been prepared only for the purpose of comparing and screening alternatives.

2.0 INTRODUCTION

Naphtha produced from liquefaction of coal by the SRC-II process contains relatively large quantities of cyclic hydrocarbons, typically 34% aromatics and 45% naphthenes, and has been demonstrated in Gulf pilot plant runs[1] to be an excellent reformer feedstock for production of high octane number gasoline. A high yield, about 90 vol %, of high octane number (97.0 Research octane number clear) reformate can be obtained from the pretreated SRC-II naphtha. Also, the yield of hydrogen from reforming is in considerable excess of that required for pretreating. However, this naphtha also has high concentrations of hetero-atom catalyst poisons: typically, 0.5% nitrogen, 0.2% sulfur, and 2.2% oxygen from liquefaction of a coal for a proposed commercial plant. The required hydrotreating conditions to produce suitable feedstock to a conventional catalytic reforming unit are severe.

One of the major problems in evaluating the economics of coal conversion processes involves determining values of coal-derived liquids. In some studies, a price has been estimated for this liquid based solely on the cost of producing it, including an arbitrary profit. Other studies have tried to estimate a value for the coal-derived liquid on the basis of some inherent properties.

The most realistic method of estimating the value of a coal-derived liquid relative to refined petroleum products is the determination of its value as a petroleum refinery feedstock. This value depends primarily on the market value of normal refined products which can be made from it and the processing costs to make such products. In this study, economics have been calculated for hydrotreating SRC-II naphtha to produce feedstock for a conventional reforming unit, and then blending the high octane number reformate into gasoline products from a large, hypothetical petroleum refinery.

Economics are evaluated for upgrading, at the SRC-II site, C_5-350°F raw naphtha produced from a projected 30,000 T/CD commercial SRC-II plant at a rate of 11,961 B/CD when charging a high-sulfur bituminous Pittsburgh Seam coal from West Virginia considered in a previous SRC-II plant design basis, and at a rate of 16,200 B/CD when charging West Virginia Panhandle (Powhatan) coal in the current design basis. The Powhatan coal has been found in pilot plant work to result in higher yields of naphtha and total liquid product and a slightly lower yield of 350°F$^+$ fuel oils than for the Pittsburgh Seam coal.

High octane number (97.0-98.0 RON clear) reformate is transported by pipeline to a hypothetical existing petroleum refinery at an assumed distance of 300 miles. In one case, it is assumed that 150,000 B/CD of West Texas sour crude is charged to the refinery which employs present-day technology including delayed coking of vacuum bottoms. Although this crude has been produced for many years, and is still a major domestic crude, its rate of production is declining and its availability may be severely curtailed within another decade. Therefore, Kuwait crude, which has a higher sulfur content than West Texas crude and is representative of the higher sulfur crudes expected to be produced at progressively higher rates in the future, has been evaluated in an alternative refinery scheme at the same throughput. Advanced refining technology, including hydrodesulfurization (HDS) of crude atmospheric residue followed by fluid catalytic cracking (FCC) of HDS residue, is employed in the Kuwait crude refining scheme.

For the West Texas crude refinery, SRC-II naphtha is evaluated on the basis of crude replacement, Case II; production of additional gasoline at the same crude rate, Case III; and production of additional No. 2 fuel oil at the same crude rate, Case IV, respectively. For the Kuwait crude refinery, SRC-II naphtha is evaluated only on the basis of increasing gasoline production, Case III-A. The originally proposed Pittsburgh Seam coal is evaluated as the coal feed in each case. In addition, Powhatan coal is evaluated as a coal feed for Case III-A, based on a Kuwait crude refinery in which SRC-II reformate is added to increase gasoline production.

The refining schemes and economics were developed using the Gulf nonlinear refinery-chemicals model. This model has the capability of optimizing feed rates to selected processing units in a specified refinery or petrochemical scheme, the severity of naphtha reforming, and the blending of a number of grades of gasoline and fuel oils to meet various product specifications. It can maximize either absolute profit or rate of return on invested capital. Nonlinear functions, such as the exponential function of investment with process unit capacity, which cannot be readily accommodated in linear programs, can be executed in the nonlinear model. Since the model is proprietary, having been developed under Gulf funding, the program details are not included in this report. Furthermore, that portion of this study involving the non-linear model was not billed to the Phase Zero contract.

Economics were calculated for hypothetical grass roots Base Case refineries charging West Texas or Kuwait crude, respectively, without SRC-II naphtha (Cases I and IA) for startup in 1979. Downstream refining units would then be increased in capacity by revamping as required to accommodate charge of SRC-II reformate to the scheme. Prices for primary petroleum products are those forecast for 1990, deflated by 6%/yr to 1979. The price of crude oil is calculated to provide 20% return on total capital before tax in the Base Case refinery.

3.0 HYDROTREATING AND REFORMING OF SRC-II NAPHTHA

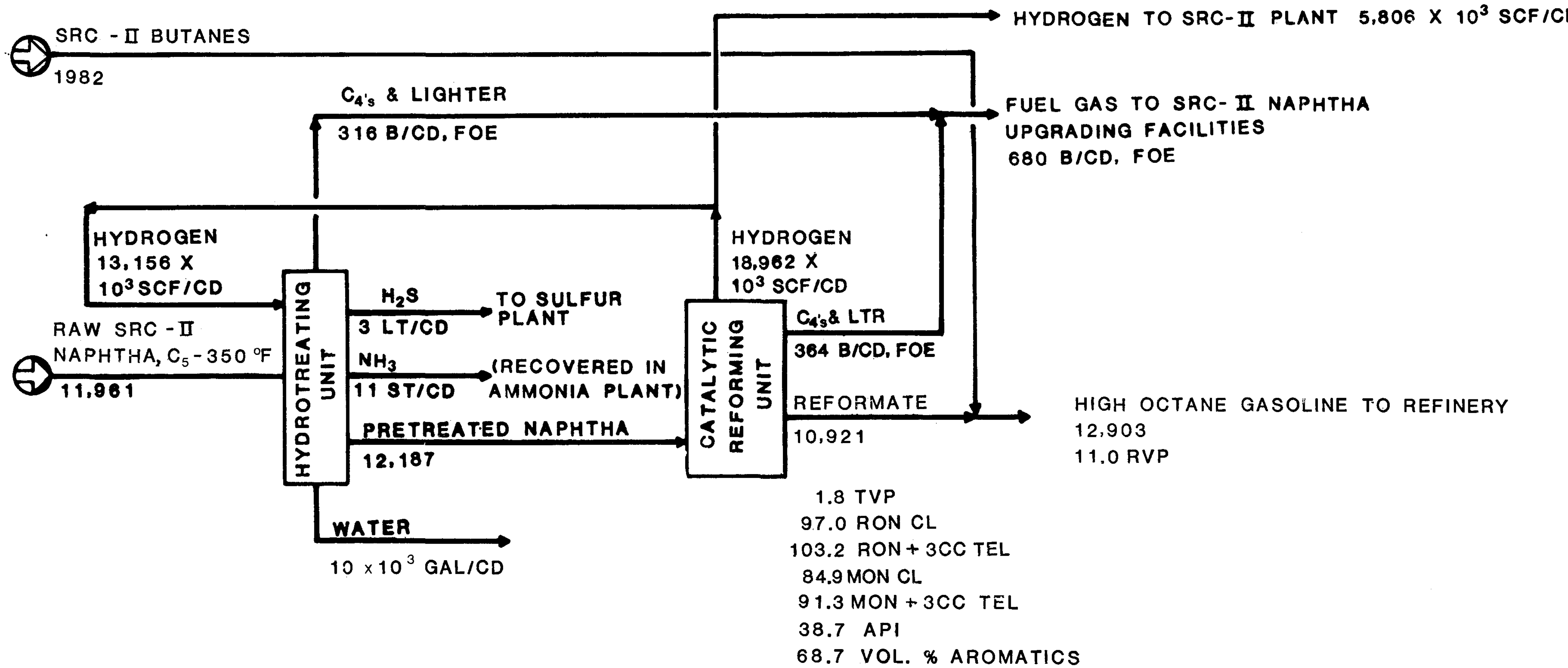
A schematic flow diagram of the naphtha upgrading plant based on charging a Pittsburgh Seam coal to the SRC-II plant, considered in the early design basis, is shown in Figure 1. Operating conditions and product yields and inspections for naphtha hydrotreating and reforming units are based on a

FIGURE 1

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

NAPHTHA FROM PITTSBURGH SEAM COAL

SCHEMATIC FLOW DIAGRAM



NOTE: FLOW RATES ARE IN B/CD UNLESS OTHERWISE SPECIFIED.

C.E.S.
GS&TC
C&MD
3/15/79

REF. DWGS.	UPGRADING OF SRC-II NAPHTHA	
	GULF RESEARCH & DEVELOPMENT CO. PITTSBURGH, PA.	
	DATE 3-15-79	SCALE
	DRAWN	APV'D. ANK 7/13/79
	TRACED	APV'D.
	CH'K'D.	APV'D.
	DWG. NO.	

Gulf pilot plant study. Raw SRC-II naphtha (C_5-350°F) is charged to a hydrotreating unit at a rate of 11,961 B/CD to produce 12,187 B/CD of reformer feedstock. Nitrogen removal to less than 0.2 ppm is the limiting constraint. Makeup hydrogen is supplied at a rate of 13.2 x 10^6 SCF/CD to provide for 900 SCF/B of chemical consumption and 200 SCF/B of solution and bleed losses. Hydrogen is supplied in a stream of 96% purity as obtained as a by-product from reforming of pretreated naphtha. A catalyst life of one year has been assumed.

Water is injected into the reactor effluent cooling train to scrub out ammonium bisulfide formed upon cooling the reaction products. Sour water from the low-pressure separator is stripped to recover ammonia and hydrogen sulfide by-products. Bleed gas from the high-pressure separator and gas from stripping of hydrotreated naphtha are charged to an existing ammonia plant for recovery of the remainder of the ammonia. Total ammonia production is 11 ST/CD. Hydrogen sulfide recovered in the sour water stripper is charged to an existing sulfur plant for recovery of 3 LT/CD of sulfur product.

Pretreated naphtha is reformed in a conventional semi-regenerative unit employing UOP bimetallic catalyst to produce 10,921 B/CD of debutanized reformate (89.6 vol % on charge) having 97.0 Research octane number clear. Hydrogen is produced at a rate of 19.0 x 10^6 SCF/CD in a stream of 96% purity. About 70% of the hydrogen is supplied to the naphtha hydrotreating unit; the remainder, 5.8 x 10^6 SCF/CD, is available for hydrogen consuming units in the SRC-II plant. Butane and lighter hydrocarbons produced from reforming, 3.0 wt % on charge, and from pretreating, 1.6 wt% on charge, supply a portion, 680 B/CD FOE, of the fuel requirements for the upgrading facilities. The remaining fuel required, 363 B/CD FOE, is supplied from the projected SRC-II plant. The production rate of 11.0 psi RVP high octane number gasoline blending component, including 1982 B/CD of SRC-II mixed butanes, is 12,903 B/CD.

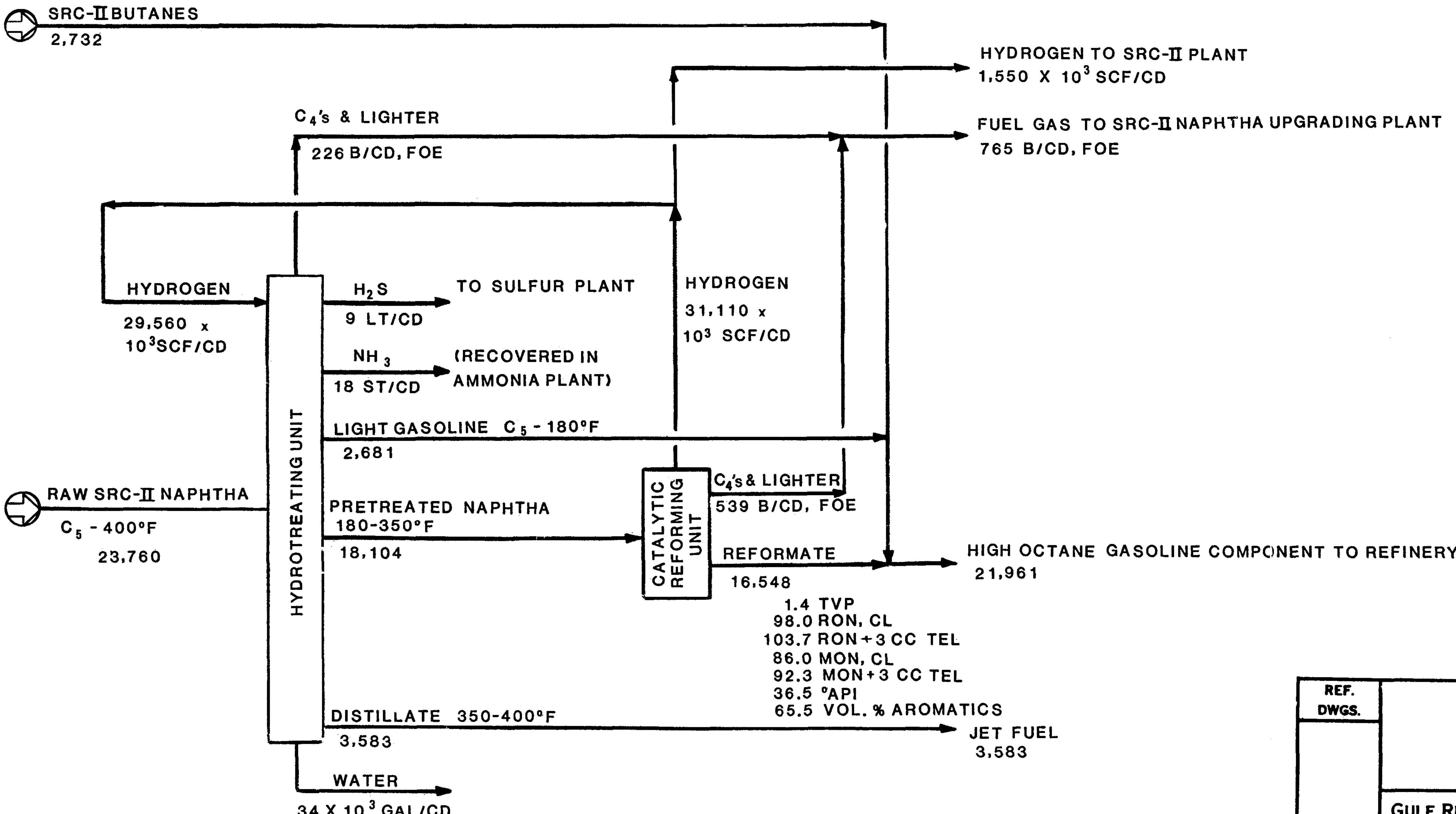
A schematic flow diagram of a projected naphtha upgrading plant based on a higher feed rate, 16,200 B/CD of C_5-350°F naphtha, obtained from Powhatan coal in the current design basis, is shown in Figure 1A. Processing

FIGURE I-A

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

NAPHTHA FROM WEST VIRGINIA POWHATAN COAL

SCHEMATIC FLOW DIAGRAM



NOTE: FLOW RATES ARE IN B/CD UNLESS OTHERWISE SPECIFIED.

C.E.S.
C&MD
GS &TC
6/29/79

REF. DWGS.	UPGRADING OF SRC-II NAPHTHA
	GULF RESEARCH & DEVELOPMENT CO. PITTSBURGH, PA.
	DATE 6-29-79 / SCALE
	DRAWN / AP'V'D.
	TRACED / AP'V'D.
	CH'K'D. / AP'V'D.
	DWG. NO.

data were obtained from a study[1] by Chevron Research Company prepared for the Department of Energy. Charge to the hydrotreating unit comprises raw C_5-400°F naphtha at a rate of 23,760 B/CD. The 180-350°F fraction of hydrotreated naphtha, 18,105 B/CD, is reformed at a high severity level to produce debutanized reformate having 98 RON clear at a rate of 16,548 B/CD. Total production of gasoline from the upgrading plant including 2681 B/CD of hydrotreated light gasoline, C_5-180°F, and 2732 B/CD of SRC-II butanes is 21,961 B/CD. Hydrotreated light gasoline and reformate are transported separately to the refinery in blocked operation in order to provide flexiblity in blending into the two grades of gasoline products.

The hydrotreated 350-400°F fraction, 3583 B/CD, could be priced as a jet fuel product. The high density of this product, 25 °API, vs manufacturing specifications of 37-51 °API for Jet A product is a reflection of the high naphthenes content, with the aromatics content indicated to be within an acceptable maximum for jet fuel of 25%.

Operating conditions and product yields for the hydrotreating unit, related to the use of Powhatan coal, are presented in Table I. With the catalyst employed by Chevron, hydrotreating operating conditions are less severe--lower hydrogen partial pressure and higher reactor space velocity, to produce naphtha meeting a 0.1 ppm nitrogen constraint--than employed in the GR&DC pilot plant. Hydrogen consumption reported by Chevron is somewhat higher than that by GR&DC which is largely a result of a feedstock containing more oxygen. Operating conditions and yields for reforming were based on the GR&DC pilot plant data and are essentially the same as those employed for reforming of naphtha derived from Pittsburgh Seam coal.

4.0 WEST TEXAS CRUDE REFINING SCHEMES

4.1 Base Case Refinery (Case I)

West Texas sour crude is charged at a rate of 150,000 B/CD to a Base Case refinery, shown schematically in Figure 2, for high gasoline production. Vacuum bottoms (1000°F+) is charged to a delayed coking unit for production of

TABLE I

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

Naphtha Hydrotreating Units
Operating Conditions and Product Yields and Inspections

Coal Source	West Virginia Powhatan
Pilot Plant Data	Chevron Research Company
Charge Stock	SRC-II Naphtha, C_5-400°F
Operating Conditions	
Total Pressure, psig	793
Hydrogen Partial Pressure, psia	645
Space Velocity, v/h/v	0.95
Recycle Gas Rate (85% H_2), SCF/B	3450
Average Temperature, °F	699
Hydrogen Consumption, SCF/B	
Chemical	1157
Solution and Bleed Losses	53
Catalyst	Chevron ICR 113 Ni-Mo on Alumina
Catalyst Life (Estimated), Years	9

Product Yields and Inspections

	Vol %	°API	Wt %	ppm S	ppm N	ppm O
Charge	100.0	38.0	100.00	2600	4200	35 100
Hydrogen, Chemical			2.02			
Total			102.02			
Products						
NH_3			0.51			
H_2S			0.27			
H_2O			3.94			
C_1			0.03			
C_2			0.12			
C_3			0.20			
C_4's			0.12			
Gasoline, C_5-180°F	11.3	63.0	9.42			
Naphtha, 180-350°F	76.2	41.1	71.75	11*	<0.10	<90
Distillate, 350-400°F	15.1	25.0	15.66			
Total	102.9		102.02			

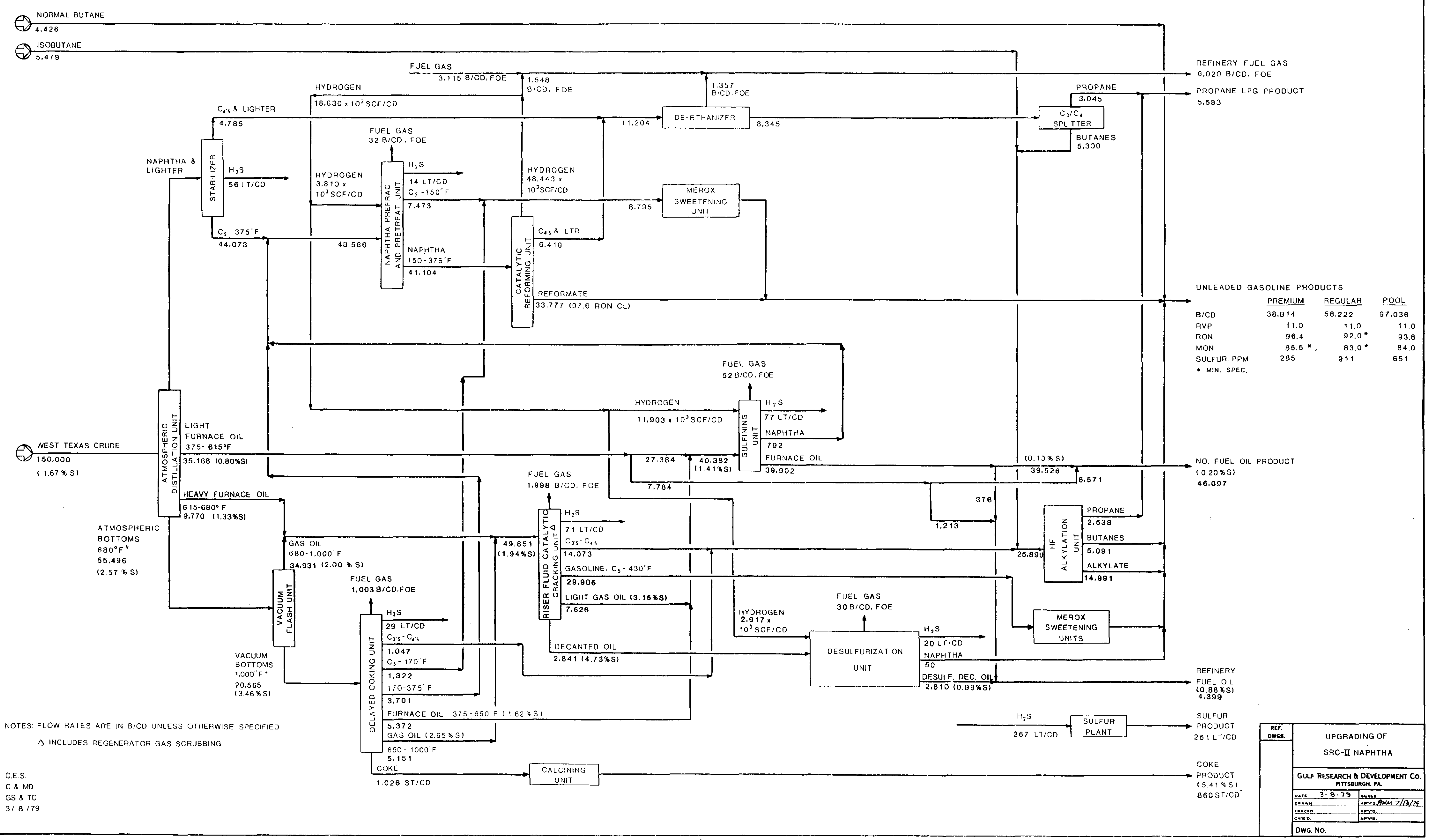
FIGURE 2

UPGRADING OF SRC - II NAPHTHA TO GASOLINE

CASE 1 - BASE CASE - WEST TEXAS CRUDE REFINING SCHEME

DELAYED COKING OF VACUUM BOTTOMS

SCHEMATIC FLOW DIAGRAM



860 ST/CD of high-sulfur, 5.4% S, coke. The coke is calcined to remove volatile matter. Straight-run gas oil (615-1000°F) plus coker gas oil (650-1000°F) are charged to a riser FCC unit employing high-activity zeolite catalyst for high conversion, 79.0 vol %, to gasoline, 430°F TBP end point, and lighter products. Propene and butenes from the FCC and coking units are alkylated over HF catalyst.

Straight-run naphtha (150-375°F) and coker (170-375°F) naphthas are reformed over UOP R-18 bimetallic catalyst to produce a debutanized reformate having 97.6 Research octane number clear. This relatively high severity is required to produce an unleaded gasoline pool containing 60% regular and 40% premium grades meeting the following minimum octane ratings forecast for 1990:

	Regular	Premium
Research Octane Number (RON)	92.0	96.0
Motor Octane Number (MON)	83.0	85.5

In this scheme, Motor octane number is controlling in the premium grade, with Research octane number exceeding the minimum by 0.4 number.

Propane produced in the reforming, FCC, alkylation, and coking units is recovered at a rate of 90% in a gas plant to LPG product. Butanes recovered from these units at a rate of 98% are supplemented with purchased iso and normal butanes to meet alkylation and gasoline vapor pressure, 11.0 psi RVP, requirements.

FCC, coker, and most of the straight-run furnace oil, 375-615°F, are Gulfined at 93% desulfurization to produce a furnace oil product containing 0.1% sulfur. A small portion of Gulfined furnace oil is blended into refinery fuel oil, with the remainder blended with undesulfurized straight-run furnace oil to produce No. 2 fuel oil product meeting a 0.2% sulfur specification.

FCC decanted oil containing 4.7% sulfur is desulfurized by 80% to a sulfur content of 1.0% and blended with portions of straight-run and Gulfined furnace oils to produce a fuel oil having a maximum sulfur specification of 0.88%. This fuel oil supplements refinery off-gas to meet refinery fuel demand.

FCC regenerator stack gas containing 1300 ppm SO_x by volume on a dry basis is scrubbed in a non-regenerative lime slurry system to reduce these emissions to a regulation maximum of 250 ppm.

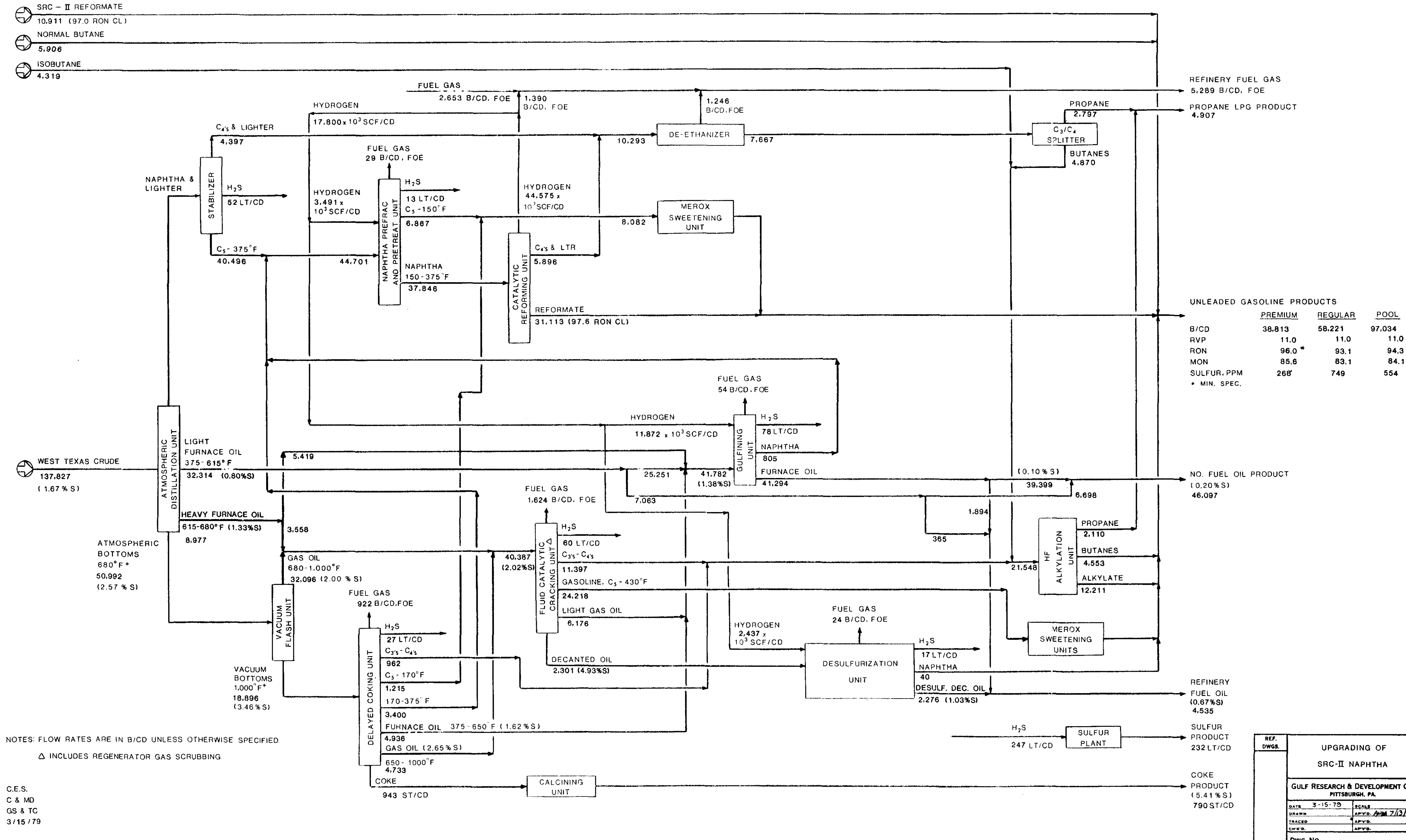
Hydrogen sulfide recovered from the desulfurization, FCC, coking, and naphtha pretreating units is converted to sulfur product in a conventional Claus unit equipped with tail gas desulfurization facilities.

4.2 Refinery with SRC-II Reformate for Crude Replacement (Case II)

In this scheme, shown in Figure 3, reformate, 10,911 B/CD, from upgrading of SRC-II naphtha is blended with refinery gasoline components at reduced production rates so that total gasoline production remains the same as in the Base Case refinery, 97,036 B/CD. West Texas crude oil charge rate is reduced from 150,000 B/CD to 137,827 B/CD with charge rates to the catalytic reforming and coking units correspondingly reduced. Charge rate to the FCC unit is reduced by an additional extent by backing out a portion of 615-680°F heavy furnace oil from the feedstock. The heavy furnace oil is then included in Gulfining unit feedstock to result in the same production rate of No. 2 fuel oil as in the Base Case refinery, 46,097 B/CD. Alkylation unit charge rate is reduced by the same extent as FCC charge.

Reforming unit severity is the same as in the Base Case refinery, with 97.6 RON clear debutanized reformate. However, to meet minimum gasoline Motor octane number specifications, more "give away" in Research octane number is required in both gasoline grades in this scheme than in the Base Case.

FIGURE 3
UPGRADING OF SRC - II NAPHTHA TO GASOLINE
WEST TEXAS CRUDE REFINING SCHEME
CASE II - SRC - II REFORMATE FOR CRUDE REPLACEMENT
SCHEMATIC FLOW DIAGRAM



NOTES: FLOW RATES ARE IN B/CD UNLESS OTHERWISE SPECIFIED
△ INCLUDES REGENERATOR GAS SCRUBBING

C.E.S.
C & MD
GS & TC
3/15/79

4.3 Refinery with SRC-II Reformate for Additional Gasoline (Case III)

In this scheme, shown in Figure 4, crude oil rate and charge rates to major downstream processing units are the same as in the Base Case refinery. SRC-II reformate is blended with gasoline components, almost totally in the premium grade, to result in about a 14% increase in gasoline production, 110,443 B/CD versus 97,036 B/CD in the Base Case. The required reforming severity, 96.8 RON clear debutanized reformate, to meet minimum gasoline octane specifications is reduced from that in the Base Case refinery, 97.6 RON clear. The higher yield of reformate at this reduced severity level accounts for a part of the overall increase in gasoline production. The large increase in normal butane blending requirement over the Base Case operation, resulting from blending of low vapor pressure SRC-II reformate, 1.8 TVP, into the gasoline pool, also contributes to increased gasoline production.

Production of No. 2 fuel oil, 45,643 B/CD, is about 1.0% lower than in Base Case refinery, as a result of the need for more liquid refinery fuel.

4.4 Refinery with SRC-II Reformate for Additional No. 2 Fuel Oil (Case IV)

In this scheme, shown in Figure 5, crude oil charge rate is the same as in the Base Case refinery, but the FCC unit charge rate is significantly reduced by backing out all heavy furnace oil (615-680°F) from the feedstock. Heavy furnace oil is processed for No. 2 fuel oil production, which is increased by 15% over the Base Case to 53,150 B/CD. Loss in FCC gasoline and alkylate from feedstock reduction is more than made up with SRC-II reformate, with gasoline production increased by 4% over the Base Case to 100,965 B/CD. Reforming severity in this scheme, 97.2 RON clear debutanized reformate, is slightly lower than in the Base Case refinery, 97.6 RON clear.

FIGURE 4
UPGRADING OF SRC - II NAPHTHA TO GASOLINE
WEST TEXAS CRUDE REFINING SCHEME
CASE III SRC-II REFORMATE FOR ADDITIONAL GASOLINE
SCHEMATIC FLOW DIAGRAM

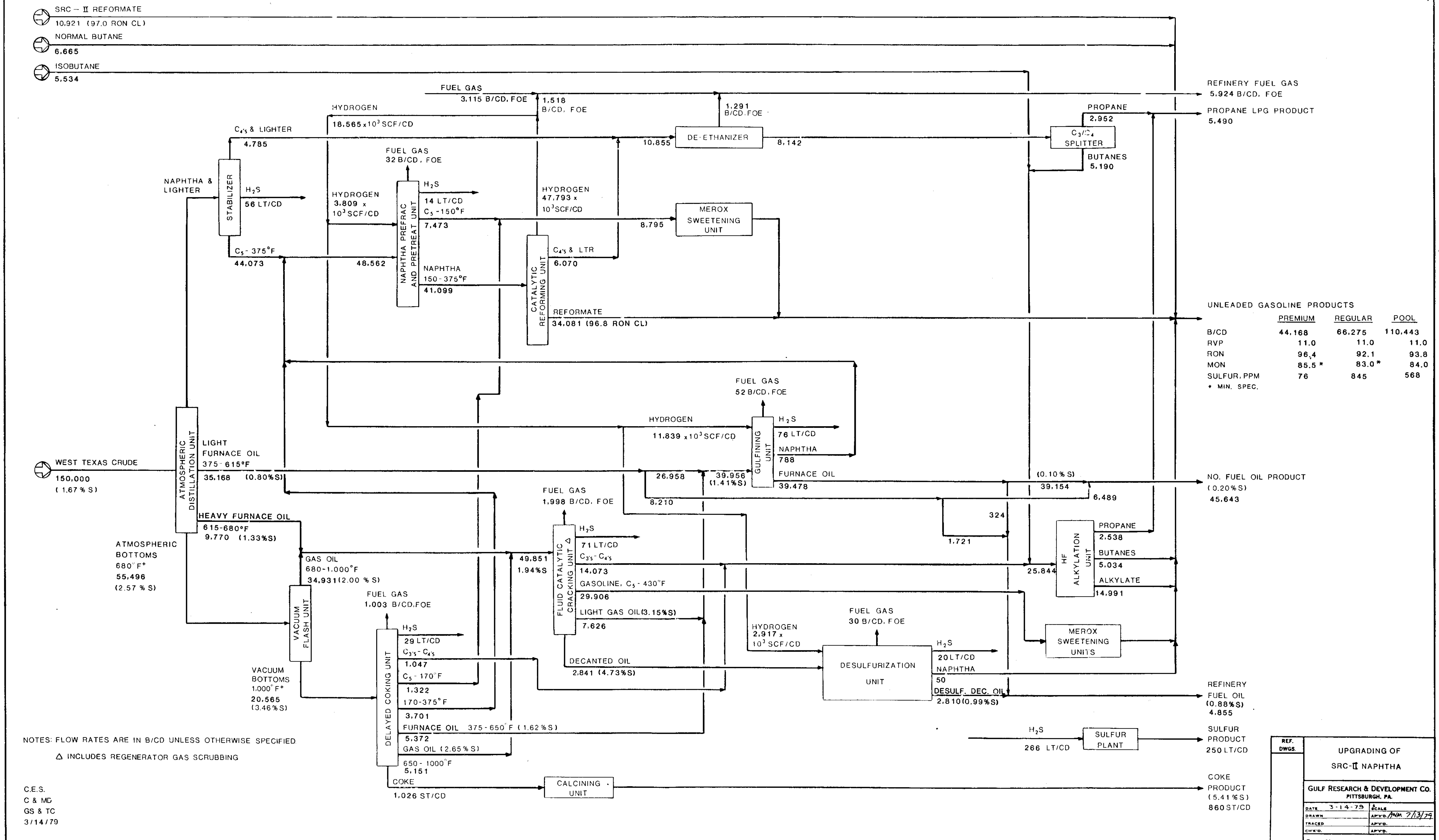


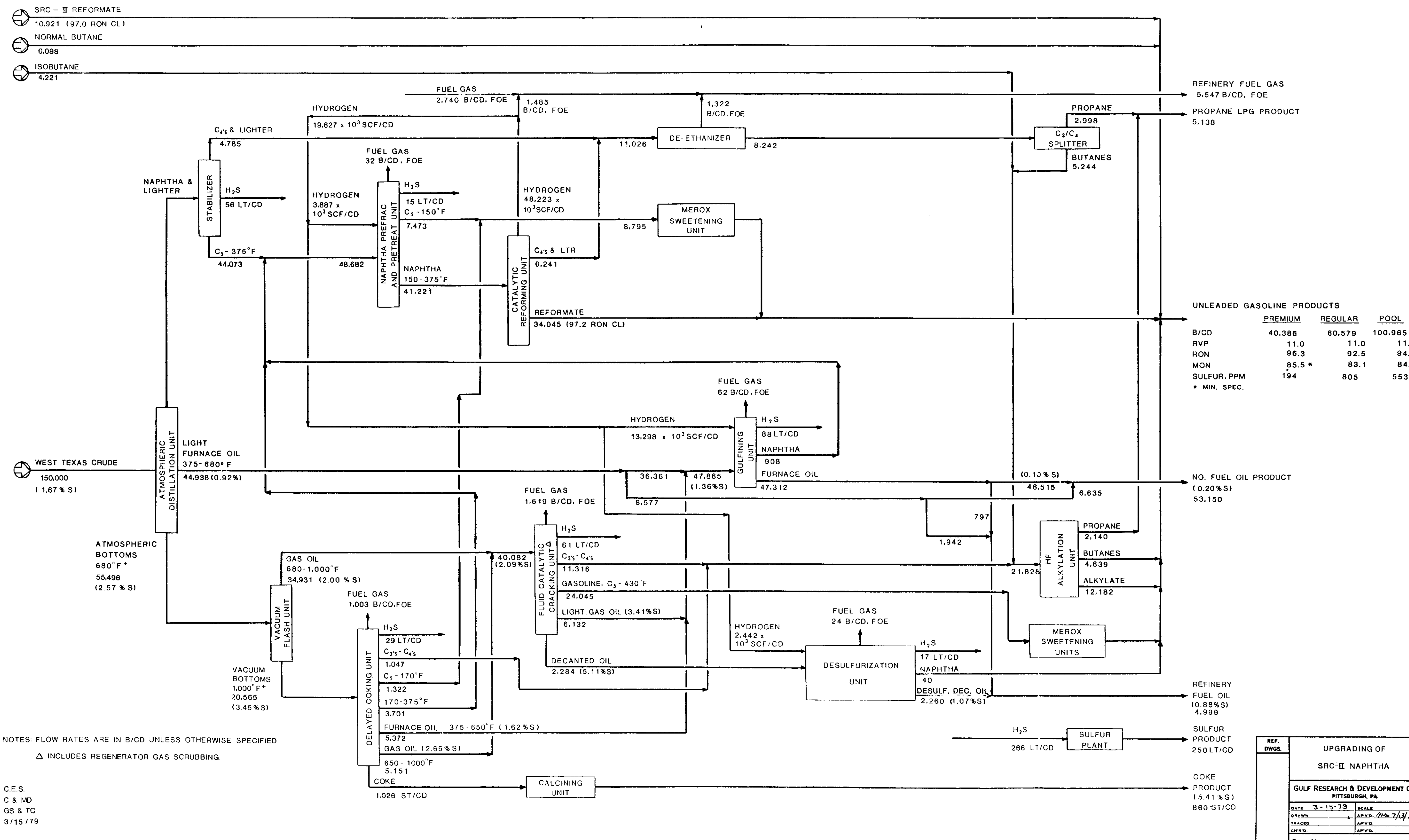
FIGURE 5

UPGRADING OF SRC - II NAPHTHA TO GASOLINE

WEST TEXAS CRUDE REFINING SCHEME

CASE IV - SRC - II REFORMATE FOR ADDITIONAL NO. 2 FUEL OIL

SCHEMATIC FLOW DIAGRAM



5.0 KUWAIT CRUDE REFINING SCHEMES

5.1 Base Case Refinery (Case I-A)

Kuwait crude is charged at a rate of 150,000 B/CD in a Base Case refining scheme, shown in Figure 6, for maximum conversion of bottoms to furnace oil and lighter products. Crude unit atmospheric bottoms (680°F+) is charged to a Gulf Type V HDS unit at a rate of 71,055 B/CD to produce 69,773 B/CD of 375°F+ bottoms containing 0.30% sulfur for charge to a riser FCC unit. The FCC unit employs a high activity zeolite catalyst at a relatively high make-up rate of 0.54 lb/B to maintain a 1000 ppm nickel equivalent metals concentration on equilibrium catalyst to off-set deactivation which would otherwise occur from the high metals content, 1.68 ppm nickel equivalent, fresh feed. A conversion of 82.9 vol % to gasoline, 430°F TBP end point, and lighter products is obtained. Regenerator stack gas is scrubbed with a lime solution to reduce SO_x emissions to a maximum specification of 250 ppm by volume on a dry basis.

Straight-run and HDS naphthas (160-375°F) are reformed over UOP R-18 bimetallic catalyst to produce debutanized reformate having 96.8 RON clear, as required for unleaded gasoline products meeting octane numbers forecast for 1990. Total gasoline production is 105,453 B/CD.

No. 2 fuel oil and refinery fuel oil processing is similar to that described previously for West Texas crude. Straight-run furnace oil is fractionated at a maximum boiling range (375-680°F) to obtain maximum yield consistent with No. 2 fuel oil product distillation specifications. No. 2 fuel oil production is 44,464 B/CD. The 0.88% maximum sulfur specification for refinery fuel oil can be met without desulfurization of FCC decanted oil as a result of FCC feedstock desulfurization.

Hydrogen is produced from FCC of high-boiling residue feedstock at a yield of 0.30 wt %, which is almost four times as great as that from a conventional gas oil feedstock. Ninety-five percent of this hydrogen is recovered

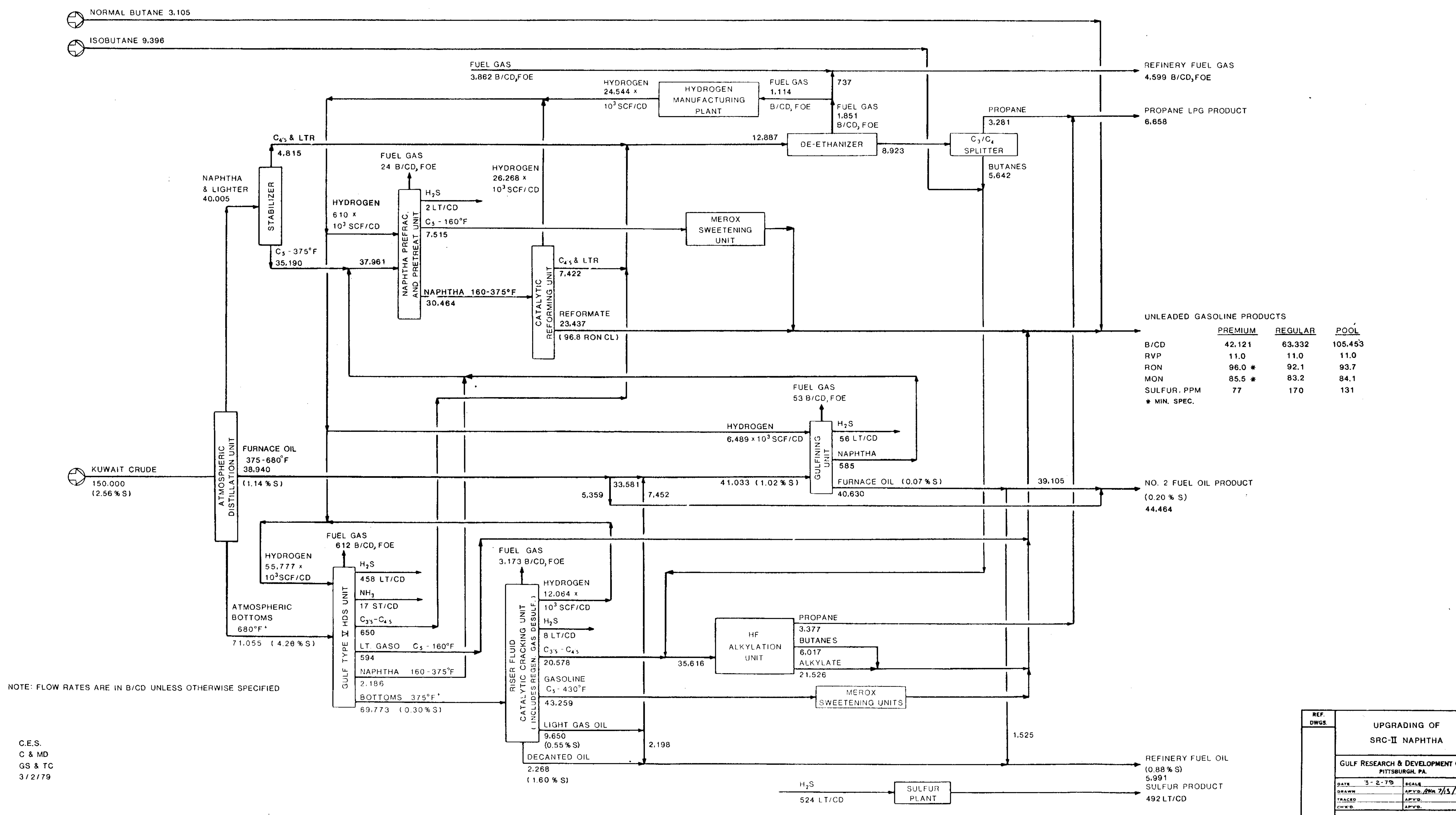
FIGURE 6

UPGRADING OF SRC - II NAPHTHA TO GASOLINE

CASE IA - BASE CASE - KUWAIT CRUDE REFINING SCHEME

TYPE V HDS OF ATMOSPHERIC BOTTOMS PLUS FCC OF HDS BOTTOMS

SCHEMATIC FLOW DIAGRAM



cryogenically to produce 12.0 x 10^6 SCF/CD of hydrogen in a stream of 93% purity. This quantity of hydrogen plus platformer hydrogen is supplemented with 24.5 x 10^6 SCF/CD of hydrogen manufactured by steam reforming of refinery off-gas to meet the requirements for the HDS, Gulfining, and naphtha pretreating units.

5.2 Refineries with SRC-II Reformate For Additional Gasoline (Case III-A)

In these schemes, shown in Figures 7 and 8, charge rates to the crude distillation, HDS, and FCC units are the same as in the Base Case refinery. With the addition of 10,921 B/CD of SRC-II reformate having 97.0 RON clear to the gasoline pool (Figure 7), naphtha reforming severity is reduced from 96.8 RON clear debutanized reformate in the Base Case to 96.1 RON clear. Gasoline production of 118,730 B/CD is obtained at an increase of 13% over that in the Base Case. Production of No. 2 fuel oil, 44,029 B/CD, is 1.0% lower than in the Base Case, as a result of an increase in refinery fuel oil requirement. As a result of the lower reforming unit severity and lower hydrogen by-product rate, hydrogen manufacturing capacity, 24.9 x 10^6 SCF/CD, is slightly greater than in the Base Case.

With the addition of 16,548 B/CD of reformate and 2681 B/CD of light hydrotreated gasoline obtained when charging Powhatan coal to the SRC-II plant (Figure 8), gasoline production in the Kuwait crude scheme is increased 22% over that in the Base Case to 128,530 B/CD. Naphtha-reforming severity is reduced from 96.8 to 95.2 RON clear debutanized reformate.

6.0 ECONOMICS

Economics for upgrading SRC-II naphtha and blending the high octane reformate with petroleum-derived gasoline were developed on a current time frame basis for comparison with the economics of other SRC-II projects. Costs of upgrading are as follows:

FIGURE 7
UPGRADING OF SRC - II NAPHTHA TO GASOLINE
KUWAIT CRUDE REFINING SCHEME
CASE III - A - SRC - II NAPHTHA FOR ADDITIONAL GASOLINE
SCHEMATIC FLOW DIAGRAM

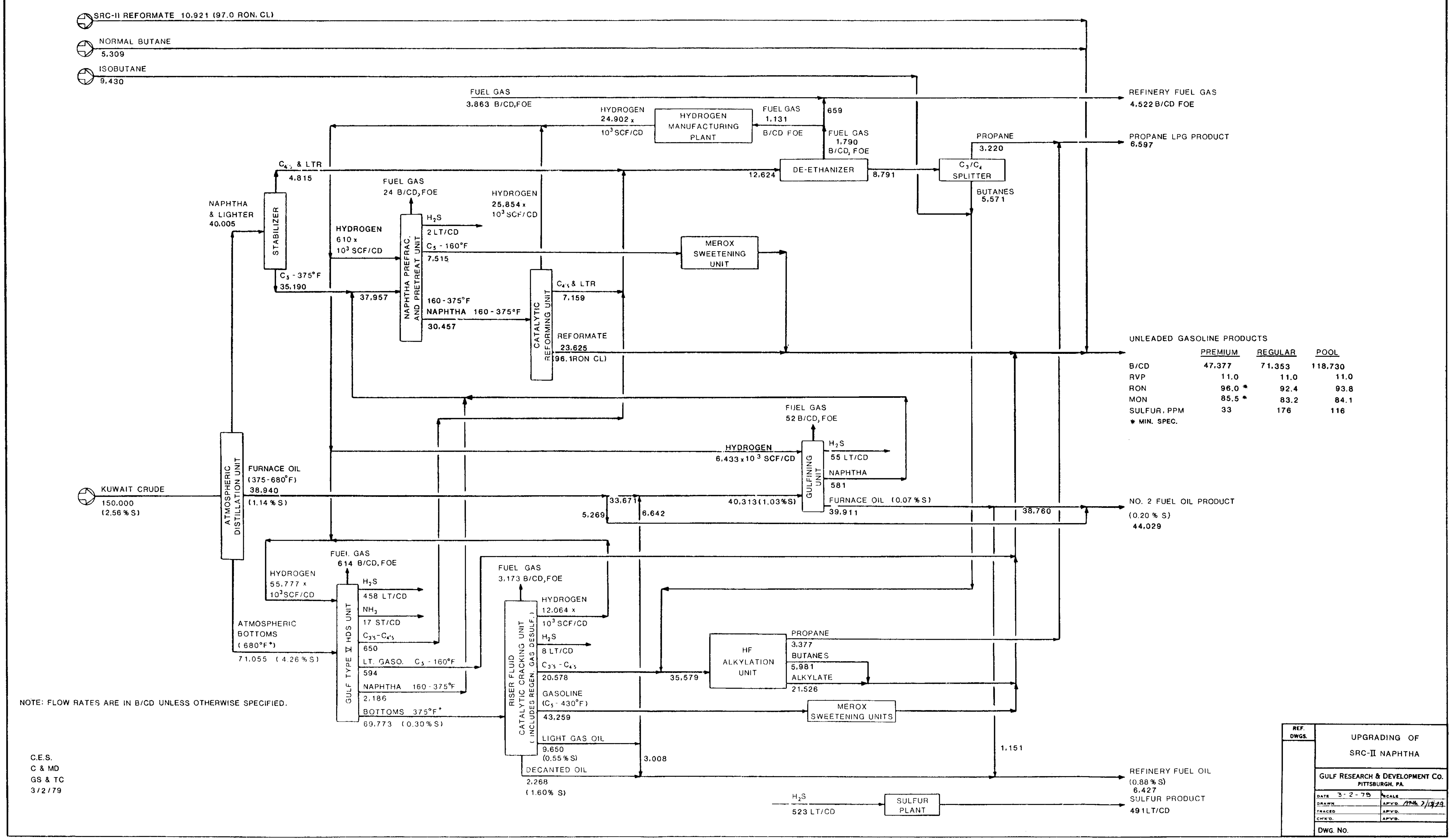
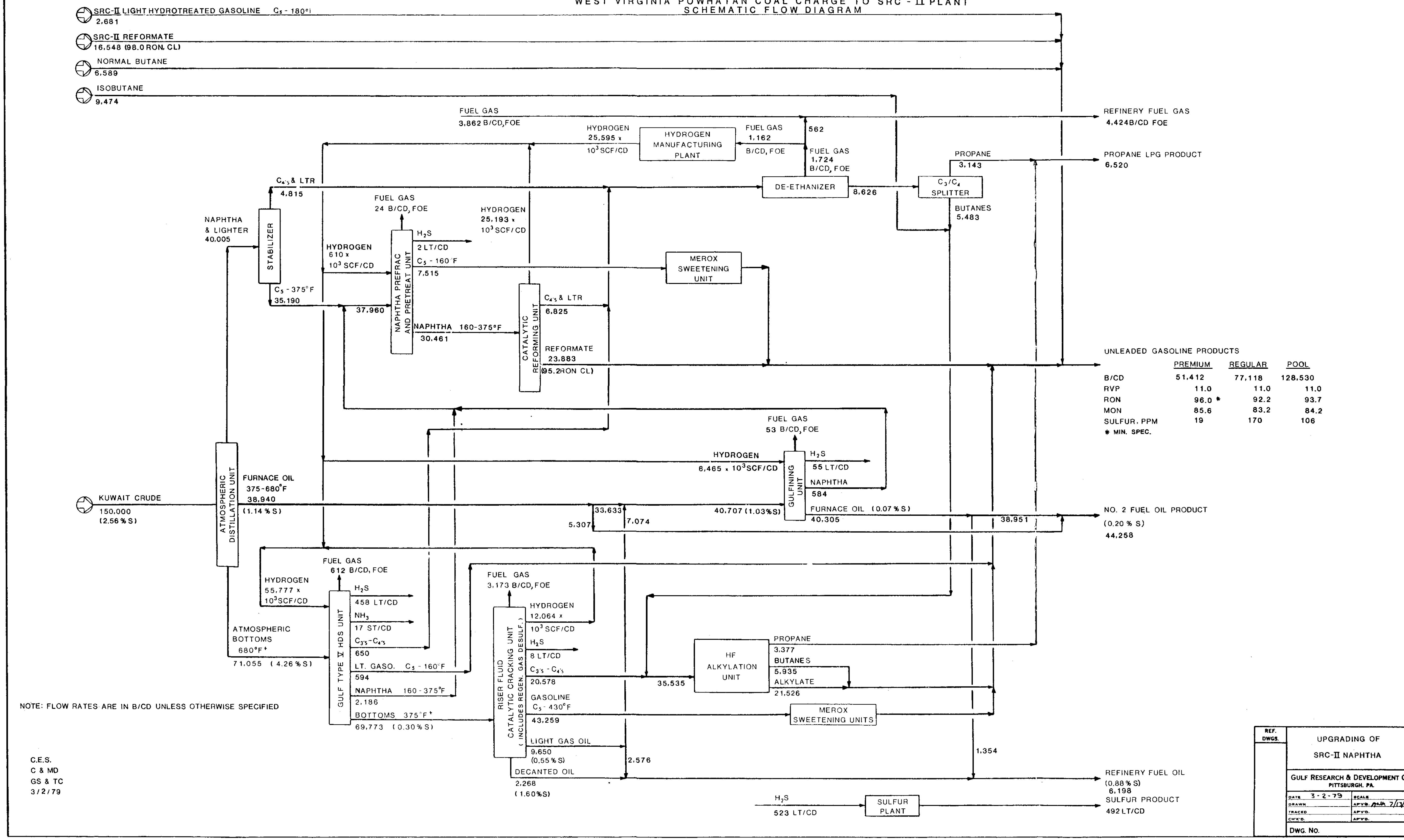


FIGURE 8

UPGRADING OF SRC - II NAPHTHA TO GASOLINE
KUWAIT CRUDE REFINING SCHEME
CASE III - A - 1 - SRC - II NAPHTHA FOR ADDITIONAL GASOLINE
WEST VIRGINIA POWHATAN COAL CHARGE TO SRC - II PLANT
SCHEMATIC FLOW DIAGRAM



Coal Source	Pittsburgh Seam	West Virginia Powhatan
Raw SRC-II Naphtha (C_5-350°F), B/CD	11,961	16,200
	Investment $\$10^6$ (1978)	
Process Unit	18.1	40.4
Catalysts and Royalties	3.6	7.7
Off-Sites Facilities	16.2	26.0
Total Plant Investment	47.9	74.1
Working Capital	5.0	9.9
Total Capital Required	52.9	84.0
Additional Cost of Upgrading	$/B SRC-II Naphtha Charge (1979)	
Total Manufacturing Expense	3.28	2.79
Cost of Capital Before Tax, 20%	2.43	1.94
Total Mfg. Expense plus Cost of Capital	5.71	4.73

Costs are included in the hydrotreating units for a stripper to recover ammonia and hydrogen sulfide from sour water from the low-pressure separator for incremental capacity in the existing ammonia plant to recover additional ammonia. The economy-of-scale from upgrading naphtha at the higher production rate results in a reduction in manufacturing expense plus cost of capital of about $1.00/B charge.

Economics for the four refining schemes charging West Texas crude and for the three schemes charging Kuwait crude are summarized in Tables II and III, respectively. The price of crude oil was calculated to provide 20% pre-tax return on capital in the Base Case refineries charging West Texas and Kuwait crudes, respectively. Economics for the naphtha upgrading facilities at the hypothetical SRC-II site are integrated with those for the hypothetical refinery. A pipeline tariff for transporting reformate from the SRC-II plant site to the refinery at an assumed distance of 300 miles is calculated at $0.22-0.30/B based on literature information.(3)

As shown in Table II, the calculated values of raw SRC-II naphtha when added to the West Texas crude refinery gasoline pool for crude replacement (Case II), increased gasoline production (Case III), or increased No. 2

TABLE II

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

WEST TEXAS CRUDE REFINING SCHEMES

ECONOMIC EVALUATION
1979 START-UP, THIRD QUARTER 1978 INVESTMENT, NRC=714

CASE	I BASE CASE	II SRC-II NAPHTHA FOR CRUDE REPLACEMENT	III SRC-II NAPHTHA FOR ADDITIONAL GASOLINE	IV SRC-II NAPHTHA FOR ADDITIONAL NO.2 FUEL OIL
COAL CHARGE TO SRC-II PLANT	-	PITTSBURGH SEAM	PITTSBURGH SEAM	PITTSBURGH SEAM
SRC-II NAPHTHA, C5-350 F, B/CD	-	11950.	11961.	11961.
PLANT CAPACITY, B/CD OF CRUDE	150000.	137827.	150000.	150000.
ESTIMATED INVESTMENT, THOUS $				
REFINERY PROCESS UNITS	265630.	291580.	291340.	293560.
CATALYSTS AND ROYALTIES	9310.	11590.	12580.	12090.
UTILITY FACILITIES	11840.	11850.	12480.	12130.
STORAGE TANKS	29560.	32410.	33590.	33420.
MISCELLANEOUS OFF-SITE FACILITIES	101320.	110830.	111350.	111910.
SUBTOTAL PLANT INVESTMENT	417660.	458260.	461340.	463110.
CONTINGENCY AT 10.%	41770.	45830.	46130.	46310.
TOTAL PLANT INVESTMENT	459430.	504090.	507470.	509420.
WORKING CAPITAL	78730.	77530.	84490.	83510.
TOTAL CAPITAL REQUIREMENT	538160.	581620.	591960.	592930.

RETURN FROM PRODUCTS		UNITS/CD	$THOUS/YEAR	UNITS/CD	$THOUS/YEAR	UNITS/CD	THOUS$/YEAR	UNITS/CD	$THOUS/YEAR
REGULAR GASOLINE,	$ 24.464/B (1)	58222.	519881.	58221.	519873.	66275.	591790.	60579.	540928.
PREMIUM GASOLINE,	$ 24.464/B (1)	38814.	346587.	38813.	346579.	44168.	394393.	40386.	360620.
PROPANE LPG,	$ 15.576/B (1)	5583.	31738.	4907.	27897.	5490.	31212.	5138.	29211.
REFINERY FUEL GAS,	$ 16.841/B FOE (1)	6020.	37007.	5971.	36705.	6603.	40590.	6228.	38283.
SULFUR,	$ 30.000/LONG TON	251.	2746.	235.	2572.	253.	2776.	253.	2776.
NO. 2 FUEL OIL,	$ 23.844/B (1)	46097.	401181.	46097.	401187.	45643.	397234.	53150.	462568.
REFINERY FUEL OIL,	$ 16.841/B (1)	4399.	27041.	4535.	27877.	4855.	29846.	4999.	30729.
COKE,	$ 80.000/SHORT TON	860.	25099.	790.	23063.	860.	25099.	860.	25099.
HYDROGEN,	$ 1.750/THOUS SCF	-	-	5801.	3705.	5806.	3709.	5806.	3709.
AMMONIA,	$120.00/SHORT TON	-	-	11.	482.	11.	482.	11.	482.
TOTAL RETURN FROM PRODUCTS			1391280.		1389940.		1517131.		1494405.
COST OF CHARGE									
WEST TEXAS CRUDE,	$19.669/B (2)	150000.	1076859.	137827.	989468.	150000.	1076859.	150000.	1076859.
ISOBUTANE,	$21.152/B (1)	5479.	42300.	4319.	33349.	5534.	42722.	4221.	32584.
NORMAL BUTANE,	$17.612/B (1)	4426.	28454.	5906.	37963.	6665.	42846.	6098.	39198.
SRC-II NAPHTHA		-	-	11950.	68516. (3)	11961.	84721. (3)	11961.	77236. (3)
TOTAL COST OF CHARGE			1147613.		1129296.		1247148.		1225877.
MANUFACTURING EXPENSE									
INVESTMENT-BASED ITEMS (4)			53128.		58267.		58617.		58864.
LABOR-BASED ITEMS, $10.78/HR (5)		912.	8710.	1068.	9627.		9627.		9627.
UTILITIES									
REFINERY FUEL OIL,	$16.841/B	4399.	27041.	4535.	27877.	4855.	29846.	4999.	30729.
REFINERY FUEL GAS,	$16.841/B FOE	6020.	37007.	5971.	36705.	6603.	40590.	6228.	38283.
POWER,	$ 0.0303/KWH	599921.	6635.	601577.	6653.	650699.	7196.	646927.	7155.
FRESH WATER,RIVER,	$ 0.00/THOUS GAL	5817.	0.	5616.	0.	6259.	0.	5925.	0.
SUBTOTAL, UTILITIES			70683.		71235.		77632.		76167.
TRANSPORTATION OF REFORMATE,	$0.302/B	-	-	10911.	1202.	10921.	1203.	10921.	1203.
CHEMICALS			1160.		1060.		1197.		1117.
CATALYSTS			2218.		2803.		3177.		2827.
ROYALTY, CRUDE DESALTING,	$0.0025/B	150000.	137.	137827.	126.	150000.	137.	150000.	137.
TOTAL MANUFACTURING EXPENSE			136036.		144320.		151590.		149942.
TOTAL EXPENSE			1283649.		1273616.		1398738.		1375819.
COST OF CAPITAL, BEFORE TAX, 20%			107631.		116324.		118393.		118586.
CALCULATED VALUE OF SRC-II NAPHTHA, $/B ($/MILL'N BTU)			-		15.708(2.88)		19.406(3.56)		17.691(3.24)

1) FORECAST PRICES IN 1990 DEFLATED TO 1979.
2) CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX IN BASE CASE-CASE 1.
3) CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX.
4) INCLUDES DEPRECIATION,MAINTENANCE,INVESTMENT BURDEN,INSURANCE AND TAXES.
5) INCLUDES DIRECT OPERATING WAGES,SUPERVISION, AND OVERHEAD.

TABLE III

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

KUWAIT CRUDE REFINING SCHEMES

ECONOMIC EVALUATION
1979 START-UP, THIRD QUARTER 1978 INVESTMENT, NRC=714

CASE	I-A BASE CASE	III-A SRC-II NAPHTHA FOR ADDITIONAL GASOLINE	III-A-1 SRC-II NAPHTHA FOR ADDITIONAL GASOLINE
COAL CHARGE TO SRC-II PLANT	-	PITTSBURGH SEAM	WEST VIRGINIA POWHATAN
SRC-II NAPHTHA, C5-350 F, B/CD	-	11961.	16200.
PLANT CAPACITY, B/CD OF CRUDE	150000.	150000.	150000.
ESTIMATED INVESTMENT, THOUS $			
REFINERY PROCESS UNITS	325490.	351400.	362920.
CATALYSTS AND ROYALTIES	16540.	19800.	22540.
UTILITY FACILITIES	14130.	14770.	16940.
STORAGE TANKS	35680.	39700.	44030.
MISCELLANEOUS OFF-SITE FACILITIES	123850.	133940.	139880.
SUBTOTAL PLANT INVESTMENT	515690.	559610.	586310.
CONTINGENCY AT 10.%	51570.	55960.	58630.
TOTAL PLANT INVESTMENT	567260.	615570.	644940.
WORKING CAPITAL	83970.	89720.	95080.
TOTAL CAPITAL REQUIREMENT	651230.	705290.	740020.

		I-A UNITS/CD	I-A THOUS$/YEAR	III-A UNITS/CD	III-A THOUS$/YEAR	III-A-1 UNITS/CD	III-A-1 THOUS$/YEAR
RETURN FROM PRODUCTS							
REGULAR GASOLINE,	$ 24.464/B (1)	63332.	565518.	71353.	637137.	77118.	688610.
PREMIUM GASOLINE,	$ 24.464/B (1)	42121.	376117.	47377.	423046.	51412.	459074.
PROPANE LPG,	$ 15.576/B (1)	6658.	37853.	6597.	37506.	6520.	37069.
JET FUEL,	$ 24.25/B	-	-	-	-	3583.	31714.
NO. 2 FUEL OIL,	$ 23.844/B (1)	44464.	386976.	44029.	383186.	44258.	385179.
SULFUR,	$ 30.000/LONG TON	492.	5385.	494.	5413.	500.	5475.
AMMONIA,	$120.00/SHORT TON	17	745.	28.	1226.	35.	1533.
HYDROGEN,	$ 1.750/THOUS SCF	-	-	5806.	3709.	1550.	990.
GAS FEED TO H2 PLANT,	$ 16.841/B FOE (1)	1114.	6847.	1131.	6952.	1162.	7143.
REFINERY FUEL GAS,	$ 16.841/B FOE (1)	4599.	28270.	5203.	31982.	5189.	31897.
REFINERY FUEL OIL,	$ 16.841/B (1)	5991.	36829.	6427.	39507.	6198.	38099.
TOTAL RETURN FROM PRODUCTS			1444540.		1569664.		1686783.
COST OF CHARGE							
KUWAIT CRUDE,	$19.137/B (2)	150000.	1047770.	150000.	1047770.	150000.	1047770.
ISOBUTANE,	$21.152/B (1)	9396.	72545.	9430.	72805.	9474.	73141.
NORMAL BUTANE,	$17.612/B (1)	3105.	19962.	5309.	34130.	6589.	42356.
SRC-II NAPHTHA,C5-350F		-	-	11961.	84199. (3)	16200.	120794. (3)
SRC-II DISTILLATE,350-400F,	$19.860/B (4)	-	-	-	-	7560.	54802.
TOTAL COST OF CHARGE			1140277.		1238904.		1338863.
MANUFACTURING EXPENSE							
INVESTMENT-BASED ITEMS (5)			65472		70994.		74290.
LABOR-BASED ITEMS, $10.78/HR (6)		888.	8481.	984.	9399.	984.	9399.
UTILITIES							
REFINERY FUEL GAS	$16.841/B FOE	4599.	28270.	5203.	31982.	5189.	31897.
REFINERY FUEL OIL	$16.841/B	5991.	36829.	6427.	39507.	7072.	43471.
POWER	$ 0.0303/KWH	947644.	10480.	998250.	11040.	1023420.	11319.
FRESH WATER, RIVER	$ 0.00/THOUS GAL	7063.	0.	7507.	0.	7792.	0.
SUBTOTAL, UTILITIES			75579.		82529.		86687.
GAS FEED TO H2 PLANT	$16.841/B FOE	1114.	6847.	1131.	6952.	1162.	7143
TRANSPORTATION OF REFORMATE		-	-	10921.	1203.	19229.	1551.
CHEMICALS			1315.		1342.		1363.
CATALYSTS			16187.		17147.		19347.
ROYALTY, CRUDE DESALTING,	$0.0025/B	150000.	137.	150000.	137.	150000.	137.
TOTAL MANUFACTURING EXPENSE			174018.		189703.		199917.
TOTAL EXPENSE			1314295.		1428607.		1538780.
COST OF CAPITAL, BEFORE TAX, 20%			130245		141057		148003
CALCULATED VALUE OF SRC-II NAPHTHA, $/B ($/MILL'N BTU)		-		19.286(3.54)		20.429(3.78)	

1) FORECAST PRICES IN 1990 DEFLATED TO 1979.
2) CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX IN BASE CASE-CASE 1-A
3) CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX.
4) CALCULATED AT $3.50/MILL'N BTU.
5) INCLUDES DEPRECIATION, MAINTENANCE, INVESTMENT BURDEN, INSURANCE AND TAXES.
6) INCLUDES DIRECT OPERATING WAGES, SUPERVISION, AND OVERHEAD.

fuel oil production (Case IV), are \$15.71/B, \$19.41/B, or \$17.69/B, respectively. These values include a cost of capital before taxes of 20% for the upgrading and incremental refining facilities, the same cost of capital used in calculating crude oil value. The highest calculated value of SRC-II naphtha, \$19.41/B, is obtained on the basis of producing additional gasoline and compares favorably with the calculated crude price of \$19.67/B. On the bases of crude replacement or increased production of No. 2 fuel oil, values of SRC-II naphtha become significantly lower as a result of operating the crude and/or major downstream refining units considerably below design capacities.

As shown in Table III, the calculated values of raw SRC-II naphtha when added to the Kuwait crude refinery gasoline pool for increased gasoline production are \$19.29/B (\$3.54/million Btu) and \$20.43/B (\$3.78/million Btu) for the lower and higher naphtha production rates, respectively. These values are higher than the calculated crude price of \$19.14/B (\$3.33/million Btu). The higher value of raw naphtha produced at the higher rate reflects economy-of-scale associated with the capital intensive upgrading facilities.

As shown by the values of SRC-II naphtha calculated in the West Texas and Kuwait crude refining schemes, respectively, the relative values of SRC-II naphtha with respect to petroleum increase as the gravity and sulfur content of petroleum increase. A trend to production of heavier and higher sulfur crude is expected, since light crude oils comprise only 49% of the known non-Communist world's crude reserves, yet they represent about 62% of its production.

The calculated value of SRC-II naphtha, produced at the rate of 11,961 B/CD, in the Kuwait crude refining scheme is summarized in Table IV as the difference between the economics of the two refinery schemes, Case III-A minus Case I-A. Total upgrading manufacturing expense plus 20% return on total upgrading capital before tax is equal to \$6.07/B of raw SRC-II naphtha. Capital cost included in this total, \$3.74/B of raw SRC-II naphtha, is the largest factor in depressing the value of this naphtha below what might otherwise be expected for a high-quality gasoline component. The unit investment

TABLE IV

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

KUWAIT CRUDE REFINING SCHEMES

ECONOMIC SUMMARY - 1979 START-UP
CASE III-A MINUS CASE I-A

CAPACITY, SRC-II NAPHTHA, C5-350F B/CD	11961
ESTIMATED INVESTMENT, THOUS $ (1978)	INVEST-MENT
SRC-II NAPHTHA UPGRADING UNITS	25560
REVAMPS TO REFINERY PROCESS UNITS	350
CATALYSTS AND ROYALTIES	3260
UTILITIES AND OFF-SITES	14750
SUB-TOTAL PLANT INVESTMENT	43920
CONTINGENCY AT 10. %	4390
TOTAL PLANT INVESTMENT	48310
WORKING CAPITAL	5750
TOTAL CAPITAL REQUIREMENT	54060

RETURN FROM PRODUCTS		UNITS/CD	THOUS$/YEAR	$/B SRC-II NAPHTHA
GASOLINES,	$ 24.464/B	13277	118548	27.15
PROPANE LPG,	$ 15.576/B	-61	-347	-0.08
NO.2 FUEL OIL,	$ 23.844/B	-435	-3790	-0.87
SULFUR,	$ 30.000/LONG TON	2	28	0.01
HYDROGEN,	$ 1.750/THOUS SCF	5806	3709	0.85
AMMONIA,	$120.000/SHORT TON	11	481	0.11
GAS FEED TO H2 PLANT,	$ 16.841/B FOE	17	105	0.03
REFINERY FUEL,	$ 16.841/B FOE	1040	6390	1.46
TOTAL RETURN FROM PRODUCTS			125124	28.66
COST OF BUTANES				
ISOBUTANE,	$ 21.152/B	34	260	0.06
NORMAL BUTANE,	$ 17.612/B	2204	14168	3.24
TOTAL COST OF BUTANES			14428	3.30
MANUFACTURING EXPENSE				
GAS FEED TO H2 PLANT,	$ 16.841/B FOE	17	105	0.03
REFINERY FUEL,	$ 16.841/B FOE	1040	6390	1.46
POWER,	$ 0.0303/KWH	50606	560	0.13
CATALYSTS,			960	0.22
CHEMICALS,			27	0.01
TRANSPORTATION OF REFORMATE,	$0.302/B	10921	1203	0.27
INVESTMENT - BASED ITEMS			5522	1.26
LABOR - BASED ITEMS			918	0.21
TOTAL MANUFACTURING EXPENSE			15685	3.59
COST OF CAPITAL, BEFORE TAX, 20%			10812	2.48
VALUE OF SRC-II NAPHTHA = PRODUCT RETURN -COST OF BUTANES - TOTAL MFG. EXPENSE - PROFIT			84199	19.29

for naphtha hydrotreating at $930 per daily barrel is about two and one-half times that for a large (30,000 B/SD) conventional naphtha pretreating unit, $380 per daily barrel.

If the investment estimated for the 11,961 B/CD SRC-II naphtha upgrading complex (hydrotreater and reformer plus associated off-sites) were decreased by 25%, the value of the SRC-II naphtha would be increased from $19.29/B to $20.18/B. Conversely, if the investment of the upgrading complex were increased by 25%, the value of the naphtha would be decreased by a like amount, from $19.29/B to $18.40/B.

If the refinery were located at a distance of 1500 miles from the SRC-II site instead of the 300 miles assumed, costs for transporting reformate would increase about five-fold and would result in decreasing the value of SRC-II naphtha by $1.08/B.

7.0 ECONOMIC BASES

Prices for gasoline, No. 2 fuel oil, propane, butanes, and refinery fuel (residual fuel oil with 1.0% sulfur guarantee) are those forecast for 1990 and then deflated by 6%/yr. Current market prices are used for calcined, high-sulfur coke, sulfur, and ammonia. The value of excess hydrogen produced from the reforming unit at the SRC-II site is calculated to be $\$1.75/10^3$ SCF for manufacture by partial oxidation of heavy oil, as planned for the projected conceptual commercial SRC-II plant complex.

Investments for naphtha upgrading and refinery process units, utility units, and oil storage tanks were taken from Gulf investment vs. capacity correlations. A contingency factor of 10% is included. Investments for the required expansion of refining units which result from blending of SRC-II reformate into the gasoline products were assumed to be equal to the difference between the investment for a new unit at the expanded capacity and that at the Base Case capacity.

Oil storage capacity is provided as follows:

	Days
Raw SRC-II Naphtha	7
SRC-II Reformate and Light Hydrotreated Gasoline, SRC-II Site	20
SRC-II Reformate and Light Hydrotreated Gasoline, Refinery Site	20
Crude Oil	20
Iso and Normal Butanes Charge to Refinery	5
Gasoline, Jet Fuel, and No. 2 Fuel Oil Products	20
LPG Product	5
Charge to Intermediate Process Units except HDS Unit	7
Charge to HDS Unit	14
Refinery Fuel Oil	10

Investments for miscellaneous off-site facilities, including site preparation, roads, fencing, buildings, communications, etc., were calculated as a percentage of the total investment for process and utility units, plus storage tanks, based on actual costs for a number of Gulf refineries. Paid-up royalties are included for all processing units as required, including those licensed by Gulf.

8.0 CONCLUSIONS

Naphtha produced from the projected commercial SRC-II plant can be upgraded, using currently available technology, to a high octane number gasoline blending component. The calculated value of raw SRC-II naphtha when upgraded and blended into gasoline products from an existing large (150,000 B/CD) petroleum refinery to increase total gasoline production by about 14% compares favorably with the value of crude oil charge to the refinery. This favorable comparison is obtained in spite of the high capital intensity required in the upgrading facilities and the economy-of-scale disadvantage for SRC-II naphtha produced at a relatively low production rate--11,961 B/CD.

Using prices forecast for petroleum products in 1990, the projected time of the conceptual commercial SRC-II plant, and then deflated to the current year, 1979, the value of SRC-II naphtha produced at a rate of 11,961 B/CD is calculated to be $19.41/B (3.56/million Btu) on the basis of increasing gasoline production in a West Texas crude refining scheme. Crude oil price is calculated at $19.67/B ($3.43/million Btu). The corresponding calculated values for naphtha in the Kuwait crude refining schemes are $19.29/B ($3.54/million Btu) at low production rate from Pittsburgh Seam coal and $20.43/B ($3.78/million Btu) at high production rate from West Virginia Powhatan coal. The price of Kuwait crude is calculated at $19.14/B ($3.33/million Btu). The relative values of SRC-II naphtha with respect to petroleum increase as the gravity and sulfur content of petroleum increase, and as the naphtha production rate increases.

SRC-II naphtha from the projected commercial plant could also effect an increase in home heating oil production of about 15%, along with increased gasoline production of 4%, in a large (150,000 B/CD) refinery, or it could maintain production rates of gasoline and home heating oil with an 8% reduction in crude rate. However, the economics for these options would be less favorable as a result of under-utilization of crude and/or major downstream refinery capacity.

References

1. "Refining and Upgrading of Synfuels from Coal and Oil Shales by Advanced Catalytic Processes," Quarterly Report for the Period January-March 1979, R. F. Sullivan, Chevron Research Company, April 1979, Prepared for the United States Department of Energy under Contract No. EF-76-C-01-2315.

2. "High-Value Energy Needs More Efficient Conversion, Transport," A. E. Uhl (Bechtel, Inc.), The Oil and Gas Journal 75, No. 31, August 1, 1977, p. 59.

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Part 2

PART 2

1.0 SUMMARY

The value of SRC-II naphtha (C_5-350°F) produced at a rate of 11,961 B/CD from a projected commercial plant charging 30,000 T/CD of Pittsburgh Seam coal has been calculated in a scheme for upgrading it to a high octane number gasoline blending component. This scheme is similar to one evaluated in Part 1, except that a wide boiling range (C_5-450°F) SRC-II fraction, instead of the naphtha fraction (C_5-350°F) only, would be charged to hydrotreating. Hydrotreating capacity would be about doubled to 27,015 B/SD and the costs are allocated between the naphtha and middle distillate (350-450°F) fractions of the feed. Twenty percent of the middle distillate fraction would be hydrocracked to naphtha boiling below 350°F.

Total plant investments and manufacturing expenses for the naphtha and middle distillate upgrading plants are as follows:

Raw SRC-II Distillate Charge	C_5-350°F	C_5-450°F
Raw SRC-II Distillate Charge, B/CD	11,961	24,314
Total Plant Investment, $\$10^6$ (1978)	47.9	71.4
Total Mfg. Expense plus Cost of Capital, $/B SRC-II Distillate Charge	5.71	4.34

Hydrotreated product would be separated into a naphtha and a middle distillate fraction. Pretreated naphtha would be reformed to a high octane number component, 97.0 RON clear, for blending into gasoline products from a large refinery charging 150,000 B/CD of Kuwait crude for about a 15% increase in gasoline production, as in Part 1. Calculated values of crude oil, raw SRC-II naphtha, and hydrotreated SRC-II middle distillate in 1990 dollars deflated to 1979 are as follows:

Part	Calculated Values for Purpose of Comparison: 1		2	
Charge Stock to Hydrotreating	Naphtha, C_5-350°F		Naphtha plus Middle Distillate, C_5-450°F	
	$/B	$/10^6 Btu	$/B	$/10^6 Btu
Kuwait Crude Oil	19.14	3.33	19.14	3.33
Raw SRC-II Naphtha, C_5-350°F	19.29	3.54	19.20	3.52
Hydrotreated Middle Distillate, 350-450°F	--	--	22.85	4.00

The value of hydrotreated middle distillate, essentially sulfur-free and with a significantly reduced nitrogen content, 0.55%, is calculated from a price of $3.50/million Btu ($19.86/B) estimated for SRC-II product plus allocated costs for hydrotreating, including 20% return on capital before tax.

The value calculated for raw SRC-II naphtha in the integrated naphtha/middle distillate upgrading scheme is slightly lower than that calculated in Part 1 for naphtha upgrading only, but is still greater than that calculated for crude petroleum. The economy-of-scale advantage from the larger hydrotreating unit enables the production of about 20% additional high octane number gasoline blending component (15,558 B/CD including butanes for 11 psi RVP vs 12,903 B/CD) plus 10,040 B/CD of hydrotreated middle distillate with only a slight loss in profitability.

2.0 INTRODUCTION

In Part 1 of this report, it was shown that the value of SRC-II naphtha which could be upgraded by pretreating and reforming to a high octane number gasoline blending component, 97.0-98.0 RON clear, is equal to or greater than that for crude petroleum. High capital costs for the severe pretreating required depress the value of this naphtha from what might be expected for a high-quality gasoline precursor.

In this part, the value of SRC-II naphtha has been calculated in a scheme similar to the one previously evaluated (Case III-A), except that the naphtha pretreating step is integrated with middle distillate hydrotreating. A wide boiling range (C_5-450°F) SRC-II fraction is charged to pretreating instead of the naphtha fraction (C_5-350°F) only. Hydrotreating capacity is about doubled to 27,015 B/SD and the costs are allocated between the naphtha and middle distillate fractions of the feed. Gulf pilot plant data indicate that about 20% of the product from hydrotreating the middle distillate fraction (350-450°F) lies in the naphtha range boiling below 350°F as a result of hydrocracking. The economy-of-scale improvement from increasing hydrotreating capacity and allocating a portion of the costs to upgrading middle distillate, plus the increase in naphtha production from hydrocracking of middle distillate could potentially result in a significant increase in value of raw SRC-II naphtha. Also, an increase in the volume of high value products could be obtained at favorable economics.

As in the previous scheme (Case III-A), high octane number, 97.0 RON clear, SRC-II reformate is blended into gasoline products from a large petroleum refinery charging 150,000 B/CD of Kuwait crude to increase gasoline production.

3.0 SRC-II PRODUCT UPGRADING SCHEME

The processing scheme evaluated to upgrade a wide boiling range (C_5-450°F) SRC-II distillate is shown in Figure 1. Charge capacity of the hydrotreating unit is 27,015 B/SD and comprises naphtha and middle distillate fractions in about a 50/50 volume ratio as follows:

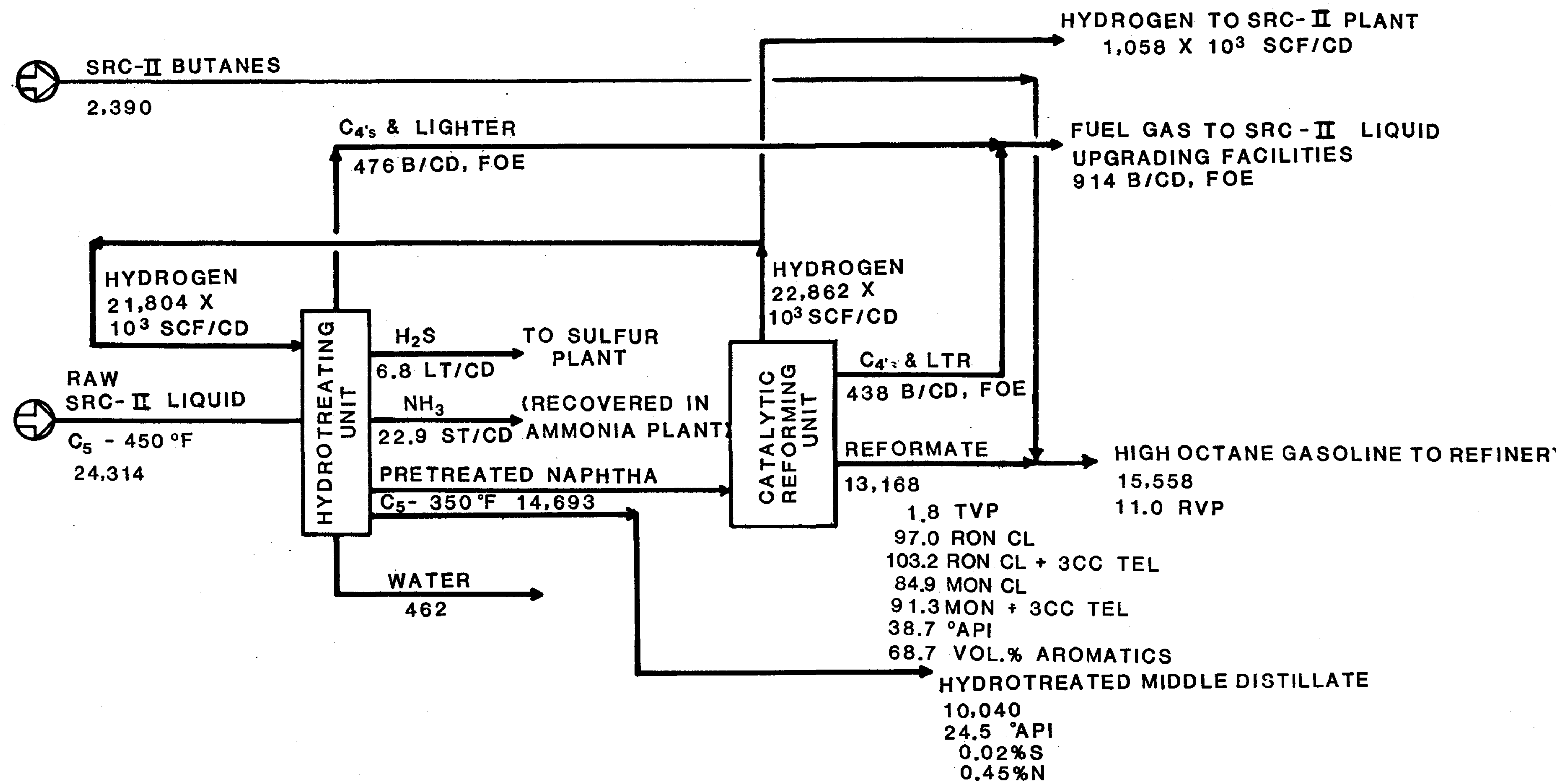
	B/SD
Naphtha, C_5-350°F	13,290
Middle Distillate, 350-450°F	13,725
Total	27,015

The naphtha fraction comprises the total naphtha production from a projected 30,000 T/CD SRC-II plant charging a Pittsburgh Seam coal, while the middle distillate fraction comprises the lower boiling 50% of the 350-600°F middle distillate product from this plant.

FIGURE 1

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

INTEGRATED NAPHTHA/MIDDLE DISTILLATE HYDROTREATING UNIT



NOTE: FLOW RATES ARE IN B/CD UNLESS OTHERWISE SPECIFIED.

C.E.S.
GS&TC
C&MD
4/27/79

REF. DWGS.	UPGRADING OF SRC-II NAPHTHA	
	GULF RESEARCH & DEVELOPMENT CO. PITTSBURGH, PA.	
	DATE 4-27-79	SCALE
	DRAWN	APV'D. ANM 7/13/79
	TRACED	APV'D.
	CH'K'D.	APV'D.
	DWG. NO.	

The operating conditions for the hydrotreating unit are those determined in a Gulf pilot plant study to be required to produce reformer feedstock, as constrained by a nitrogen content of less than 0.2 ppm, and are those used in the previous section for upgrading naphtha only. Under these conditions, the nitrogen content of the middle distillate fraction is indicated from pilot plant data to be reduced from 0.95% to 0.55%, while sulfur content is reduced from 0.2% to 0.02%.

Twenty percent of the liquid product from hydrotreating middle distillate is indicated in Gulf pilot plant work to be hydrocracked to naphtha boiling below 350°F. Hydrogen is consumed at a rate of 900 SCF/B in the naphtha fraction and at a rate of 550 SCF/B in the middle distillate fraction of the feed.

Water is injected into the hydrotreater effluent cooling train to scrub out ammonium bisulfide which forms upon cooling the reaction products. Sour water from the low-pressure separator is charged to a sour water stripper to recover hydrogen sulfide and ammonia by-products. The remainder of the ammonia produced, contained in vapors from the high-pressure separator and the product stripper, is recovered in an existing ammonia plant. Hydrogen sulfide is converted to product sulfur at a rate of 7 LT/CD in an existing sulfur plant. Total ammonia production is 23 ST/CD.

Hydrotreated liquid product is fractionated to recover 14,693 B/CD of C_5-350°F naphtha and 10,040 B/CD of 350-450°F middle distillate. Hydrotreated naphtha is charged to a conventional semi-regenerative reforming unit operating at conditions used in the pilot plant to produce a debutanized reformate having 97.0 RON clear, as in Part 1. Operating conditions and product yields and properties for the reforming unit are also presented in Table I. Reformate is produced at a yield of 89.6 vol% on charge and at a rate of 13,168 B/CD. With the addition of 2390 B/CD of butanes from the SRC-II plant, 15,558 B/CD of 11 psi RVP high octane number gasoline blending component is produced.

TABLE I

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

KUWAIT CRUDE REFINING SCHEMES

ECONOMIC EVALUATION
1979 START-UP, THIRD QUARTER 1978 INVESTMENT, NRC=714

CASE	I-A BASE CASE	III-A SRC-II NAPHTHA FOR ADDITIONAL GASOLINE	III-B SRC-II NAPHTHA FOR ADDITIONAL GASOLINE
COAL CHARGE TO SRC-II PLANT	-	SRC-II C5-350F	SRC-II C5-450F
SRC-II NAPHTHA, C5-350 F, B/CD	-	11961.	16200.
PLANT CAPACITY, B/CD OF CRUDE	150000.	150000.	150000.
ESTIMATED INVESTMENT, THOUS $			
REFINERY PROCESS UNITS	325490.	351400.	363380.
CATALYSTS AND ROYALTIES	16540.	19800.	21780.
UTILITY FACILITIES	14130.	14770.	15460.
STORAGE TANKS	35680.	39700.	41560.
MISCELLANEOUS OFF-SITE FACILITIES	123850.	133940.	138730.
SUBTOTAL PLANT INVESTMENT	515690.	559610.	580910.
CONTINGENCY AT 10.%	51570.	55960.	58090.
TOTAL PLANT INVESTMENT	567260.	615570.	639000.
WORKING CAPITAL	83970.	89720.	94390.
TOTAL CAPITAL REQUIREMENT	651230.	705290.	733390.

RETURN FROM PRODUCTS		UNITS/CD	THOUS$/YEAR	UNITS/CD	THOUS$/YEAR	UNITS/CD	THOUS$/YEAR
REGULAR GASOLINE,	$ 24.464/B (1)	63332.	565518.	71353.	637137.	72987.	651727.
PREMIUM GASOLINE,	$ 24.464/B (1)	42121.	376117.	47377.	423046.	48468.	432788.
PROPANE LPG,	$ 15.576/B (1)	6658.	37853.	6597.	37506.	6586.	37440.
HYDROTREATED DISTILLATE,	$ 22.853/B (2)	-	-	-	-	10040.	83747.
NO. 2 FUEL OIL,	$ 23.844/B (1)	44464.	386976.	44029.	383186.	43654.	379920.
SULFUR,	$ 30.000/LONG TON	492.	5385.	494.	5413.	497.	5443.
AMMONIA,	$120.00/SHORT TON	17	745.	28.	1226.	40.	1752.
HYDROGEN,	$ 1.750/THOUS SCF	-	-	5806.	3709.	1058.	676.
GAS FEED TO H2 PLANT,	$ 16.841/B FOE (1)	1114.	6847.	1131.	6952.	1132.	6959.
REFINERY FUEL GAS,	$ 16.841/B FOE (1)	4599.	28270.	5203.	31982.	5424.	33342.
REFINERY FUEL OIL,	$ 16.841/B (1)	5991.	36829.	6427.	39507.	6803.	41819.
TOTAL RETURN FROM PRODUCTS			1444540.		1569664.		1675613.
COST OF CHARGE							
KUWAIT CRUDE,	$19.137/B (3)	150000.	1047770.	150000.	1047770.	150000.	1047770.
ISOBUTANE,	$21.152/B (1)	9396.	72545.	9430.	72805.	9437.	72855.
NORMAL BUTANE,	$17.612/B (1)	3105.	19962.	5309.	34130.	5761.	37032.
SRC-II NAPHTHA,C5-350F		-	-	11961.	84199. (4)	11961.	83827. (4)
SRC-II DISTILLATE,350-400F,$19.860/B (5)		-	-	-	-	12353.	89559.
TOTAL COST OF CHARGE			1140277.		1238904.		1331043.
MANUFACTURING EXPENSE							
INVESTMENT-BASED ITEMS (6)			65472		70994.		73657.
LABOR-BASED ITEMS, $10.78/HR (7)		888.	8481.	984.	9399.	1008.	9627.
UTILITIES							
REFINERY FUEL GAS	$16.841/B FOE	4599.	28270.	5203.	31982.	5424.	33342.
REFINERY FUEL OIL	$16.841/B	5991.	36829.	6427.	39507.	6803.	41819.
POWER	$ 0.0303/KWH	947644.	10480.	998250.	11040.	1038020.	11480.
FRESH WATER, RIVER	$ 0.00/THOUS GAL	7063.	0.	7507.	0.	7876.	0.
SUBTOTAL, UTILITIES			75579.		82529.		86641.
GAS FEED TO H2 PLANT	$16.841/B FOE	1114.	6847.	1131.	6952.	1132.	6959
TRANSPORTATION OF REFORMATE		-	-	10921.	1203.	13168.	1452.
CHEMICALS			1315.		1342.		1353.
CATALYSTS			16187.		17147.		18066.
ROYALTY, CRUDE DESALTING,	$0.0025/B	150000.	137.	150000.	137.	150000.	137.
TOTAL MANUFACTURING EXPENSE			174018.		189703.		197892.
TOTAL EXPENSE			1314295.		1428607.		1528935.
COST OF CAPITAL, BEFORE TAX, 20%			130245		141057		146678
CALCULATED VALUE OF SRC-II NAPHTHA, $/B ($/MILL'N BTU)			-		19.286(3.54)		19.201(3:52)

1) FORECAST PRICES IN 1990 DEFLATED TO 1979.
2) PRICE OF HYDROTREATED DISTILLATE CALCULATED AT $22.853/B($4.00/MILL'N BTU) INCLUDING ALLOCATED COSTS ON HYDROTREATING UNIT.
3) CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX IN BASE CASE-CASE 1-A.
4) CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX.
5) CALCULATED AT $3.50/MILL'N BTU.
6) INCLUDES DEPRECIATION, MAINTENANCE, INVESTMENT BURDEN, INSURANCE AND TAXES.
7) INCLUDES DIRECT OPERATING WAGES, SUPERVISION, AND OVERHEAD.

At the operating conditions employed to upgrade naphtha to reformer feedstock quality, the hydrotreated distillate product (350-450°F) is not sufficiently upgraded to produce specification jet fuel product. In addition to the relatively high nitrogen content, 0.55%, the aromatics content is in excess of the maximum of 25% considered to be acceptable for jet fuel. A second-stage jet fuel hydrotreating unit would be required. However, the moderately hydrotreated SRC-II distillate could be blended with high-quality petroleum-derived jet fuels to extend production of transportation fuels.

Hydrogen is produced from the reforming unit at a rate of 1560 SCF/B or 22.9 x 10^6 SCF/CD in a stream of 96% purity. This production of hydrogen is almost totally consumed in hydrotreating of naphtha and middle distillate with a small excess, 1.1 x 10^6 SCF/CD, being available for hydrogen consuming units in the SRC-II plant.

4.0 KUWAIT CRUDE REFINING SCHEMES

Schematic flow diagrams for a Base Case refinery and a refinery in which SRC-II reformate is blended into the gasoline products are shown in Figures 2 and 3, respectively. The Base Case refinery scheme, Figure 2, is the same as that presented in Figure 6 in Part 1. The SRC-II reformate blending scheme in Figure 3 is similar to that in Figure 7 of Part 1, except that SRC-II reformate is available at about a 20% higher rate, 13,168 B/CD versus 10,921 B/CD. Kuwait crude charge rate, 150,000 B/CD, and charge rates to major downstream units - residual HDS, FCC, and alkylation units - are the same in each scheme.

In the scheme charging SRC-II reformate (Figure 3), the severity required in the naphtha reforming unit to enable production of unleaded gasoline products meeting minimum octane number specifications forecast for 1990 is somewhat lower than in the Base Case refinery, 96.0 versus 96.8 RON clear debutanized reformate. This results from the relatively high octane rating of SRC-II reformate - 97 RON clear. The gasoline pool contains 60% regular and 40% premium grades meeting octane number specifications forecast by Gulf for 1990 as follows:

FIGURE 2

UPGRADING OF SRC - II NAPHTHA TO GASOLINE

CASE IA - BASE CASE - KUWAIT CRUDE REFINING SCHEME

TYPE V HDS OF ATMOSPHERIC BOTTOMS PLUS FCC OF HDS BOTTOMS

SCHEMATIC FLOW DIAGRAM

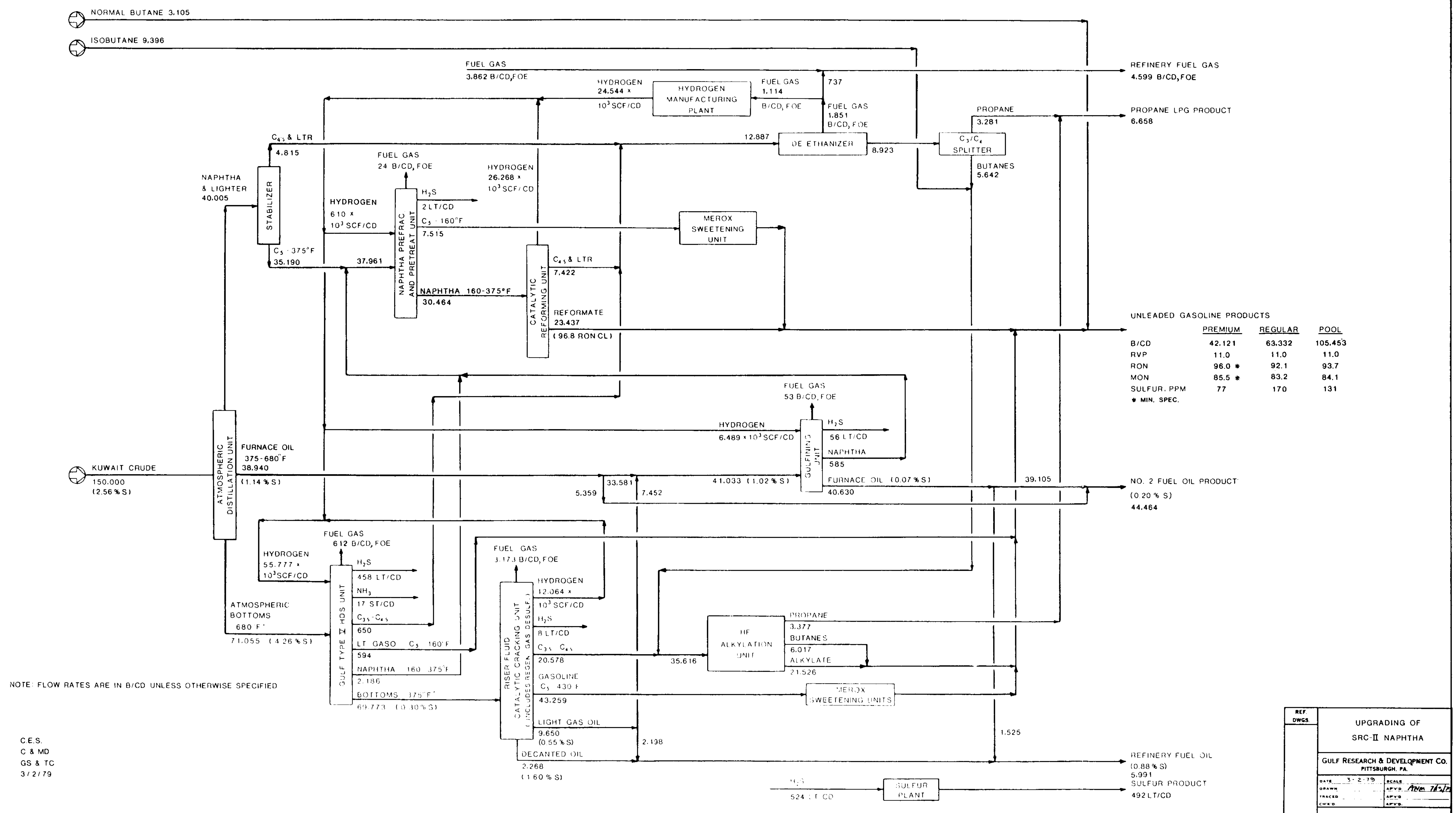


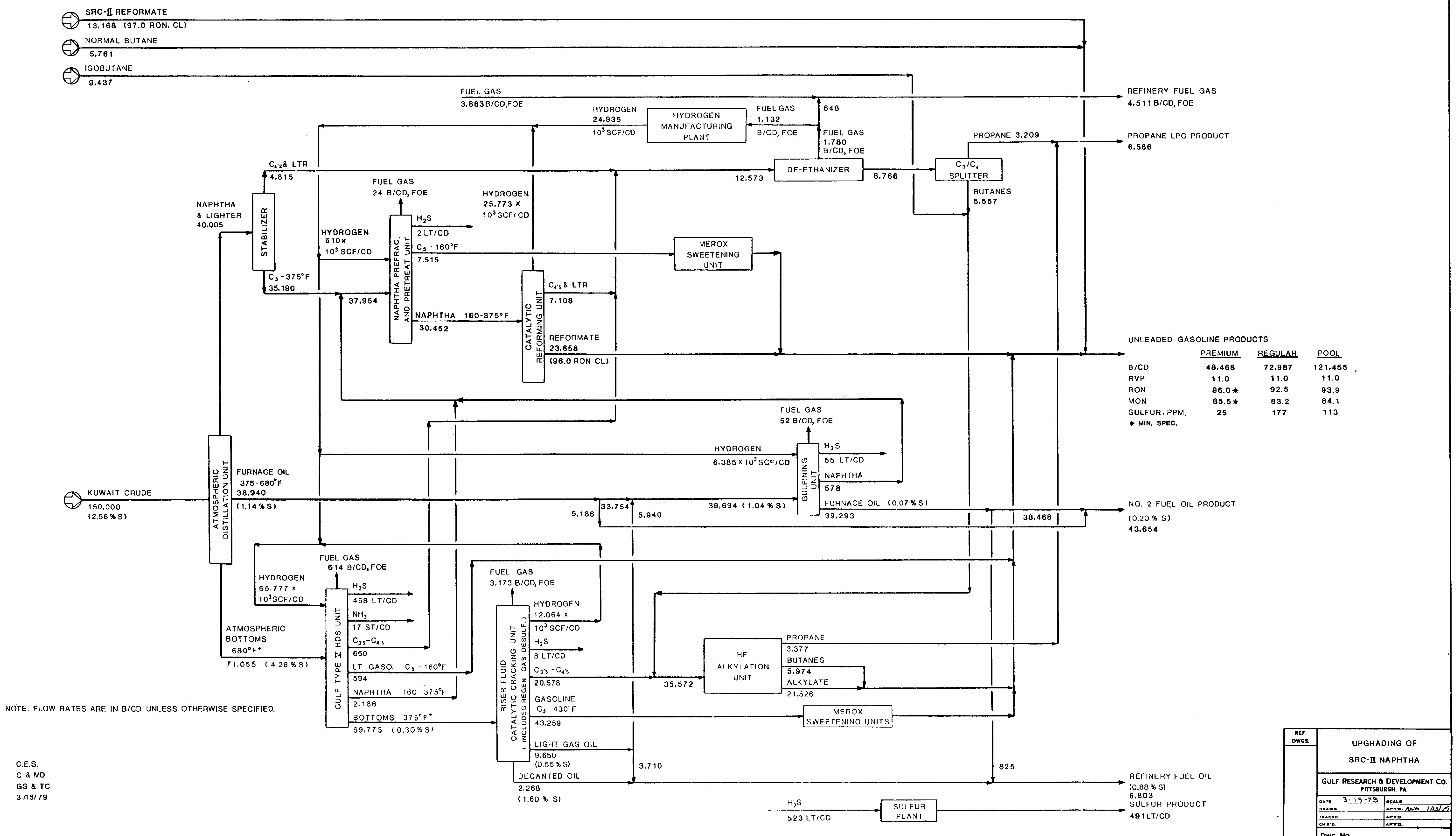
FIGURE 3

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

KUWAIT CRUDE REFINING SCHEME

CASE IIIB - SRC-II NAPHTHA FOR MAXIMUM GASOLINE

SCHEMATIC FLOW DIAGRAM



REF. DWGS.

UPGRADING OF SRC-II NAPHTHA

GULF RESEARCH & DEVELOPMENT CO.
PITTSBURGH, PA.

DATE 3-15-79 SCALE
DRAWN APVD.
TRACED APVD.
CHK'D. APVD.
DWG. NO.

Gasoline Grade	Regular	Premium
Research Octane Number (RON)	92.0	96.0
Motor Octane Number (MON)	83.0	85.5

Total gasoline production in the scheme with SRC reformate is about 15% greater than in the Base Case refinery: 121,455 B/CD versus 105,453 B/CD. Higher yield of reformate from operating the refinery reforming unit at lower severity, increase in butanes blending resulting from low vapor pressure SRC-II reformate, as well as the addition of SRC-II reformate itself, contribute to the increase in gasoline production.

5.0 ECONOMICS

Economics for upgrading raw SRC-II liquid were developed on a current time frame basis for comparison with the economics of other SRC-II projects. Costs of upgrading are as follows:

Raw SRC-II Distillate Charge	C_5-350°F	C_5-450°F
Raw SRC-II Distillate Charge, B/CD	11,961	24,314
	Investment, $\$10^6$ (1978)	
Process Unit	28.1	4.2
Catalysts and Royalties	3.6	5.8
Off-Site Facilities	16.2	24.4
Total Plant Investment	47.9	71.4
Working Capital	5.0	9.6
Total Capital Required	52.9	81.0
	$/B SRC-II Distillate Charge (1979)	
Additional Cost of Upgrading		
Total Manufacturing Expense	3.28	2.52
Cost of Capital	2.43	1.81
Total Mfg. Expense plus Cost of Capital	5.71	4.34

Costs are included in the hydrotreating units for a stripper to recover ammonia and hydrogen sulfide from sour water from the low-pressure separator and for incremental capacity in the existing ammonia plant to recover additional ammonia. Also, costs are included in the hydrotreating unit charging wide-boiling range distillate for a product fractionator to separate naphtha and middle distillate. The economy-of-scale from the larger upgrading plant charging the wide boiling range distillate results in a reduction in manufacturing expense plus cost of capital of $1.37/B of charge.

The economics for the following three schemes charging 150,000 B/CD of Kuwait crude are presented in detail in Table I:

Case I-A - Base Case Refinery - same as in Part 1.

Case III-A - Refinery with addition of reformate from SRC-II naphtha upgrading only for increase in gasoline production - same as in Part 1.

Case III-B - Refinery with addition of reformate from integrated SRC-II naphtha and middle distillate upgrading for increase in gasoline production.

Prices for petroleum products are those forecast for 1990 when the SRC-II process is expected to be commercialized, but deflated by an inflation rate of 6%/yr to 1979. The price of crude oil was calculated to provide 20% return before tax on total capital in the Base Case refinery, Case I-A.

The calculated value of SRC-II naphtha in the integrated naphtha and middle distillate upgrading scheme (Case III-B) is slightly lower, $19.20/B ($3.52/million Btu), than the value of $19.29/B ($3.54/million Btu) calculated in the scheme evaluated in Part 1 (Case III-A) based on upgrading of naphtha only. This results from the relatively high value placed on SRC-II middle distillate, and as a result, on the value of synthetic naphtha produced from middle distillate. However, this value is still greater than that calculated for crude petroleum, $19.14/B ($3.33/million Btu). The economy-of-scale advantage from the larger hydrotreating unit enables the production of about 20% additional high octane number gasoline blending component (15,558 B/CD including butane for 11 psi RVP vs 12,903 B/CD) plus 10,040 B/CD of hydrotreated middle distillate with only a slight loss in profitability.

The value of hydrotreated middle distillate (350-450°F) is calculated to be $22.85/B ($4.00/million Btu) by adding allocated hydrotreating costs including 20% cost of capital to a value of $19.86/B ($3.50/million Btu) which has been developed for raw SRC-II distillate. Economics for upgrading of middle distillate are presented in Table II. If raw middle distillate were valued at only 17.03/B ($3.00/million Btu), the value of raw SRC-II naphtha would increase to $19.81/B ($3.63/million Btu), which is significantly greater than obtained when charging naphtha only to hydrotreating, $19.29/B ($3.54/million Btu). Thus, if the refiner is charged less for raw SRC-II middle distillate, he could afford to pay more for raw SRC-II naphtha.

6.0 CONCLUSIONS

Upgrading of SRC-II naphtha to gasoline has been evaluated in a scheme in which a wide boiling range SRC-II product, including naphtha and middle distillate, is charged to a large hydrotreating unit to achieve economy-of-scale and to take advantage of conversion of middle distillate to naphtha. Hydrotreating costs are allocated between the naphtha and middle distillate fractions of the feed. The value of raw SRC-II naphtha in the integrated hydrotreating scheme is only slightly lower than in the scheme including hydrotreating of naphtha only, $19.20/B ($3.52/million Btu) versus $19.29/B ($3.54/million Btu), and is still greater than that calculated for crude petroleum. A 20% increase in high octane number gasoline blending component and production of about 10,000 B/CD of hydrotreated middle distillate can be obtained with only a slight decline in profitability.

Hydrotreated light distillate (350-450°F) produced along with upgraded naphtha is calculated to have a value of $22.85/B ($4.00/million Btu) based on the value for raw SRC-II distillate and allocated costs for hydrotreating. This distillate is not sufficiently upgraded to jet fuel quality, but it could be blended with high quality petroleum-derived jet fuel to extend production of transportation fuels.

TABLE II

UPGRADING OF SRC-II NAPHTHA TO GASOLINE

HYDROTREATING OF SRC-II MIDDLE DISTILLATE

ECONOMIC EVALUATION
1979 START-UP, THIRD QUARTER 1978 INVESTMENT, NRC=714

CAPACITY, SRC-II MIDDLE DISTILLATE CHARGE,350-450F,B/CD	12353
HYDROTREATED MIDDLE DISTILLATE, C5-450F,B/CD	12547

	CAPACITY, BU/SD	INVESTMENT
ESTIMATED INVESTMENT, THOUS $		
ALLOCATED HYDROTREATING UNIT	13725	11435
CATALYST AND ROYALTY		1625
ALLOCATED UTILITY FACILITIES		570
STORAGE TANKS		1300
MISC. OFF-SITE FACILITIES		4390
SUBTOTAL PLANT INVESTMENT		19320
CONTINGENCY AT 10. %		1930
TOTAL PLANT INVESTMENT		21250
WORKING CAPITAL		5200
TOTAL CAPITAL REQUIREMENT		26450

RETURN FROM PRODUCTS		UNITS/CD	THOUS$/YEAR
HYDROTREATED DISTILLATE	$ 22.853/B (1)	12547	104659
FUEL GAS,	$ 16.841/B FOE	160	984
SULFUR,	$ 30.000/LT	3	33
AMMONIA,	$120.000/ST	12	525
TOTAL RETURN FROM PRODUCTS			106201
COST OF CHARGE			
SRC-II MIDDLE DISTILLATE,	$19.863/B (2)	12353	89559
MANUFACTURING EXPENSE			
FUEL,	$16.841/B FOE	285	1752
POWER,	$ 0.0303/KWH	35585	394
CHEMICALS,			7
CATALYSTS,			901
HYDROGEN,	$ 1.750/THOUS SCF	8647	5523
LABOR BASED ITEMS,	$26.167/HR	36.6	349
INVESTMENT-BASED ITEMS			2426
TOTAL MANUFACTURING EXPENSE			11352
TOTAL EXPENSE			100911
COST OF CAPITAL, BEFORE TAX, 20%			5290
CALCULATED VALUE OF HYDROTREATED SRC-II MIDDLE DISTILLATE, $/B ($/MILL'N BTU)			22.853(4.00)

1 CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX.
2 CALCULATED AT $3.50/MILL'N BTU.

GULF SCIENCE AND TECHNOLOGY COMPANY
PITTSBURGH, PENNSYLVANIA

TRANSPORTATION FUELS FROM SRC–II PRODUCTS

C. E. SINNETT
D. L. STAHLFELD

CHEMICALS AND MINERALS DIVISION
REPORT NO. 633RK108
JULY, 1979

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1.0 SUMMARY

A processing scheme and economics have been developed for upgrading liquid product from a projected conceptual commercial SRC-II plant to high yields of transportation fuels--gasoline, jet fuel, and diesel fuel. The C_5-400°F naphtha fraction and most of the 400°F+ distillate fraction would be hydrotreated at severe conditions in separate, single-stage units over a commercial catalyst and using available technology to produce liquid products containing less than 0.1 and 0.5 ppm of nitrogen, respectively. High conversion to naphtha and light distillate would accompany hydrotreating of the SRC-II 400°F+ distillate fraction.

The composited 180-350°F hydrotreated naphtha from each feedstock would be reformed in a conventional unit to produce a high yield, 91%, of high octane number, 98.0 Research octane number (RON) clear, gasoline blending stock. The hydrotreated light distillate (350-550°F) would contain about 93% naphthenes and less than 5% aromatics, and would meet the smoke point and stringent thermal stability specifications for jet fuel. The quality of jet fuel that could be produced from the upgrading of SRC-II product would be superior to that of typical jet fuels produced from petroleum with respect to a lower freezing point and to a higher volumetric heating value due to the high naphthenes content. Hydrotreated heavy distillate (550°F+) would be blended with about the same quantity of hydrotreated light distillate (350-550°F) to diesel fuel product.

Projected rates of transportation fuels from and plant investment and manufacturing expense for a total liquid upgrading plant for a commercial conceptual SRC-II complex charging 30,000 T/CD of West Virginia Panhandle Powhatan coal are compared with those reported previously for upgrading C_5-400°F naphtha only in the following table. These economics are designed to present relative comparisons and for expediency, use a simple 20% before tax return for treatment of capital costs. The values of both crude oil and SRC-II feedstocks are calculated values in order to be on the same basis. The values derive from revenue requirements based on 1990 forecasted prices,

deflated to 1979, to which are added manufacturing costs. Thus, the absolute values are not reflective of any specific real price expectations, although they are reasonable figures in the context of this preliminary screening study.

	Naphtha Upgrading	Naphtha and Distillate Upgrading
Hydrotreating Unit Charge, B/CD		
Naphtha, C_5-400°F	24,000	24,000
Distillate, 400°F - 550°F	-	{31,000
38% of 550°F+	-	
Total	24,000	55,000
Transportation Fuels, B/CD		
Gasoline, Incl. SRC-II Butanes	22,000	32,000
Jet Fuel	3,600	24,000
Diesel Fuel	-	4,000
Total	25,600	60,000
Total Plant Investment, $\$10^6$ (1978)	74	207
	$/B Transportation Fuels (1979)	
Additional Cost of Upgrading		
Total Manufacturing Expense	2.60	4.54
Total Mfg. Expense plus Cost of Capital	4.40	6.86

The higher projected cost for transportation fuels from the total liquid upgrading plant results directly from the much higher capital, manufacturing expense, and hydrogen consumption for hydrotreating 400°F+ distillate compared with hydrotreating C_5-400°F naphtha.

Reformate and hydrotreated light gasoline, (C_5-180°F), would be blended into gasoline products from a hypothetical existing refinery charging 150 000 B/CD of Kuwait crude to increase gasoline production. As the result of the high octane number (98.0 RON, Clear) of SRC-II reformate, the severity of the refinery reforming unit is reduced from 96.8 to 93.2 RON, Clear. The yield of reformate could be increased significantly with reduction in reforming severity level, from 76.9 vol % on charge in the Base Case to 80.3 vol % in the refinery adding reformate from upgrading of SRC-II total liquid. As a result of this increase in yield and the increase in butanes requirement when blending low-vapor pressure SRC-II reformate, as well as the additional SRC-II gasoline, refinery gasoline production is increased as follows:

	Base Case	Addition of SRC-II Gasoline: Naphtha Upgrading	Addition of SRC-II Gasoline: Total Liquid Upgrading
Gasoline Pool, B/CD	105,000	128,000	141,000
Increase in Gasoline Production, Vol %	0	22	34

The value of raw SRC-II naphtha calculated to provide 20% return on capital before tax on the total liquid upgrading plant and expansion of refinery units is compared with that reported previously for a naphtha-only upgrading plant and with the similarly calculated value of Kuwait crude at the same return on capital in each case as follows:

Value of Refinery Feedstock	Naphtha Upgrading		Naphtha and Distillate Upgrading	
	$/B	$/million Btu	$/B	$/million Btu
Kuwait Crude Oil	19.14	3.33	19.14	3.33
Raw SRC-II Naphtha, C_5-350°F	20.43	3.78	16.52	3.06

The calculated value of raw SRC-II naphtha in the total liquid upgrading plant is less than crude value and less than that when upgrading naphtha only because of the higher costs for distillate hydrotreating. These costs cannot be recovered with the prices forecast in 1990 for gasoline and distillate products, which are, however, only $24.50/B for gasoline and $24.25/B for jet fuel. (These projections were made before the recent abrupt increases in the price of crude petroleum.) The comparative economics for the total liquids upgrading scheme would be substantially improved based on current outlooks for premium petroleum transportation products. Values for petroleum in this study are based on current costs with the same cost of capital as used in the SRC-II liquid upgrading plant.

With the higher costs expected in finding and drilling for petroleum in the future, compared with modest expected increases in constant dollar costs for the SRC-II process after successful demonstration and initial commercialization, the value of raw SRC-II feedstocks as charge to the transportation fuels upgrading plant could become equal to or exceed that for crude petroleum. Because of the recent emphasis by Saudi Arabia on production of heavier crudes and the rising price of premium transportation fuels, it appears probable that an economic situation is developing that will favor increasing conversion of future coal liquids, such as SRC-II fuels, into a spectrum of transportation fuels.

2.0 INTRODUCTION

In Part 7 of this volume, processing schemes and economics were developed for upgrading the naphtha fraction of the liquid product from a projected conceptual commercial SRC-II plant, which in turn is predicated on a successful SRC-II demonstration plant. The upgrading scheme would yield a high octane number gasoline blending component. From a conceptual commercial SRC-II plant charging 30,000 T/CD of West Virginia Panhandle Powhatan coal, 21,961 B/CD of high octane number (96 Research octane number clear) blending component could be produced along with a relatively small quantity, 3583 B/CD, of a light distillate (350-400°F). The light distillate product could qualify as a jet fuel component.

When blending the high octane number component into gasoline products from a large hypothetical existing refinery charging 150,000 B/CD of Kuwait crude, the calculated value of raw SRC-II naphtha would be greater than that for Kuwait crude at the same return on capital.

In the present volume, a processing scheme and economics are presented for approaching maximization of the production of transportation fuels -- gasoline, jet fuel, and diesel fuel -- from total SRC-II liquid. West Virginia Powhatan coal is charged to the SRC-II plant at a rate of 30,000 T/CD as in the previous evaluation. A recent pilot plant study[1] on catalytic hydrotreating of SRC-II naphtha and distillates by Chevron Research Company, funded by the Department of Energy, has been employed. Pilot plant runs of up to about 3000 hours duration were made, after which the catalyst was still active. The runs were terminated voluntarily. This study is based on use of Chevron commercial catalysts and pertains only to the extent that a commercial plant is feasible using presently existing technology. Yields and properties were taken from the Chevron pilot plant data.

The following text describes the projected processing in the upgrading of SRC-II liquids. The SRC-II C_5-400°F naphtha fraction would be charged to a hydrotreating unit operating at severe conditions to produce reformer feedstock as developed and presented in the previous part of Volume 1 for upgrading naphtha only. The limiting constraint would be nitrogen content, which would be reduced to less than 0.1 ppm. In addition, most of the SRC-II 400°F+ distillate would be charged to a second hydrotreating unit operating at more severe conditions to produce a liquid product containing less than 0.5 ppm nitrogen. Along with denitrogenation extensive conversion to naphtha and light distillate products would be achieved. The severity requirement for this unit would be fixed by producing a jet fuel product passing the stringent thermal stability test.

The composited 180-350°F naphtha fraction from each hydrotreating unit would be reformed in a conventional unit to produce a high yield, 91%, of high octane number, 98.0 RON clear, gasoline blending stock. Yields and properties of products from reforming were based on GR&DC pilot plant data. Hydrotreated middle distillate (350-550°F) would have an aromatics content of less than 5% and would meet the smoke point and the stringent JFTOT thermal stability specifications for jet fuel. A relatively small production of hydrotreated heavy distillate (550°F+) would be blended with approximately the same quantity of hydrotreated light distillate (350-550°F) to produce diesel fuel product.

Production rates of transportation fuels from this scheme would be as follows:

	B/CD
Gasoline, Including SRC-II Butanes	32,451
Jet Fuel	23,776
Diesel Fuel	4,100
Total	60,327

The upgraded gasoline, including reformate and hydrotreated C_5-180°F light gasoline, would be transported by pipeline to a large, hypothetical existing refinery charging 150,000 B/CD of Kuwait crude, where it would be blended into gasoline products.

The values of raw SRC-II naphtha were determined on the basis of 1990 forecasted prices of petroleum products, and deflated by 6%/yr to 1979. A cost of capital of 20% was provided for the upgrading plant and expansion of various refinery facilities. The price forecasts are not up-to-date and do not reflect the upsurge in petroleum prices or the dislocations in petroleum refining economics, but instead are based on traditional refining costs and margins added to the previous outlook for crude oil prices.

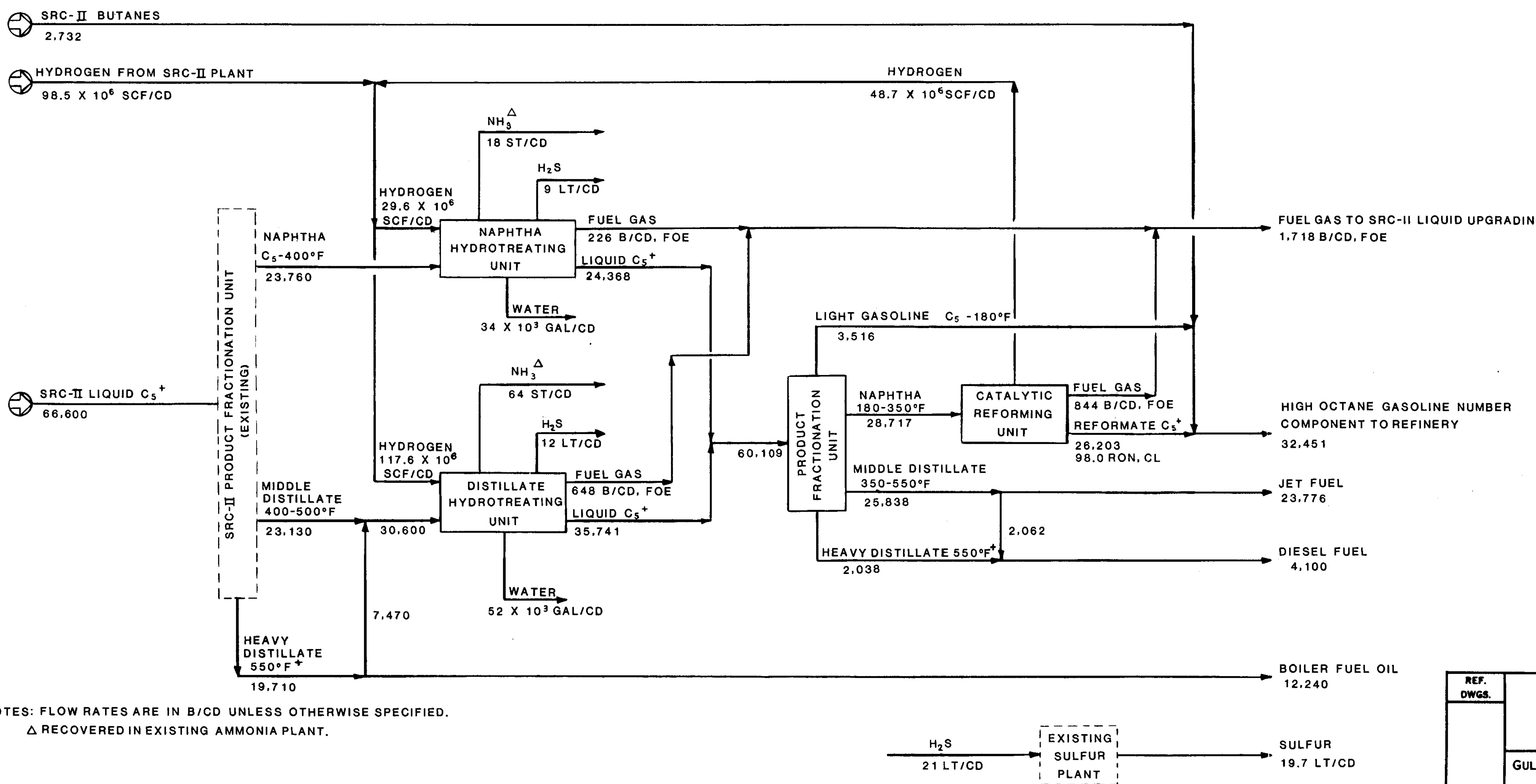
3.0 SRC-II PRODUCT UPGRADING SCHEME

A schematic flow diagram of the upgrading plant to maximize production of transportation fuels is shown in Figure 1. SRC-II naphtha and distillate rates shown are from a conceptual commercial plant charging 30,000 T/CD of West Virginia Powhatan coal using the design basis prepared by Gulf Minerals Resources Company.

Raw SRC-II naphtha (C_5-400°F) would be charged at a rate of 23,260 B/CD to a hydrotreating unit identical to that in the evaluation presented in Part 7 of Volume 1 for upgrading naphtha only, as shown in Figure 2. The unit would be operated at about 800 psig and 0.95 h^{-1} volumetric space velocity to produce reformer feedstock quality naphtha with a limiting constraint of less than 0.2 ppm of nitrogen. Hydrogen would be consumed at a rate of about 1200 SCF/B. Projected operating conditions and product yields are presented in Table I.

FIGURE 1

UPGRADING OF SRC-II LIQUID PRODUCT
TO MAXIMUM TRANSPORTATION FUELS
NAPHTHA FROM WEST VIRGINIA POWHATAN COAL
SCHEMATIC FLOW DIAGRAM



NOTES: FLOW RATES ARE IN B/CD UNLESS OTHERWISE SPECIFIED.
△ RECOVERED IN EXISTING AMMONIA PLANT.

C.E. S.
GS & TC
C & MD
7/12/79

REF. DWGS.	UPGRADING OF SRC-II LIQUID PRODUCT	
	GULF RESEARCH & DEVELOPMENT CO. PITTSBURGH, PA.	
	DATE 7-12-79	SCALE
	DRAWN	APV'D. ANM 7/13/79
	TRACED	APV'D.
	CH'K'D.	APV'D.
	DWG. NO.	

FIGURE 2

UPGRADING OF SRC-II NAPHTHA TO GASOLINE
NAPHTHA FROM WEST VIRGINIA POWHATAN COAL
SCHEMATIC FLOW DIAGRAM

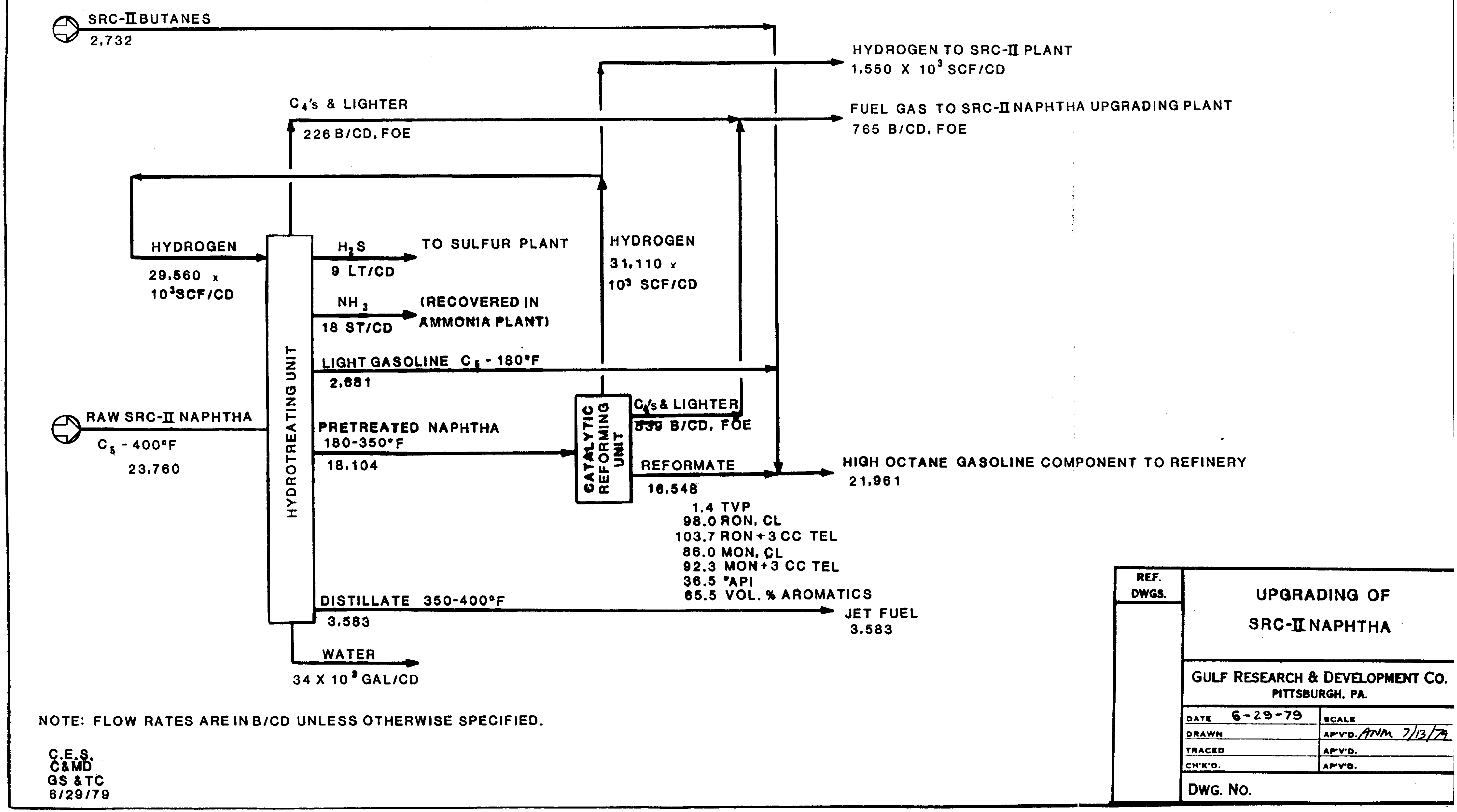


TABLE I

UPGRADING OF SRC-II LIQUID TO MAXIMUM TRANSPORTATION FUELS

Hydrotreating and Reforming Units
Operating Conditions and Product Yields and Inspections

Hydrotreating Units

Charge Stock	SRC-II Naphtha, C_5-400 °F	SRC-II Distillate, 400°F Plus
Operating Conditions		
Total Pressure, psig	793	2495
Hydrogen Partial Pressure, psia	645	2306
Space Velocity, v/h/v	0.95	0.52
Recycle Gas Rate, SCF/B	3450	17590
Average Temperature, °F	699	746
Hydrogen Consumption, SCF/B		
Chemical	1157	3733
Solution Losses	53	111
Catalyst	Chevron ICR 113 Ni-Mo on Silica Alumina	Chevron ICR 106 Ni-W on Silica Alumina
Catalyst Life (Estimated), Months	9	9

Product Yields and Inspections	Vol %	°API	Wt %	ppm S	ppm N	ppm O	Vol %	°API	Wt %	ppm S	ppm N	ppm O
Charge	100.0	31.0	100.00	2600	4200	35100	100.0	11.0	100.00	2500	9900	36300
Hydrogen, Chemical			2.02						5.72			
Total			102.02						105.72			
Products												
NH_3			0.51						1.20			
H_2S			0.27						0.26			
H_2O			3.94						4.08			
C_1			0.03						0.20			
C_2			0.12						0.36			
C_3			0.20						0.47			
C_4's			0.12						0.21			
Gasoline, C_5-180°F	11.3	63.0	9.42				2.7	65.9	1.97			
Naphtha, 180-350°F	76.2	41.1	71.75	11*	<0.10	90	34.7	47.6	27.58			
Light Distillate, 350-400°F	15.1	25.0	15.66				-	-	-	57*	0.43	50
Middle Distillate, 350-550°F	-	-	-				72.7	32.1	63.36			
Heavy Distillate, 550°F Plus	-	-	-				6.7	26.0	6.03			
Total	102.6		102.02				116.8		105.72			

----

* Believed to be high due to incomplete stripping.

SRC-II middle distillate (400-550°F) plus 38% of the SRC-II heavy distillate (550°F+), totaling 30,600 B/CD, would be charged to a second hydrotreating unit operating at about 2500 psig and 0.52 h^{-1} volumetric space velocity to reduce the nitrogen content to less than 0.5 ppm. Under these conditions, the stringent JFTOT thermal stability test for jet fuel at the high temperature of 280°C could be met. The TBP distillation of the feedstock to this hydrotreating unit approximates that for the 400°F+ feedstock in the Chevron pilot plant run,[1] which feedstock contains a relatively high ratio (about 6/1) of middle to heavy distillates. The ratio of 350-550°F middle distillate to 550°F+ heavy distillate in the Gulf SRC-II design basis[2] is about 1.6/1. Therefore, a substantial quantity, 62%, of 550°F+ heavy distillate is rejected from hydrotreating unit feedstock. In actual practice, the separation would probably be made by fractionation to minimize high boiling point feed to hydrotreatment.

Hydrogen would be consumed at a rate of about 3850 SCF/B. Projected operating conditions and product yields are also shown in Table I. Concomitant with high denitrogenation would be extensive conversion to light distillate products as shown by a 37.4 vol % yield of C_5-350°F naphtha and a 72.7 vol % yield of light distillate (350-550°F). Although Chevron refers to this unit as a hydrotreating unit, the term hydrocracking would be more appropriate since high conversion of heavy oil to naphtha and light distillate is obtained over a hydrocracking catalyst and at conventional hydrocracking conditions.

Water would be injected into the reactor effluent cooling train in each hydrotreating unit to scrub out ammonium bisulfide formed upon cooling the reaction products. Sour water from the low-pressure separator would be stripped to recover ammonia and hydrogen sulfide by-products. Bleed gas from the high-pressure separator and gas from product stripping would be charged to the SRC-II ammonia plant for recovery of the remainder of the ammonia. Total ammonia production from both units is 82 ST/CD. Hydrogen sulfide that would be recovered in the sour water stripper would be charged to the sulfur plant in the conceptual commercial SRC-II plant for recovery of 19.7 LT/CD of sulfur.

The 180-350°F hydrotreated naphtha fraction that could be produced at a rate of 28,717 B/CD would be reformed in a conventional, semi-regenerative unit to produce a high yield, 91 vol %, of debutanized reformate having 98.0 Research octane number, clear. Projected operating conditions and product yields are based on GR&DC pilot plant data, as used in the evaluation for upgrading naphtha only. Reformate, hydrotreated light distillate (C_5-180°F) and SRC-II butanes would comprise the high octane number, 96 RON clear, gasoline blending stock which would be transported to the petroleum refinery. The reformate plus butanes and the hydrotreated light gasoline would be transported separately by pipeline in blocked operation in order to provide flexibility in blending into two grades of gasoline.

The light hydrotreated distillate (350-550°F) would have a high naphthenes content, about 93%, would contain less than 5% aromatics, and would meet the smoke point specification and the stringent JFTOT thermal stability test at the high temperature of 280°C for jet fuel. The aromatics content would be significantly below the maximum acceptable limit of 25% for jet fuel. A hydrotreated jet fuel sample containing 27% aromatics that could be obtained at less severe operating conditions could not meet the JFTOT thermal stability test at 280°C. The projected low API gravity of 32.1 for this distillate compared with a minimum specification of 37 for jet fuel reflects the high naphthenes content and is not expected to preclude its use as a jet fuel. Also, as a result of the anticipated high naphthenes content, the freezing point, about -69°F, would be significantly lower than that for typical petroleum-derived jet fuels. The coal-based jet fuel could be blended with a high freezing point petroleum jet fuel to produce specification product.

A relatively small production of hydrotreated heavy distillate, (550°F+) would be blended with an approximately equal quantity of hydrotreated light distillate (350-550°F) to produce diesel fuel product meeting the ASTM distillation specification.

Production rates of transportation fuels in this scheme would be as follows:

	B/CD
Gasoline, Including SRC-II Butanes	32,451
Jet Fuel	23,776
Diesel Fuel	4,100
Total	60,327

In addition, excess heavy SRC-II distillate (550°F+) would be produced at a relatively low rate of 12,240 B/CD for use as a boiler fuel.

Hydrogen would be produced as a by-product from naphtha reforming at a relatively high rate of about 1700 SCF/B in a stream of 96% purity. The production rate, 48.7×10^6 SCF/CD, could supply one-third of the total requirements for the two hydrotreating units. The remainder, 98.5×10^6 SCF/CD, would be supplied from expanded hydrogen manufacturing capacity in the conceptual commercial SRC-II plant. It should be pointed out that in the naphtha-only upgrading scheme, evaluated in Part 7 of Volume 1, hydrogen consumption would be far less than in the maximum upgrading scheme evaluated herein. A small net production of hydrogen was projected in that scheme for use in the conceptual commercial SRC-II plant.

Fuel gas comprising butanes and lighter hydrocarbons and solution hydrogen from the hydrotreating and reforming units would be produced at a rate of 1718 B/CD, FOE. This gas would supply about 55% of the fuel requirement for the upgrading plant, including fuel for steam generation. The remainder of the fuel requirement, 1384 B/CD, FOE, would be supplied from external sources.

4.0 KUWAIT CRUDE REFINING SCHEMES

Schematic flow diagrams for a Base Case refinery (Case IA), a refinery in which reformate and hydrotreated light gasoline (C_5-180°F) from projected SRC-II naphtha upgrading would be blended into the gasoline products

(Case III-A-1), and a refinery in which gasoline from projected maximum upgrading of SRC-II total liquid is blended into the gasoline products (Case V-A), are presented in Figures 3, 4, and 5, respectively. The Base Case refinery and refinery with addition of projected SRC-II gasoline from naphtha upgrading only are identical to those developed and presented previously for evaluation of the potential of SRC-II naphtha upgrading.

Kuwait crude would be charged at a rate of 150,000 B/CD to the refinery designed for maximum conversion of atmospheric bottoms to furnace oil and lighter products. Crude unit atmospheric bottoms (680°F+) would be charged to a Gulf Type V HDS unit at a rate of 71,055 B/CD to produce 69,773 B/CD of 375°F+ bottoms containing 0.30% sulfur for charge to a riser FCC unit. The FCC unit employs a high activity zeolite catalyst at a relatively high make-up rate of 0.54 lb/B to maintain a 1000 ppm nickel equivalent (Ni + 0.2V) metals concentration on equilibrium catalyst to offset deactivation which would otherwise occur from the high metals content, 1.68 ppm nickel equivalent, fresh feed. A conversion of 82.9 vol % to gasoline, 430°F TBP end point, and lighter products is obtained. Regenerator stack gas is scrubbed with a lime solution to reduce SO_x emissions to a maximum specification of 250 ppm by volume on a dry basis. Capacities of the crude, HDS, FCC and alkylation units are the same in each of the three refining schemes evaluated.

Straight-run and HDS naphthas (160-375°F) are reformed over UOP R-18 bimetallic catalyst to produce debutanized reformate having a sufficently high octane number to meet octane numbers forecast by Gulf for unleaded gasoline grades in 1990, as follows:

Grade	Regular	Premium
Vol % Pool	60.0	40.0
Research Octane Number (RON)	92.0	96.0
Motor Octane Number (MON)	83.0	85.5

FIGURE 3

UPGRADING OF SRC-II NAPHTHA TO GASOLINE
CASE IA - BASE CASE - KUWAIT CRUDE REFINING SCHEME
TYPE V HDS OF ATMOSPHERIC BOTTOMS PLUS FCC OF HDS BOTTOMS
SCHEMATIC FLOW DIAGRAM

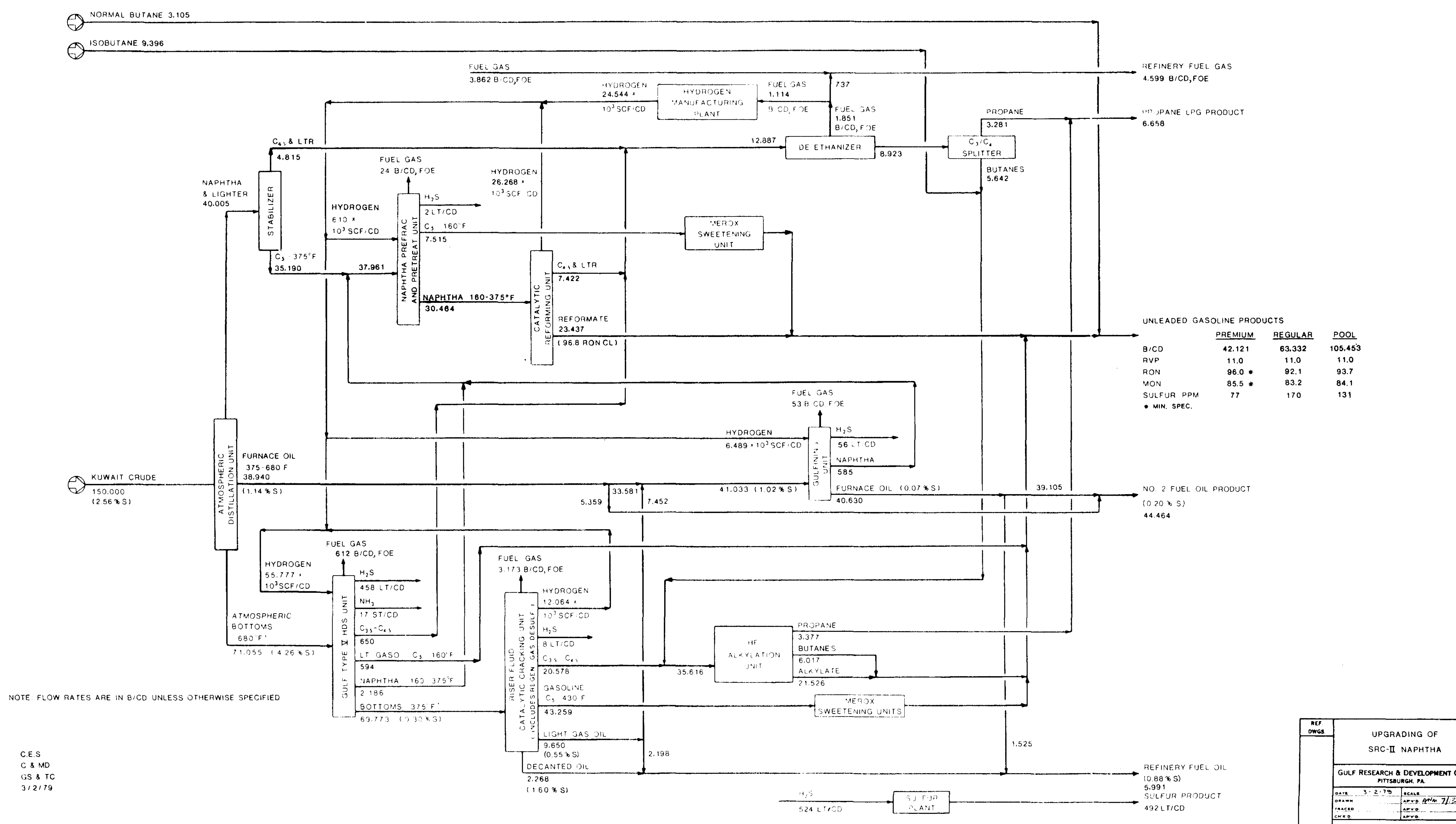
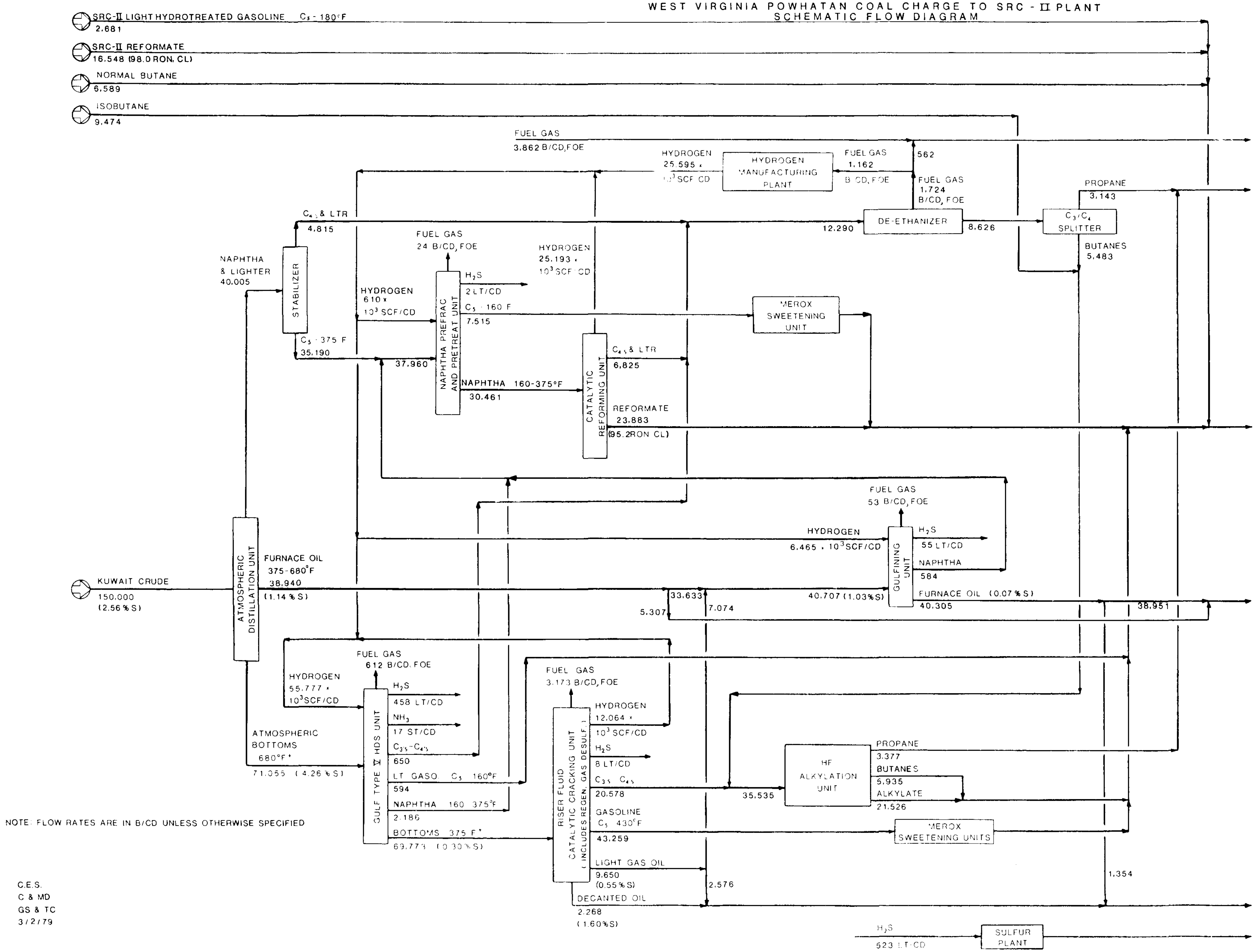


FIGURE 4

UPGRADING OF SRC - II NAPHTHA TO GASOLINE
KUWAIT CRUDE REFINING SCHEME
CASE III - A - 1 - SRC - II NAPHTHA FOR ADDITIONAL GASOLINE
WEST VIRGINIA POWHATAN COAL CHARGE TO SRC - II PLANT
SCHEMATIC FLOW DIAGRAM



	PREMIUM	REGULAR	POOL
B/CD	51.412	77.118	128.530
RVP	11.0	11.0	11.0
RON	96.0 *	92.2	93.7
MON	85.6	83.2	84.2
SULFUR, PPM	19	170	106

* MIN. SPEC.

NOTE: FLOW RATES ARE IN B/CD UNLESS OTHERWISE SPECIFIED

C.E.S.
C & MD
GS & TC
3/2/79

REF. DWGS	UPGRADING OF SRC-II NAPHTHA
	GULF RESEARCH & DEVELOPMENT CO. PITTSBURGH, PA.
DATE 3-2-79	SCALE
DRAWN	APVD.
TRACED	APVD.
CHKD	APVD.
DWG. NO.	

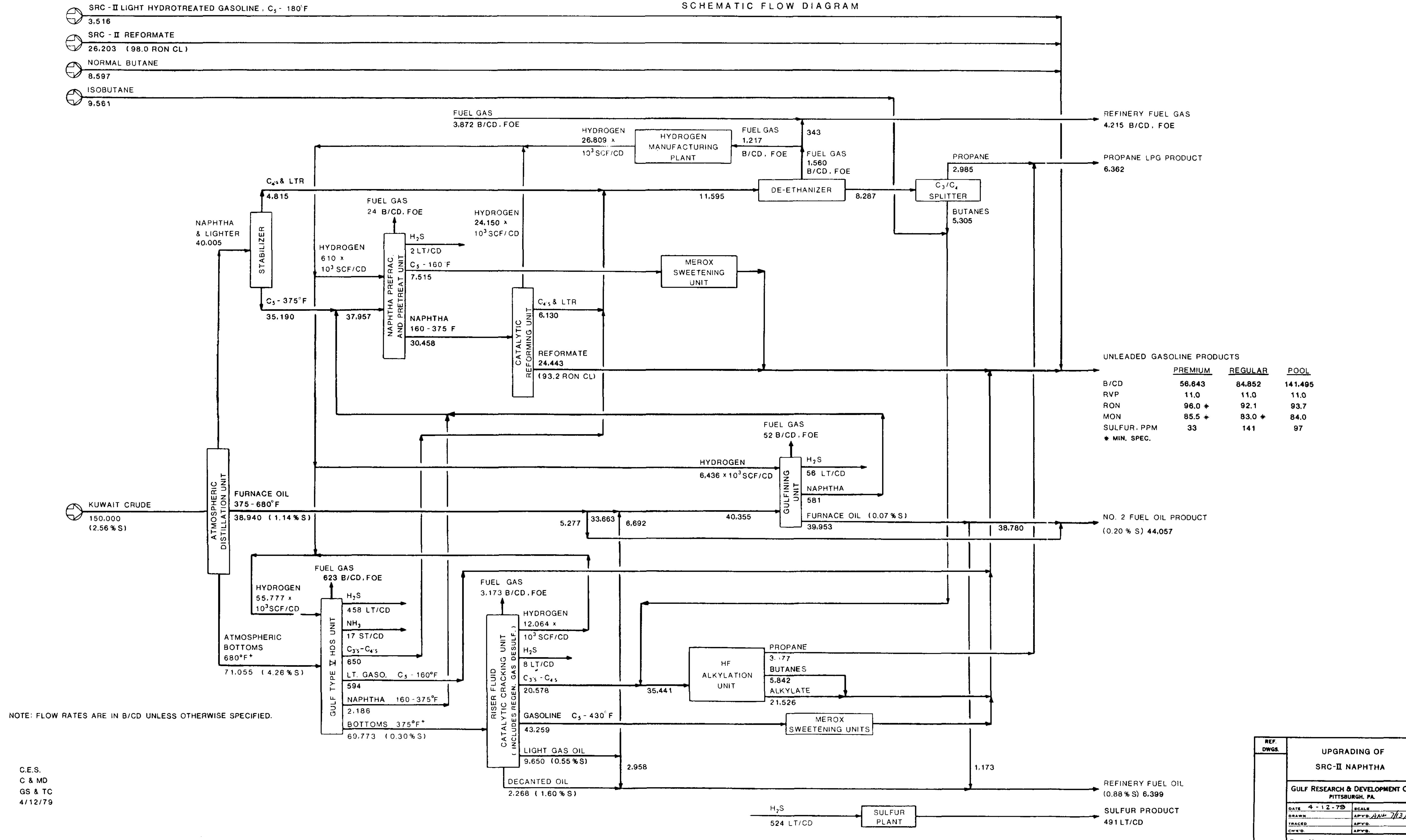
FIGURE 5

UPGRADING OF SRC - II NAPHTHA TO GASOLINE

KUWAIT CRUDE REFINING SCHEME

CASE V-A SRC - II REFORMATE FOR ADDITIONAL GASOLINE

SCHEMATIC FLOW DIAGRAM



The required reformate octane number in the Base Case is 96.8 RON clear. As a result of the high octane number of the projected SRC-II reformate, 98.0 RON, clear, the required severity level of the reforming unit in the refinery is reduced to 95.2 RON clear, when blending 16,548 B/CD of reformate from projected upgrading of SRC-II naphtha only, and to 93.2 RON, clear, when blending 26,203 B/CD of reformate from projected upgrading of SRC-II total liquid. The yield of reformate could be increased significantly with reduction in reforming severity level, from 76.9 vol % on charge in the Base Case to 80.3 vol % in the refinery adding reformate from projected upgrading of SRC-II total liquid. As a result of this projected increase in yield, the increase in butanes requirement when blending low-vapor pressure SRC-II reformate, as well as the potential additional SRC-II gasoline, refinery gasoline production would be increased as follows:

Case	I-A	III-A-1	V-A
		Addition of SRC-II Gasoline	
	Base Case	Naphtha Upgrading	Total Liquid Upgrading
Gasoline Pool, B/CD	105,453	128,530	141,495
Increase in Gasoline Production, Vol. %	0	21.9	34.2

Production of No. 2 fuel oil product from the refinery with addition of projected SRC-II gasoline is only slightly lower, by less than 1.0 percent, than that in the Base Case refinery, 44,464 B/CD.

Hydrogen requirements for the HDS, naphtha pretreating, and furnace oil Gulfining units would be supplied from by-product hydrogen from the naphtha reforming unit, hydrogen recovered cryogenically from FCC off-gas, and hydrogen manufactured by steam-reforming of refinery off-gas. Hydrogen manufacturing capacity would be increased from 24.5 x 10^6 SCF/CD in the Base Case to 26.8 x 10^6 SCF/CD in the refinery with addition of the projected SRC-II gasoline from total liquid upgrading. The increase would result from the decrease in by-product hydrogen from naphtha reforming as severity is decreased.

5.0 ECONOMICS

Economics have been calculated for upgrading SRC-II liquid for projected maximum production of transportation fuels and are compared with those developed in Part 7 of Volume 1 for upgrading SRC-II naphtha only. Economics are on a current time frame basis for comparison with the economics of the proposed SRC-II project. Plant investments, manufacturing expenses, and return on capital for the upgrading plants in each case are summarized as follows:

Case	III-A-1 Upgrading of SRC-II Naphtha Only	V-A Upgrading of SRC-II Total Liquid
	Investment, $\$10^6$ (1978)	
Process Units	40.4	134.5
Catalysts and Royalties	7.7	18.6
Off-Site Facilites	26.0	68.2
Total Plant Investment	74.1	221.3
Working Capital	9.9	34.3
Total Capital Required	84.0	255.6
Additional Cost of Upgrading	$/B Transportation Fuels (1979)	
Total Manufacturing Expense	2.60	4.54
Cost of Capital, Before Tax, 20%	1.80	2.32
Total Mfg. Expense plus Cost of Capital	4.40	6.86

Costs are included in the hydrotreating units for a stripper to recover ammonia and hydrogen sulfide from sour water from the low-pressure separator, for incremental capacity in the SRC-II ammonia plant to recover additional ammonia, and for a product fractionator. Investments include 10% contingency. Inclusion of the heavy distillate hydrotreater, operating at much more severe conditions and with a much higher hydrogen consumption rate, in the projected upgrading plant, would result in a 200% increase in upgrading plant investment and in a 56% increase in upgrading manufacturing expense plus return. These costs do not include costs for SRC-II feedstock or butanes, or credits for by-products, such as fuel gas, sulfur and ammonia.

Economics for the three refining schemes charging Kuwait crude are summarized in Table II. The calculated price of crude oil is \$19.14/B (\$3.33/million Btu) including cost of capital at 20% before tax in the Base Case refinery - Case I-A.

Economics for the product upgrading plant at the SRC-II site are integrated with those for the hypothetical refinery. Prices for the SRC-II 350-400°F and 350°F+ distillate charge to the upgrading plant and the 550°F+ heavy distillate boiler fuel product were based on an estimated raw SRC-II product cost of \$3.50/million Btu. A pipeline tariff for transporting reformate and hydrotreated light gasoline from the hypothetical SRC-II plant site to the refinery at an assumed distance of 300 miles is calculated at \$0.22/B based on literature inflrmation.[1]

As shown in Table II, a value of \$16.52/B (\$3.06/million Btu) is calculated for raw SRC-II naphtha (C_5-350°F) including 20% cost of capital before tax for the hypothetical total liquid upgrading plant plus incremental refining capacity. This value is significantly lower than the value of \$20.43/B (\$3.78/million Btu) calculated for raw SRC-II naphtha in the hypothetical plant to upgrade naphtha only, and is also lower than the value calculated for Kuwait crude of \$19.14/B (\$3.33/million Btu). The lower value for naphtha charge to the hypothetical total liquid upgrading plant results from the significantly higher investment, hydrogen consumption, and manufacturing expense required for hydrotreating heavy (400°F+) SRC-II distillate compared with that for hydrotreating C_5-400°F naphtha only. Price estimates used in this study for transportation fuel products are not adequate to recover the higher costs for the hypothetical maximum fuels upgrading plant.

The quality of jet fuel that could be produced from the upgrading of SRC-II product could be superior to that of typical jet fuels produced from petroleum with respect to a lower freezing point and to a higher volumetric heating value due to the high naphthenes content. If a premium of at least \$6.00/B which has been forecast for high quality distillate fuels in the mid-

TABLE II

UPGRADING OF SRC-II LIQUID TO MAXIMIZE TRANSPORTATION FUELS

KUWAIT CRUDE REFINING SCHEMES

ECONOMIC EVALUATION
1979 START-UP, THIRD QUARTER 1978 INVESTMENT, NRC=714

CASE	I-A BASE CASE	III-A-1 SRC-II NAPHTHA FOR ADDITIONAL GASOLINE	V-A SRC-II LIQUID TO MAX. TRANSP. FUELS
COAL CHARGE TO SRC-II PLANT	-	WEST VIRGINIA POWHATAN	WEST VIRGINIA POWHATAN
SRC-II NAPHTHA, C5-350 F, B/CD	-	16200.	16200.
HYDROTREATER CHARGE, C5-400F, B/CD	-	23760.	23760.
HYDROTREATER CHARGE, 400F+, B/CD	-	-	30600.
PLANT CAPACITY, B/CD OF CRUDE	150000.	150000.	150000.
ESTIMATED INVESTMENT, THOUS $			
REFINERY PROCESS UNITS	325490.	362920.	449240.
CATALYSTS AND ROYALTIES	16540.	22540.	33450.
UTILITY FACILITIES	14130.	16940.	18020.
STORAGE TANKS	35680.	44030.	52090.
MISCELLANEOUS OFF-SITE FACILITIES	123850.	139880.	171390.
SUBTOTAL PLANT INVESTMENT	515690.	586310.	724190.
CONTINGENCY AT 10.%	51570.	58630.	72420.
TOTAL PLANT INVESTMENT	567260.	644940.	796610.
WORKING CAPITAL	83970.	95080.	119840.
TOTAL CAPITAL REQUIREMENT	651230.	740020.	916450.

RETURN FROM PRODUCTS		UNITS/CD	THOUS$/YEAR	UNITS/CD	THOUS$/YEAR	UNITS/CD	THOUS$/YEAR
REGULAR GASOLINE,	$ 24.464/B (1)	63332.	565518.	77118.	688610.	84852.	757674.
PREMIUM GASOLINE,	$ 24.464/B (1)	42121.	376117.	51412.	459074.	56643.	505786.
PROPANE LPG,	$ 15.576/B (1)	6658.	37853.	6520.	37069.	6362.	36170.
JET FUEL,	$ 24.25/B	-	-	3583.	31714.	23776.	210447.
DIESEL FUEL,	$ 23.844/B (1)	-	-	-	-	4100.	35683.
NO. 2 FUEL OIL,	$ 23.844/B (1)	44464.	386976.	44258.	385179.	44057.	383431.
BOILER FUEL OIL,	$ 22.02/B (2)	-	-	-	-	12240.	98377.
SULFUR,	$ 30.000/LONG TON	492.	5385.	500.	5475.	511.	5595.
AMMONIA,	$120.00/SHORT TON	17.	745.	35.	1533.	99.	4336.
HYDROGEN,	$ 1.750/THOUS SCF	-	-	1550.	990.	-	-
GAS FEED TO H2 PLANT,	$ 16.841/B FOE (1)	1114.	6847.	1162.	7143.	1217.	7481.
REFINERY FUEL GAS,	$ 16.841/B FOE (1)	4599.	28270.	5189.	31897.	5933.	36470.
REFINERY FUEL OIL,	$ 16.841/B (1)	5991.	36829.	6198.	38099.	6399.	39334.
TOTAL RETURN FROM PRODUCTS			1444540.		1686783.		2120784.
COST OF CHARGE							
KUWAIT CRUDE,	$19.137/B (3)	150000.	1047770.	150000.	1047770.	150000.	1047770.
ISOBUTANE,	$21.152/B (1)	9396.	72545.	9474.	73141.	9561.	73816.
NORMAL BUTANE,	$17.612/B (1)	3105.	19962.	6589.	42356.	8597.	55265.
SRC-II NAPHTHA, C5-350F		-	-	16200.	120794. (4)	16200.	97695. (4)
SRC-II DISTILLATE, 350-400F,	$19.860/B (2)	-	-	7560.	54802.	-	-
SRC-II DISTILLATE, 350F PLUS,	$20.92/B (2)	-	-		-	50400.	384844.
TOTAL COST OF CHARGE			1140277.		1338863.		1659390.
MANUFACTURING EXPENSE							
INVESTMENT-BASED ITEMS (5)			65472.		74290.		91657.
LABOR-BASED ITEMS, $10.78/HR (6)		888.	8481.	984.	9399.	1056.	10086.
UTILITIES							
REFINERY FUEL GAS,	$16.841/B FOE	4599.	28270.	5189.	31897.	5933.	36470.
REFINERY FUEL OIL,	$16.841/B	5991.	36829.	7072.	43471.	7786.	47860.
POWER,	$ 0.0303/KWH	947644.	10480.	1023420.	11319.	1153370.	12756.
FRESH WATER, RIVER,	$ 0.00/THOUS GAL	7063.	0.	7792.	0.	9364.	0.
SUBTOTAL, UTILITIES			75579.		86687.		97086.
HYDROGEN,	$ 1.150/THOUS SCF	-	-	-	-	98500.	41345.
GAS FEED TO H2 PLANT	$16.841/B FOE	1114.	6847.	1162.	7143.	1217.	7481.
TRANSPORTATION OF REFORMATE		-	-	19229.	1551.	29719.	2397.
CHEMICALS			1315.		1363.		1405.
CATALYSTS			16187.		19347.		26511.
ROYALTY, CRUDE DESALTING,	$0.0025/B	150000.	137.	150000.	137.	150000.	137.
TOTAL MANUFACTURING EXPENSE			174018.		199917.		278105.
TOTAL EXPENSE			1314295.		1538780.		1937495.
COST OF CAPITAL, BEFORE TAX, 20%			130245.		148003.		183289.
CALCULATED VALUE OF SRC-II NAPHTHA, $/B ($/MILL'N BTU)			-		20.429(3.78)		16.522(3.06)

1) FORECAST PRICES IN 1990 DEFLATED TO 1979.
2) CALCULATED AT $3.50/MILL'N BTU.
3) CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX IN BASE CASE-CASE 1-A.
4) CALCULATED TO GIVE 20.0% RETURN ON CAPITAL BEFORE TAX.
5) INCLUDES DEPRECIATION, MAINTENANCE, INVESTMENT BURDEN, INSURANCE AND TAXES.
6) INCLUDES DIRECT OPERATING WAGES, SUPERVISION, AND OVERHEAD.

1990's would be realized, the economics for upgrading SRC-II distillate to maximum transportation fuels would be considerably enhanced. If the premium for high quality distillates could not be obtained, upgrading of only SRC-II naphtha would be the most economical route. The value of SRC-II naphtha by this route is shown to exceed that for petroleum, even on the basis of current estimated costs.

The value calculated for crude oil in this study is based on current costs with the same cost of capital as employed in upgrading of SRC-II liquid product. The trend projected in the future is for increasing production of heavier and higher sulfur crudes with higher refining costs. Production of light, low-sulfur crudes in the free world is currently abut 62% of the total, while known reserves are estimated to be only about 49% of the total. Also, with the expected higher costs for finding oil and for deeper drilling, and with the much higher costs for enhanced recovery techniques, substantial growth in petroleum costs beyond 1990 is forecast. At the same time, based on historical experience in process development, constant dollar costs for the SRC-II process would be expected to increase only modestly after demonstration and possible subsequent initial commercialization as a result of technological improvements. Thus, it can be projected that the value of SRC-II distillate for upgrading to maximum yield of transportation fuels will become equal to or exceed the value of crude petroleum.

6.0 ECONOMIC BASES

Prices for pool gasoline, No. 2 fuel oil, propane, butanes, and refinery fuel (residual fuel oil with 1.0% sulfur guarantee) are those forecast for 1990, deflated by 6% yr to 1979. Price for diesel fuel is assumed to be equal to that for No. 2 fuel oil, while the price for jet fuel is approximately midway between that for pool gasoline and No. 2 fuel oil. Current market prices are used for sulfur and ammonia. The value of excess hydrogen produced from the reforming unit in the plant to upgrade SRC-II naphtha only is calculated to be $\$1.75/10^3$ SCF for manufacture by partial oxidation of heavy oil, as planned for the conceptual commercial SRC-II complex. Hydrogen supplied from the projected SRC-II plant to the plant for upgrading of SRC-II liquid to maximum transportation fuels is priced at the estimated SRC-II product cost, \$3.50/million Btu (\$1.15/10^3 SCF), since excess synthesis gas production at the SRC-II plant would be consumed as plant fuel.

The investment and operating requirements for the hydrotreating unit charging 400°F+ SRC-II distillate at high pressure, 2500 psig, were based on cost data given by the Chevron Research Company in a study(3) for refining of shale oil, funded by the Department of Energy. Costs for the naphtha hydrotreating unit operating at lower pressure, 800 psig, and for other refinery process units, utility units, and oil storage tanks, were based on capacity correlations developed by Gulf. A contingency factor on investment of 10% is included. Costs for the required expansion of refining facilities in the schemes in which SRC-II reformate and hydrotreated light gasoline would be blended into gasoline products were assumed to be equal to the difference between the investment for a new unit at the expanded capacity and that at the Base Case capacity.

Oil storage capacity is provided as follows:

	Days
Raw SRC-II Naphtha	7
Raw SRC-II Distillates	10
SRC-II Reformate and Hydrotreated Light Gasoline, SRC-II Site	20
SRC-II Reformate and Hydrotreated Light Gasoline, Refinery Site	20
Crude Oil	20
Iso and Normal Butanes Charge to Refinery	5
Gasoline and Distillate Products	20
LPG Product	5
Charge to Intermediate Process Units, except HDS Units	7
Charge to HDS Unit	14
Refinery Fuel Oil	10

Investments for miscellaneous off-site facilities, including site preparation, roads, fencing, buildings, communications, etc., were calculated as a percentage of the total investment for process and utility units, and storage tanks, based on actual costs for a number of Gulf refineries. Paid-up royalties are included for all processing units as required, including those licensed by GR&DC.

7.0 CONCLUSIONS

SRC-II liquid product could be converted to high yields of transportation fuels--gasoline, jet fuel, and diesel fuel--by hydrotreating in single-stage catalytic units, at high operating severity, in which the nitrogen content would be reduced to less than 0.5 ppm. Total production of transportation fuels would be 87% of raw C_5+ SRC-II liquid. Process yields and properties were taken from pilot plant runs of long duration. The study is based on use of Chevron commercial catalysts and pertains only to the extent that a commercial plant would be possible using existing technology.

The potential value of SRC-II naphtha in an upgrading scheme to maximize transportation fuels is calculated to be less than its value in a scheme to upgrade naphtha only, and less than that calculated for crude petroleum. Higher costs for converting 400°F+ SRC-II distillate, compared with upgrading naphtha only, cannot be recovered in prices forecast for gasoline and distillate products. However, this conclusion is based on forecasts made prior to the recent upsurge in crude oil prices and subsequent demand for transportation fuels, as well as before the recent emphasis by Saudi Arabia on production of heavier crudes. These factors, in relation to demand, are causing the price of premium transportation fuels to rise. It appears probable, therefore, that an economic situation is developing that will favor increasing conversion of future coal liquids, such as SRC-II fuels, into a spectrum of transportation fuels.

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APPENDIX A

Refining the Products From the SRC Coal Liquefaction Process

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REFINING THE PRODUCTS FROM THE SRC COAL LIQUEFACTION PROCESS

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SRC-I and SRC-II are two versions of the Solvent Refined Coal Liquefaction process under development by the Pittsburg and Midway Coal Mining Company in programs funded by the Department of Energy (DOE). The primary products from these processes have sharply contrasting characteristics. SRC-I product is a high melting solid that is a relatively unattractive feed for conversion to distillate fuels using petroleum hydroprocessing technology. In contrast, the SRC-II product is a distillate that is a more attractive feed for conversion to transportation fuels. Essentially all of the nitrogen, sulfur, and oxygen can be removed in a single catalytic hydrotreating stage to yield a naphtha that is an excellent feed for a catalytic reformer and a middle distillate fraction that is a premium jet fuel meeting smoke point and stability specifications. Only 5-6% of the hydrotreated SRC-II product boils above the jet boiling range. In this paper, several possible refining routes for the production of distillate fuels from SRC-II process product are considered, based on results from several demonstration pilot plant tests recently completed by Chevron Research Company in a program sponsored by the DOE. In all cases, the initial hydrotreating step is the key step in the processing sequence. High hydrogen consumption, however, leads to high refining costs for this feedstock.

INTRODUCTION

Chevron Research Company, under a contract sponsored by the U.S. Department of Energy, is conducting a program to determine the feasibility and estimate the costs of refining a number of synthetic crude feedstocks to distillate fuels such as high octane gasoline and jet fuel using modern petroleum processing technology. Pilot plant tests for the key processing steps are being conducted to the extent needed to make reasonable estimates of commercial plant performance.

The first feedstock studied under this contract was Paraho shale oil. In a series of recent papers (1-4) and a DOE report (5), three basic shale oil processing routes for the production of transportation fuels were studied: hydrotreating followed by hydrocracking, hydrotreating followed by fluid catalytic cracking (FCC), and severe coking followed by hydrotreating. It was concluded that shale oil can be refined to high quality transportation fuels via modern state-of-the-art refining technology and that it can serve as a substitute for crude oil in a refinery equipped with modern hydrotreating facilities. The key to successful shale oil refining is the initial hydrotreating step which removes contaminants (nitrogen, sulfur, oxygen, olefins, and metallic contaminants) and permits the use of conventional refining facilities.

The present paper is a progress report on the results of a similar study to determine the feasibility of converting solvent refined coal to transportation fuels.

The Solvent Refined Coal Process in its two forms is one of the major processes under current study in programs sponsored by the DOE for conversion of coal to either (1) a solid deashed low sulfur product or (2) a low boiling liquid.

Much of the current work on the project is being carried out by the Pittsburg and Midway Coal Mining Company in a 50 T/D pilot plant at Ft. Lewis, Washington. In the original Solvent Refined Coal Process (6), now designated as SRC-I, coal is dissolved under moderate hydrogen pressure in an internally generated heavy aromatic solvent to yield a product from which most of the mineral matter is removed by filtration. Solvent is recovered for reuse by vacuum distillation. The distillation residue is a solid under ambient conditions known as Solvent Refined Coal or SRC-I.

In an alternate version of the process (7,8), designated SRC-II, a portion of the coal solution is recycled as solvent in place of the distillate solvent of the SRC-I process. The filtration step is eliminated; and, typically, the process operates at a higher pressure, higher temperature, and longer residence time than the SRC-I process. Hydrogen consumption and the conversion of dissolved coal is increased, and the primary product is a liquid rather than the solid product of the SRC-I process. Typical yields and conditions for the processes as reported by Anderson (7) are shown in Table I.

The primary intended use for SRC-I is as a low sulfur boiler fuel. Its successful use in this service has been reported in a test conducted by Southern Company Services, Inc., in a Georgia Power Company coal-fired utility boiler (9). SRC-II has been tested successfully as a substitute fuel oil in a New York power generating plant of Consolidated Edison (10). Results were well within Environmental Protection Agency's (EPA) 1978 standards for emissions.

A limited amount of work has been done on the upgrading of SRC-I. Because SRC-II is a newer process, little work has been reported on its upgrading.

Stein et al. (11,12) studied the hydroprocessing of SRC-I to produce turbine fuel and concluded that such a process is feasible. High severity hydrotreating was required to produce acceptable product.

de Rosset, Tan, and Gatsis (13) concluded SRC-I, when mixed with hydrotreated product, could be upgraded by high severity hydroprocessing. Heptane insolubles were reduced from 89% to 5-12% in a 450-hour run. Hydrogen consumption was 3800 SCF/bbl. The rate of catalyst deactivation was high.

Givens et al. (14,15) studied the hydrotreating of SRC-I in the presence of creosote oil. Creosote oil was chosen as a substitute solvent because of its similarity in ring composition to SRC distillate and because it is derived from coal. Runs up to 180 hours were made in an upflow mode, and distillate yields of up to 46% were obtained. It was concluded that numerous questions remain before a clear-cut evaluation of catalytic SRC-I upgrading can be made.

Currently, work is in progress in which SRC-I product is being upgraded using the LC-fining expanded bed process (16). This combination process has the advantage of avoiding plugging problems that accompany high conversion fixed bed processes.

Some problems have been encountered in upgrading of coal liquids to specification transportation fuels. For example, Kalfadelis and Shaw (17) attempted to make jet fuels from several shale oils and coal-derived liquids by hydroprocessing them over nickel-molybdenum and cobalt-molybdenum catalysts. The products from shale oil met the specifications for jet fuel, but those from coal-derived liquids did not.

Gallagher, Humes, and Siemssen (18) studied the FCC as a process to upgrade syncrudes from coal. The raw syncrudes gave poor selectivity. However, with sufficient hydrogenation pretreatment, these stocks were shown to give motor fuels of quality comparable to conventional fuels.

Research at the University of Delaware (19-21) indicates that removal of nitrogen from multiring compounds such as found in coal-derived liquids is quite difficult. Hydrodenitrogenation takes place with much lower selectivity than does hydrodesulfurization, for example. The cyclic requires extensive hydrogenation before the C-N bonds are weak enough to break. There is no known catalyst that can selectively break C-N bonds in the manner desulfurization catalyst breaks C-S bonds and leaving most of the aromatic rings unhydrogenated. Similarly, saturation of any aromatic ring attached to an oxygen is required before breaking the C-O bond in oxygen-containing compounds (22).

Oyxgen can be even more difficult to remove than nitrogen. Furimsky (23) showed that for catalytic removal from heavy gas oils, the relative rate of desulfurization is fastest, followed by hydrodenitrogenation and, finally, hydrodeoxygenation. He related the rates of reaction to differences in bond strengths. With coal-derived liquids, Stein et al. report that the degree of heteroatom removal is sulfur>oxygen>nitrogen (12). Detailed studies of representative pure heterocompounds that appear in petroleum residua, shale oils, and coal oil show that comparison of reactivities of heteroatoms is complex and that it is difficult to generalize which heteroatom is most difficult to remove. The rate of reaction depends on the specific compound type. For example, Rollman (22) ranked hydrogenation reactivities by chemical type as follows: sulfides>> p-alkylphenols>benzothiophenes>quinoline~o-alkylphenols ~indoles~benzofurans~naphthalenes>dibenzofurans>>mononuclear aromatics.

FEEDS

The Pittsburg and Midway Coal Mining Company provided the products from the SRC process that were used as feeds in these studies. They were produced in the Ft. Lewis, Washington, pilot plant. Three fractions of SRC-II were provided, and these were reblended to represent the net liquid product production in a ratio recommended by the DOE to constitute the net whole liquid process product from "typical" SRC-II operation as best could be estimated at the time this project was started (April 1978). The blend was a mixture of product prepared from Kentucky No. 9, Kentucky No. 14, and Illinois No. 6 coals.

Table II shows the inspections of SRC-I and SRC-II and compares these materials with creosote oil (obtained in the destructive distillation of coal), shale oil, and a Middle Eastern petroleum crude.

The chloride was not removed from the samples before hydroprocessing. However, it was shown that water washing can remove most of the chloride. It is our understanding that this would typically be done at the upstream (SRC) processing facility in a commercial situation.

PROCESSING OF SRC-I

Figure 1 is a simplified schematic flow diagram of our pilot plant studies for processing SRC-I. Because of the very high pour point of SRC-I, it was necessary to mix it with a solvent in order to be able to pump the SRC in our pilot plant equipment. SRC-I recycle solvent was unavailable; so, therefore, at the suggestion of DOE, 50% creosote oil by weight was used as the solvent. Hydrotreating tests with the 50/50 SRC-I/creosote oil were made until sufficient 350-850°F product was available to serve as a "simulated recycle solvent" in subsequent tests. In downstream processing tests, the 350-850°F product prepared from SRC-I/creosote oil was further hydrotreated and then either (1) hydrocracked or (2) fluid catalytic cracked. The catalyst used in the hydrotreating tests was ICR 106 containing nickel, tungsten, silica, and alumina.

Figure 2 shows the effect of time on stream on catalyst activity. Catalyst temperatures are normalized to 0.5% nitrogen in the product. The catalyst deactivated with time. Purity of the recycle hydrogen decreased as the run progressed due to formation of light gases. At about 550 hours, a recycle gas bleed was started; and some activity was regained due to the hydrogen pressure increase. The run was shut down after over 1100 hours on stream with the SRC-I/creosote oil blend. There was no plugging of the catalyst bed or preheat, but several times during the run the unit completely depressured suddenly when a pressure relief valve gave way as a result of plugging in feed or product lines. Undoubtedly, some of the catalyst deactivation was the result of coking that occurred due to the abrupt losses of pressure.

Table III shows results for the 50/50 SRC-I/creosote oil blend. These yields were obtained after several hundred hours on stream. Appreciable catalyst deactivation had occurred.

Tests were made in which the feed was a 50/50 blend of SRC-I and the 350-850°F product of the previous run ("simulated recycle solvent"). At the same conditions as the previous run, conversion levels were much higher than with the SRC-I/creosote oil feed. For example, Table IV shows results of a yield period taken after about 24 hours onstream. Because of the short time onstream, this sample is not necessarily representative of lined out operation. However, we were unable to obtain satisfactory yield periods after longer periods of time onstream due to plugging problems in the catalyst bed. All of our attempts to run at this severity with the SRC-I/solvent resulted in catalyst bed plugging. Probably, the coke-forming tendencies of the SRC-I and the relatively high metals content both contributed to the plugging problem; this problem is particularly severe if the more-easily-converted feed molecules are cracked, leaving behind the harder-to-convert feed molecules.

Our general conclusion is that at low conversions such as obtained with the SRC-I/creosote oil, the plugging problem can be avoided; but the conversion is lower than desired. At higher conversion with the SRC-I/hydrogenated solvent, plugging is a severe problem that would have to be solved before it could be said that SRC processing in a fixed bed is commercially feasible. Because the conversion of SRC-I to transportation fuels by these routes does not appear promising, we did not attempt to solve this problem and concentrated our efforts on the processing of SRC-II.

PROCESSING OF SRC-II

Pilot plant tests were made to hydrotreat the whole SRC-II process product blends to remove nitrogen, sulfur, oxygen, and metals using fixed catalyst beds.

These tests were designed (1) to evaluate two Chevron catalysts, ICR 106 (containing nickel, tungsten, silica, and alumina) and ICR 113 (containing nickel, molybdenum, silica, and alumina), (2) to select appropriate processing conditions, (3) to determine yields and hydrogen consumptions, (4) to determine product properties, and (5) to determine catalyst deactivation rates.

The first pilot plant test was a 3030-hour run using ICR 106 catalyst. The second was a 2650-hour test using ICR 113 catalyst. Tables V and VI summarize the conditions and results of the test. Both runs were terminated voluntarily, and the catalysts were still active. Each catalyst lost a maximum of about 25°F activity during the fouling rate portion of the test and showed no significant activity change in the portions of the run at higher pressure. The ICR 113 is somewhat less active than ICR 106. However, it is a less expensive catalyst and, therefore, may be the catalyst of choice for cases at the lower severities.

For convenience, the tests at 0.5 LHSV and 2300 psia hydrogen will be referred to as "high severity." Those at 1.5 LHSV and 2300 psia will be referred to as "intermediate severity," and those at either 1.0 or 1.5 LHSV and 1750 psia will be referred to as "moderate severity." Results at other conditions will not be discussed in detail.

At high severity, the nitrogen, sulfur, and oxygen were all removed to the lower limit of our ability to measure with the analytical techniques employed (Dohrmann nitrogen, Dohrmann sulfur, neutron activation for oxygen). The product was water-white in appearance.

Table VII compares yields and whole liquid product properties obtained at both high and intermediate severities with ICR 106. Results are shown both when the catalyst is fairly fresh and after aging. Table VIII compares results with partially aged ICR 106 and ICR 113 caalysts at moderate severity.

Figure 3 shows the relationship between product nitrogen and hydrogen consumption.

PRODUCT PROPERTIES

A. Naphtha

Table IX shows properties of the naphtha product at the high and intermediate severities. The naphtha produced at the highest severity is essentially free of nitrogen, sulfur, and oxygen and can be fed directly to a catalytic reformer. The naphtha produced at the lower severities can be further processed in a conventional naphtha hydrotreater before reforming. Table X shows product properties at moderate severity.

Table XI shows the breakdown by carbon number and group type of the 180-300°F naphtha at high severity. The very high yield of cyclics shows that this material would produce a high octane gasoline upon reforming. Furthermore, the exceptionally high yields of C_6-C_9 aromatics after reforming would make this naphtha an unusually attractive feed for petrochemicals production if a use other than as a transportation fuel is desired.

B. Jet Fuel

Table XII shows the properties of the 300-550°F product obtained at the two higher severities.

Table XIII summarizes some of the current commercial (Jet A) specifications (24) compared to the properties of the 300-550°F fraction from the severely hydrotreated SRC-II. One production inspection of particular interest is the JFTOT thermal stability rating of the jet boiling range product (ASTM D 1660). The product from the high severity ICR 106 tests easily passes this test at 280°C with a No. 1 rating, even when the smoke point is 20 mm, the minimum range of the current commercial specification. Current jet specification requires a No. 1 rating at 260°C, a temperature considerably less severe than the 280°C used here. Therefore, smoke point appears to be the limiting specification, and jet fuel produced from SRC-II is expected to pass the thermal stability specification at any condition that the smoke point is acceptable.

One property outside current limits is the API gravity. This specification was originally set because of limitations of available fuel flow controllers. Today's controllers are no longer limited in this way, but there has been no incentive to lower the specification because until now there was no available product below 37°API which met the other specifications.

The gravimetric heat of combustion is only slightly above the current specification. This quantity determines the range of newly designed airplanes. The volumetric heat of combustion, which determines the range of an existing plane with fixed fuel tanks, is quite high because the fuel is unusually dense.

A recent study by Franck et al. at the Institut Francais du Petrole (25) indicates that naphthenic hydrocarbons with two or three rings (molecular weight between

120 and 200) give the best compromise among the different properties of jet fuels. Of the types of compounds studied, these are the only ones to show, simultaneously, the following properties: (1) high calorific value by volume, (2) satisfactory calorific value by weight, (3) acceptable thermal stability, (4) very low freezing point, (5) acceptable low temperature viscosity, (6) low volatility and high flash point, and (7) acceptable flame characteristics. Our results show that these observations apply, in particular, to the jet fuel produced by hydroprocessing SRC-II.

C. Heating Oil

The bottoms prepared at moderate severity could be marketed as No. 2 heating oil or as a heating oil blending component if a product other than transportation fuel is desired. Our tests show that the 300°F+ fraction of partially hydrotreated SRC-II will pass the ASTM specifications for No. 2 heating oil except for gravity. The 1978 ASTM minimum specification for gravity is 30; if the 300°F+ fraction of SRC-II is hydrotreated to 30 gravity, it becomes acceptable jet fuel and, therefore, would not be marketed as heating oil; heating oil from SRC-II is only acceptable for applications in which the low gravity is not a problem. Because of the high heating value on a volumetric basis, the low gravity may be advantageous in some applications.

PROCESSING OF HYDROTREATED SRC-II

If the SRC-II is hydrotreated at an intermediate or moderate severity, further processing will be necessary to upgrade the product to specification transportation fuels. A number of pilot plant tests have been made to determine the conditions and to demonstrate the feasibility of these processing steps. We are analyzing results of these tests, which will be reported in future papers. Figure 4 outlines the pilot plant programs.

At both the moderate and intermediate severities, it is necessary to further hydrotreat the naphtha before it can be fed to a bimetallic reforming catalyst. The jet boiling range product requires further hydrotreating before it can be marketed as jet fuel.

Tests were made to demonstrate the feasibility of fluid catalytic cracking and hydrocracking of the heavier product fractions produced at moderate severity. In both cases, the processes are being examined for applications in which gasoline is to be the major refinery product.

Another processing option for SRC-II is initially to hydrotreat the naphtha separately from the heavier distillates rather than processing the whole SRC-II product in a single hydrotreater. Experiments were made to compare the processing schemes. The results showed no cost incentive for separate processing; although for a large refinery, some benefit might be achieved in added processing flexibility.

REFINING PLANS AND PRELIMINARY COST ESTIMATES

Studies are in progress to evaluate alternate refining schemes for SRC-II oil. Based on the pilot plant work, the schemes can be classified into three general types:

1. High severity hydrotreating to produce reformer feed and jet fuel. The only further downstream processing would be reforming to produce motor gasoline.

2. Intermediate severity hydrotreating with downstream naphtha and middle distillate hydrotreating. Motor gasoline and jet fuel would again be the main products. If maximum gasoline is desired rather than jet fuel, the higher boiling hydrotreated product could also be further converted.

3. Moderate severity hydrotreating to produce naphtha and heating oil or feed to a downstream conversion process such as FCC or hydrocracking.

A. High Severity Hydrotreating

A simplified flow diagram of a refinery centered around high severity hydrotreating is shown in Figure 5. Along with the main processing plants, auxiliary facilities are included to meet air and water pollution requirements while producing sulfur and ammonia as valuable by-products. All refinery fuel and hydrogen are derived from SRC-II or its products; depending on how they are supplied, the overall refinery yields and consumptions would range as follows:

	LV % of Whole SRC-II Oil Feed
Net Yields	
Motor Gasoline	23-27
Kerosene Jet Fuel	59-62
Internal Consumptions	
Refinery Fuel	(11)
Hydrogen Plant Feed	(16)

Motor gasoline would meet current specifications, including 93 clear research octane minimum and ASTM 50% point 250°F maximum. Kerosene jet fuel quality would be as shown for the high severity case on Table XII discussed previously.

Total investment for this refinery sized to produce 50,000 barrels per day of gasoline plus jet fuel is between 500 and 600 million dollars. This estimate includes the processing plant shown in Figure 5, all required offplot facilities required for a grassroots refinery, and reasonable estimates for working capital and contingencies based on a mid-continent location as of Third Quarter, 1979. Feed to the refinery is SRC-II oil, assumed to be received at the refinery via pipeline. Overall processing cost would be between $10 and $12 per barrel of product at the 50,000 barrels per day product rate, assuming a 15% discounted cash flow rate of return on investment. The processing costs are exclusive of all SRC-II oil feed and exclude transportation costs for feed and products.

B. Intermediate Severity Hydrotreating

One possible alternate refining scheme is centered around the intermediate severity hydrotreating operation discussed earlier. The expectation here is that the cost savings in hydrogen manufacture and initial hydrotreating would more than compensate for the added costs of naphtha and middle-distillate hydrotreating. The refining arrangement is generally similar to the severe hydrotreating case, except for the added downstream processing. The product slate is also similar. Cost studies on this alternate are in progress as well as studies of further downstream conversion.

C. Moderate Severity Hydrotreating

Figure 6 shows some on the refining alternates based on moderate severity hydrotreating. A product suitable for home or industrial heating oil can be produced directly by this type of operation, as discussed previously. The range of refinery yield structure for this alternate is:

	LV % of Whole SRC-II Feed
Net Yields	
Motor Gasoline	28-30
No. 2 Oil	58-60
Internal Consumptions	
Refinery Fuel	(7)
Hydrogen Plant Feed	(8)

Total investment for this refinery on the same basis as the severe hydrotreating case would be roughly 400 million dollars with an overall processing cost of between $7 and $9 per product barrel, again at 50,000 barrels per day product rate. Again, the processing costs are exclusive of all SRC-II oil feedstock costs and exclude transportation costs for feed and products.

D. Moderate Severity Hydrotreating Plus Cracking

Another alternate, also shown in Figure 6, is the use of a conversion process on the material boiling above reformer feed to maximize the yield of gasoline. This feed should be nearly ideal as a hydrocracking stock in that its boiling range could allow relatively low severity processing and its hydrocarbon composition (about 90% double rings) is expected to result in a high ultimate gasoline yield. Evaluations of this type of processing are in progress, as are studies of FCC as the conversion step.

Studies of alternates for each of the refining plans are continuing. The influence of refinery capacity and the possibilities for integration with the SRC-II process are being evaluated as well as further variations on the refining arrangement.

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TABLE I

COMPARISON SRC I AND SRC II RUNS
(DATA OF ANDERSON, REFERENCE 7)

	SRC I	SRC II
Reaction Conditions		
Pressure, psig	1425	2000
Dissolver Temperature, °C	450-455	465
Nominal Residence Time, Hr	0.4	1.0
Hydrogen Feed, Wt % of Slurry	2.1	4.9
Feed Slurry Composition		
Coal, %	39.0	35.0
Coal Solution, %	-	65.0
Distillate Solvent, %	61.0	
Yields, Wt % Based on Coal		
C_1-C_4	2.9	12.2
CO, CO_2, H_2S	3.0	3.5
Water	6.4	11.4
Light Oil, C_5-249°C	16.4	19.5
Heavy Distillate, <249°C	-	24.8
Total Oil	16.4	44.3
SRC (Organic Residue)	56.3	15.9
IOM (Insoluble Organic Matter)	5.3	4.5
Ash	11.3	11.5
H_2 Reacted (Gas Balance)	1.6	4.0

TABLE II

PROPERTIES OF SRC PROCESS PRODUCT, CREOSOTE OIL, SHALE OIL, AND ARABIAN LIGHT CRUDE

Description	SRC I	Whole SRC II	Creosote Oil*	Paraho Shale Oil	Arabian Light
Inspections					
Gravity, °API	-14.6	18.6	-4.9	20.2	33.4
Aniline Point, °F	–	<30	77		
Sulfur, Wt %	0.89	0.29	0.64	0.66	1.8
Total Nitrogen, Wt %	2.04	0.85	0.78	2.18	0.09
Basic Nitrogen, Wt %	0.86	0.7	0.45		
Oxygen, Wt %	4.52	3.79	1.11	1.16	
Hydrogen, Wt %	6.1	9.1	6.1	11.3	
Hydrogen/Carbon Atom Ratio	0.8	1.3	0.7	1.6	
Chloride, ppm	50	50	9	<0.2	
Pour Point, °F	>400	Below -80		90	-30
Ramsbottom Carbon, Wt %	58	0.70	0.60	2.5	4.7
Hot Heptane Insolubles, Wt %	96	0.05	0.002	0.2	0.44
Benzene Insolubles, Wt %	60	<0.03	<0.03		
Ash, Wt %	0.22	0.004	<0.003		
Molecular Weight		132	179	326	375
TBP Distillation, °F					
St/5	159/943	56/189	343/406	183/347	112/219
10/30	1017/1161	241/379	446/534	453/610	281/472
50		424	599	736	655
70/90		473/562	631/680	842/963	857/
95/99		642/820	687/726		

*70% Overhead from Allied Chemical's Creosote Oil

TABLE III

HYDROTREATING OF 50/50 SRC/CREOSOTE OIL WITH AGED ICR 106 AT 750°F, 0.2 LHSV

Description	50/50 SRC/ Creosote Oil	Product
Gravity, °API	-7.4	3.6
Nitrogen, ppm	14,600	4300
Sulfur, Wt %	0.90	0.04
Oxygen, Wt %	2.70	0.57
H/C Atom Ratio	0.76	1.13
Hot C_7 Insolubles, %	52.2	8.01
Ramsbottom Carbon, %	29.0	10.5
Benzene Insolubles, %	30.2	2.5
850°F+, LV % of C_5+	45	28
Chemical H_2 Consumption, SCF/Bbl		2300

TABLE IV

HYDROTREATING OF 50/50
SRC/PRODUCT SOLVENT WITH
ICR 106 AT 750°F, 0.2 LHSV*

Description	50/50 SRC/Solvent	Product
Gravity, °API	-2.9	16.1
Nitrogen, ppm	12,900	240
Sulfur, Wt %	0.54	0.004
Oxygen, Wt %	2.53	0.13
H/C Atom Ratio	0.94	1.44
Hot C_7 Insolubles, %	40.8	0.18
Ramsbottom Carbon, %	28.6	0.74
Benzene Insolubles, %	14.0	0.04
850°F+, LV % of C_5+	45	3
Chemical H_2 Consumption, SCF/Bbl		3900

*After only 24 hr on stream with this feed. (See text.)

TABLE V

HYDROTREATING OF SRC II WITH ICR 106 CATALYST
RUN CONDITIONS AND PRODUCT NITROGEN

Time On Stream, Hr	LHSV	Avg. Cat. Temp., °F	Pressure Total, psig	Pressure H_2, psia[1] (Approx.)	Nitrogen Content of Whole Liquid Product, ppm	Catalyst Deactivation Observed
0-350	0.5	750	2500	2300	<0.5	No
350-425	1.0	750	2500	2300	<0.5	No
425-620	1.5	750	2500	2300	20	No
620-2100	1.5	750	2000	1700-1800[2]	100-800	Yes
2100-2300	1.5	750	2500	2300	450	No
2300-2500	1.5	775	2500	2300	20	No
2500-3030	0.5	775	2500	2300	<0.5	No

[1]Recycle gas rate at 2500 psig, 15,000 SCF/bbl; at 2000 psig, 8000 SCF/bbl.

[2]Intermittent periods of partial plugging in reactor exit due to formation of ammonium chloride resulted in lower recycle gas rates and lower hydrogen pressures.

TABLE VI

HYDROTREATING OF SRC II WITH ICR 113 CATALYST
RUN CONDITIONS AND PRODUCT NITROGEN
AVERAGE CATALYST TEMPERATURE 750°F

Time On Stream, Hr	LHSV	Avg. Cat. Temp., °F	Pressure Total, psig	Pressure H_2, psia[1] (Approx.)	Nitrogen Content of Whole Liquid Product, ppm	Catalyst Deactivation Observed
0-490	1.0	750	2500	2300	10	No
490-1660	1.0	750	2000	1700-1800[2]	60-700	Yes
1660-1860	1.0	750	2500	2300	130	No
1860-2115	1.0	775	2500	2300	10	No
2115-2650	0.5	775	2500	2300	0.5	No

[1]Recycle gas rate at 2500 psig, 15,000 SCF/bbl; at 2000 psig, 8000 SCF/bbl.

[2]Intermittent periods of partial plugging in reactor exit lines due to formation of ammonium chloride resulted in lower recycle gas rates and lower hydrogen pressures.

TABLE VII

HYDROTREATING OF SRC II
COMPARISON OF YIELDS AND LIQUID PRODUCT PROPERTIES
AT HIGH AND INTERMEDIATE SEVERITIES
WITH ICR 106 CATALYST

2300 psia H_2

Severity	High (0.5 LHSV)				Intermediate (1.5 LHSV)			
Catalyst Condition	Fresh		Aged		Fresh		Aged	
Avg. Cat. Temp., °F	750		776		750		768	
	Wt %	Vol %	Wt %	Vol %	Wt %	Vol %	Wt %	Vol %
No Loss Prod. Yields								
C_1	0.17		0.22		0.08		0.12	
C_2	0.29		0.37		0.16		0.20	
C_3	0.37		0.53		0.17		0.24	
iC_4	0.03	0.05	0.05	0.08	0.02	0.02	0.02	0.03
nC_4	0.18	0.28	0.21	0.34	0.09	0.14	0.12	0.19
C_5-180°F	4.7	6.1	5.6	7.1	5.1	6.5	5.4	6.9
180-300°F	26.5	32.3	27.1	32.7	25.6	30.8	24.5	29.3
300-550°F	61.4	67.9	60.6	66.6	58.6	62.4	58.6	62.0
550°F-EP	5.8	6.1	4.4	4.7	7.9	8.0	8.4	8.3
Total C_5+	98.4	112.4	97.8	111.0	97.1	107.7	96.9	106.6
H_2 Cons. (Chemical), SCF/Bbl	3100		2900		2000		2000	
Inspections of Whole Liquid Product								
Gravity, °API	39.3		38.4		34.1		32.7	
Aniline Point, °F	116.9		105.4		67.5		54.8	
Sulfur, ppm	5		3		5		21	
Nitrogen, ppm	0.25		0.27		20		52	
Hydrogen, Wt %	13.8		13.6		12.0		11.8	
Oxygen, ppm*	40		40		630		680	
Group Type, LV %								
Paraffins	3.0		3.9		4.1		4.5	
Naphthenes	93.2		81.2		54.7		46.8	
Aromatics	3.9		14.9		41.2		48.7	
TBP Distillation, °F								
St/5	65/181		65/181		59/175		66/179	
10/30	215/276		215/277		209/279		213/286	
50	365		364		370		380	
70/90	410/499		410/498		435/518		442/528	
95/99	546/648		543/657		570/685		581/705	

*Includes dissolved water.

TABLE VIII

HYDROTREATING OF SRC II WITH
AGED CATALYSTS AT MODERATE SEVERITY
750°F--1750 PSIA H_2
8000 SCF/BBL RECYCLE GAS RATE

Catalyst	ICR 106		ICR 113	
LHSV	1.5		1.0	
	Wt %	Vol %	Wt %	Vol %
No Loss Prod. Yield				
C_1	0.10		0.10	
C_2	0.17		0.19	
C_3	0.19		0.19	
iC_4	0.02	0.03	0.02	0.03
nC_4	0.09	0.14	0.08	0.13
C_5-180°F	4.3	5.5	3.8	4.8
180-350°F	32.5	38.4	31.8	37.5
350°F-EP	59.8	60.6	61.1	61.9
Total C_5+	96.6	104.5	96.6	104.2
H_2 Consumption (Chemical), SCF/Bbl	1725		1740	
Whole Liquid Product Properties				
Gravity, °API	30.3		29.9	
Aniline Point, °F	38.3		38.1	
Sulfur, ppm	10		8	
Nitrogen, ppm	560		453	
Hydrogen, Wt %	11.0		11.2	
Oxygen, ppm	4300		2800	
Group Type, LV %				
Paraffins	4.4		3.9	
Naphthenes	38.7		40.1	
Aromatics	56.9		55.9	
TBP Distillation, °F				
St/5	58/174		59/181	
10/30	212/292		216/309	
50	394		401	
70/90	446/539		454/545	
95/99	600/728		608/746	

TABLE IX

HYDROTREATING OF SRC II WITH ICR 106 CATALYST
NAPHTHA PRODUCT PROPERTIES
2300 PSIA H_2

Severity	High		Intermediate	
Condition of Catalyst	Fresh	Aged	Fresh	Aged
Inspections, C_5-180°F Product				
Gravity, °API	60.7	58.6	59.9	59.9
Group Type, LV %				
Paraffins	31.2	26.2	30.0	32.8
Naphthenes	65.4	66.3	62.3	58.0
Aromatics	3.4	7.4	7.7	9.1
Olefins		0.1		0.1
Octane Number				
F-1 Clear	81.1	82.2	81.5	81.1
F-2 Clear	77.0		77.4	
Inspections, 180-300°F Product				
Gravity, °API	51.6	49.9	49.0	48.4
Aniline Point, °F	108.8	96.8	87.4	79.2
Nitrogen, ppm	0.1	0.1	2.1	2
Oxygen, ppm	<20	<20	280	300
Group Type, LV %				
Paraffins	7.1	10.1	10.1	11.3
Naphthenes	90.1	79.5	75.8	69.3
Aromatics	2.8	10.4	14.1	19.4
Octane Number				
F-1 Clear	66.0	68.4	69.2	72.0
F-2 Clear	66.0	66.9	67.3	68.5
TBP Distillation, °F				
St/5	98/177	157/184	159/184	147/183
10/30	180/220	206/233	203/232	202/232
50	246	252	252	251
70/90	269/294	277/299	280/302	279/301
95/99	303/-	306/317	310/323	308/319

TABLE X

PROPERTIES OF PRODUCT FRACTIONS
HYDROTREATING OF SRC II WITH AGED CATALYSTS
AT MODERATE SEVERITY

Catalyst	ICR 106	ICR 113
LHSV	1.5	1.0
Inspections		
C_5-180°F Naphtha		
Gravity, °API	59.3	60.9
Group Type, LV %		
Paraffins	29.1	31.6
Naphthenes	63.3	60.3
Aromatics	7.6	8.1
Octane Number		
F-1 Clear	81.5	81.4
F-2 Clear	77.4	
180-350°F Naphtha		
Gravity, °API	45.7	45.5
Aniline Point, °F	78.8	79.1
Total Nitrogen, ppm	119	126
Oxygen, ppm	3200	1700
Group Type, LV %		
Paraffins	11.3	8.0
Naphthenes	65.5	69.2
Aromatics	23.2	22.8
Octane Number		
F-1 Clear	72.2	69.3
F-2 Clear	67.7	67.1
TBP Distillation, °F		
St/5	157/180	161/182
10/30	212/231	213/230
50	269	269
70/90	297/336	301/337
95/99	346/363	347/362
350°F+ Product		
Gravity, °API	20.6	20.7
Aniline Point, °F	<32	<32
Total Nitrogen, ppm	702	529
Oxygen, ppm	5500	3400
Smoke Point, mm	5	8
Cetane No.	<21	<21
Pour Point, °F	-55	-38
Group Type, LV %		
Paraffins	5.2	4.7
Naphthenes	28.0	26.6
Aromatics	66.8	68.8
TBP Distillation, °F		
St/5	329/363	329/362
10/30	381/417	381/417
50	451	452
70/90	496/585	497/586
95/99	648/771	648/765

TABLE XI

HYDROTREATING OF SRC II
WITH AGED ICR 106 AT HIGH SEVERITY

Composition of
180-300°F Naphtha

Composition, LV %	
Paraffins	
C_6-	0.3
i-C_7	1.7
n-C_7	1.8
i-C_8	2.1
n-C_8	1.8
C_9's	2.1
C_{10}'s	0.2
C_{11}'s	0.1
Total Paraffins	10.1
Naphthenes	
Cyclopentane	0.04
Methylcyclopentane	0.7
Cyclohexane	7.7
C_7-Cyclopentanes	6.3
Methylcyclohexane	26.8
C_8-Cyclopentanes	4.3
C_8-Cyclohexanes	26.0
C_9-Cyclopentanes	2.2
C_9-Cyclohexanes	5.5
Total Naphthenes	79.5
Aromatics	
Benzene	0.4
Toluene	4.6
Ethylbenzene	1.5
Xylenes	3.6
C_9 Aromatics	0.2
Total Aromatics	10.3

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TABLE XII

HYDROTREATING OF SRC II WITH ICR 106 CATALYST
300-550°F PRODUCT PROPERTIES
2300 PSIA HYDROGEN

Severity	High		Intermediate	
Condition of Catalyst	Fresh	Aged	Fresh	Aged
Inspections of 300-550°F Product				
Gravity, °API	34.6	33.3	28.5	27.3
Aniline Point, °F	122.3	110.2	60.3	41.8
Nitrogen, ppm	0.1	0.1	19	40
Oxygen, ppm	<50	<50	970	1400
Smoke Point, mm	22	20	11	11
Freeze Point, °F	<-94	-81	-76	-76
Group Type, LV %				
Paraffins	3.3	3.9	4.0	4.3
Naphthenes	92.6	81.1	47.1	36.7
Aromatics	4.0	15.0	48.8	59.0
TBP Distillation, °F				
St/5	266/304	277/312	264/313	281/318
10/30	316/365	329/373	332/382	334/386
50	404	403	419	421
70/90	451/504	442/497	460/510	459/507
95/99	520/560	516/542	526/550	522/544
Viscosity				
cSt at -40°F	12.79	11.92	13.17	
cSt at 100°F		1.640	1.591	
Existent Gum, mg/100 ml	2	0	High	High
Flash Point, Tag Closed, °F	132	134	142	142
Neutralization No. (Acid), mg/g	0.019	0.05	0.034	0.05
Naphthalenes, %	0.11	0.22	0.94	1.35

TABLE XIII

COMPARISON OF JET FUEL
FROM SRC II WITH
CURRENT COMMERCIAL SPECIFICATION FOR JET A

Property	Jet A Specification (ASTM D 1655-78)	Properties of 300-550°F Jet Fraction from Severely Hydrotreated SRC II Oil
Gravity, °API	37-51	33-35
Hydrogen, Wt %	None as Yet	13.8
Net Heating Value		
Btu/Lb	18,400 Minimum	18,450
Btu/Gal.	-	130,500
Thermal Stability, JFTOT Break Point, °C	260 Minimum	>280
Smoke Point, mm	20 Minimum	20-22
Aromatics, LV %	20 Maximum	5-15
Freezing Point, °F	-40	<-80
Existent Gum, mg/100 ml, Max.	7	2
Corrosion, Copper Strip, Two Hours at 212°F	No. 1	No. 1

FIGURE 1

PILOT PLANT TESTS
PROCESSING OF SRC-1

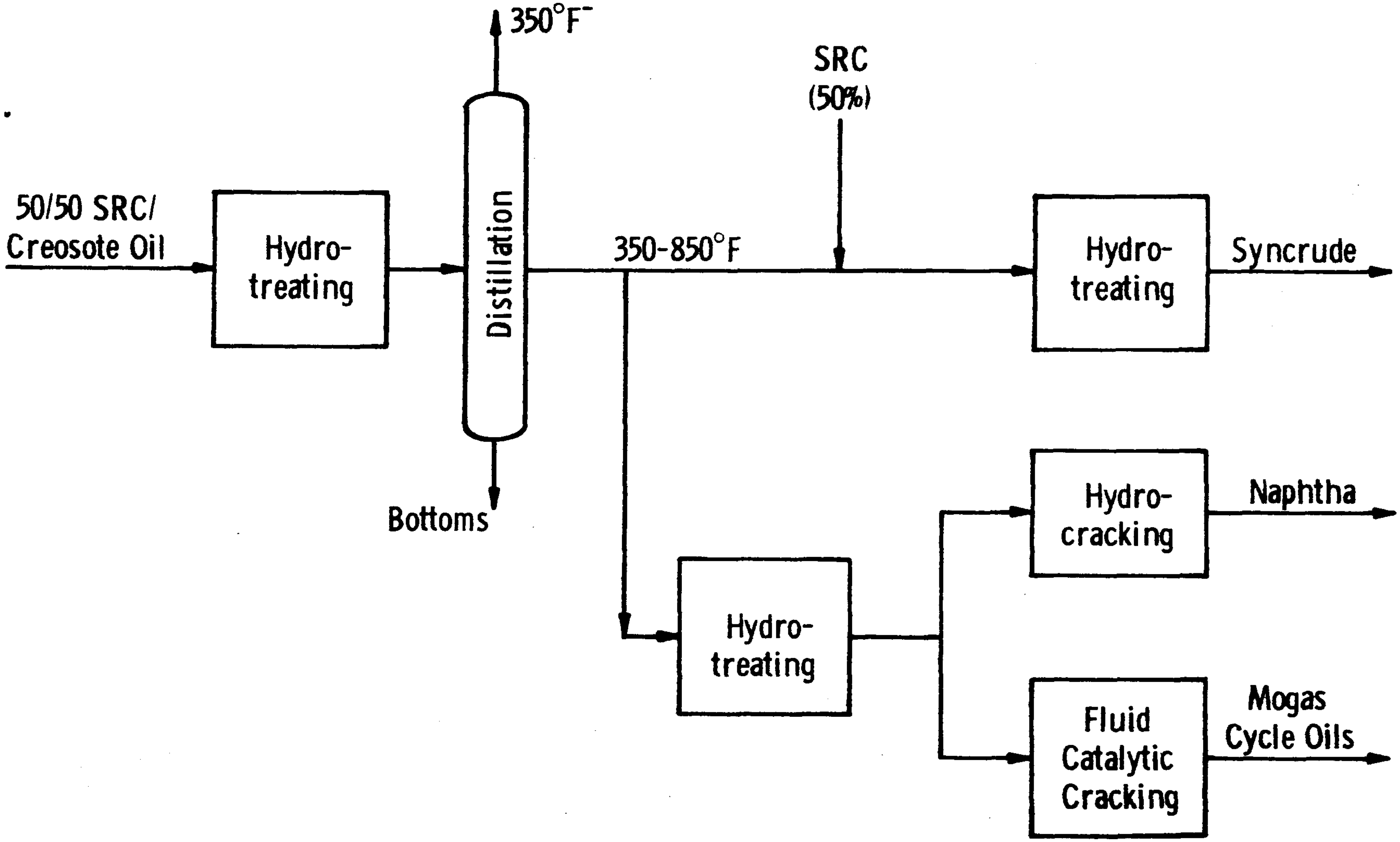

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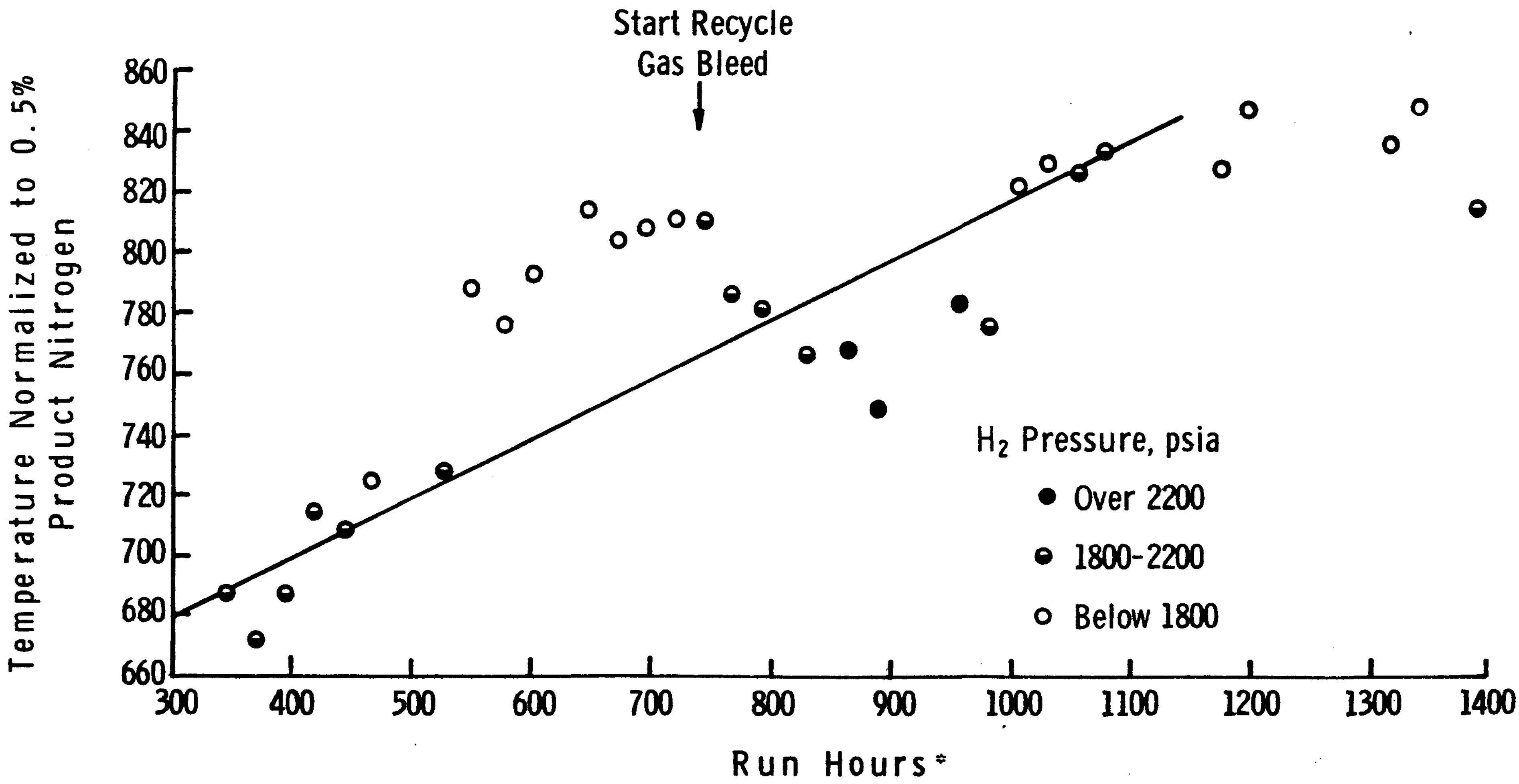
FIGURE 2

HYDROPROCESSING OF 50/50 SRC-I CREOSOTE OIL
BLEND AT 0.2 LHSV WITH ICR 106



*200 Hr Prior Service with Creosote Oil

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FIGURE 3

PRODUCT NITROGEN VERSUS
HYDROGEN CONSUMPTION
HYDROTREATING OF
SRC-II AT 750°F

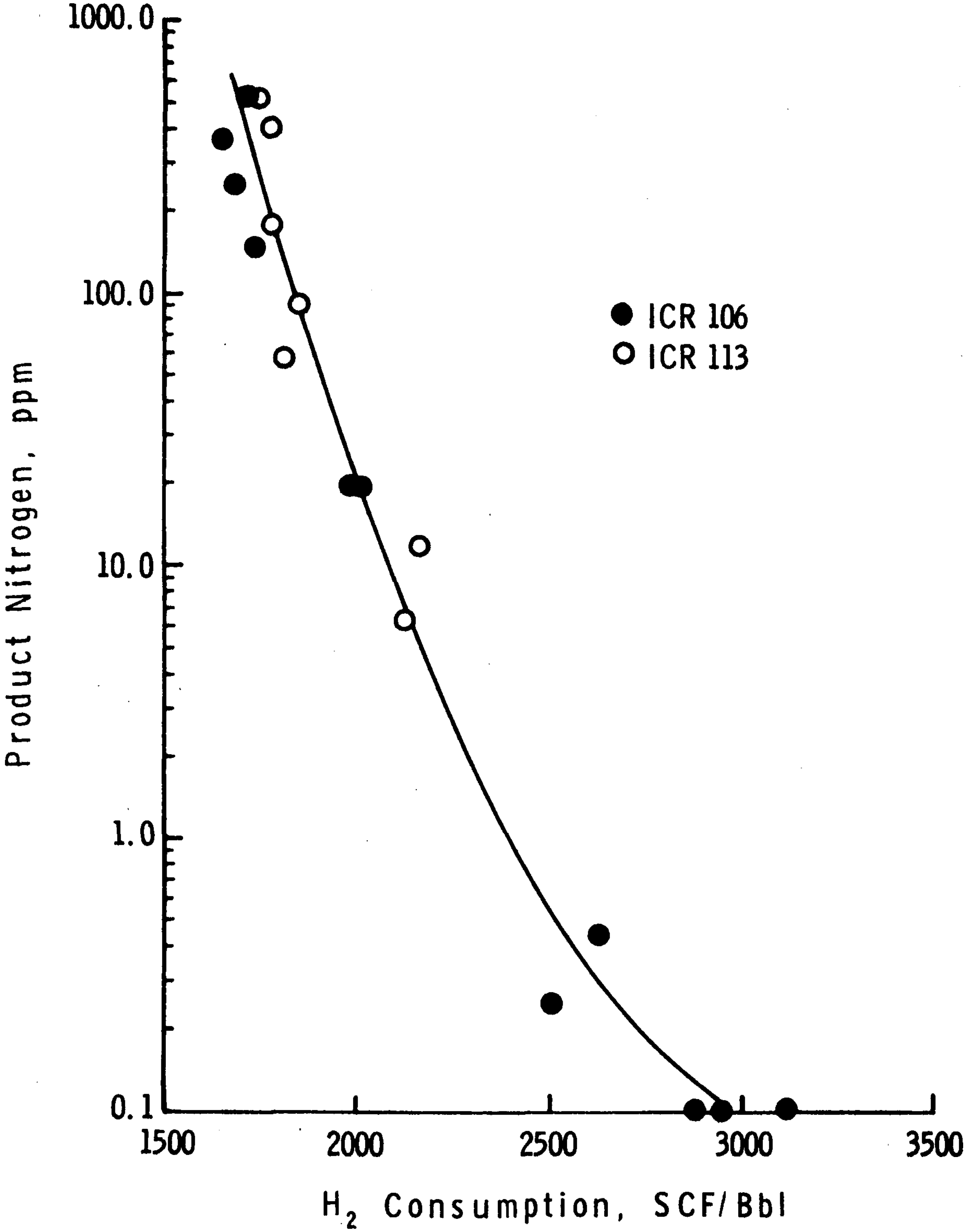

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FIGURE 4

SIMPLIFIED SCHEMATIC OF PILOT PLANT TESTS FOR UPGRADING OF SRC-II PROCESS PRODUCT

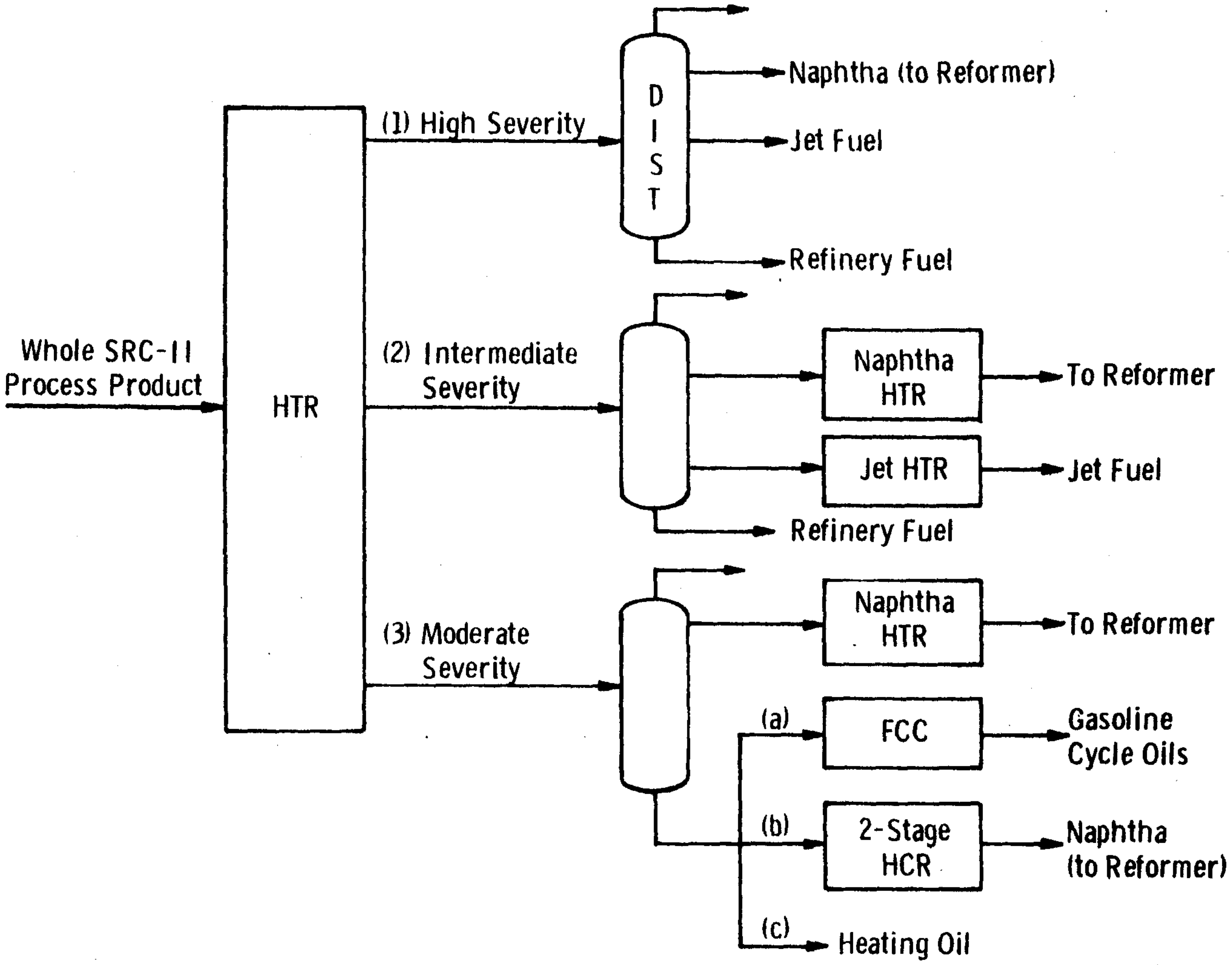

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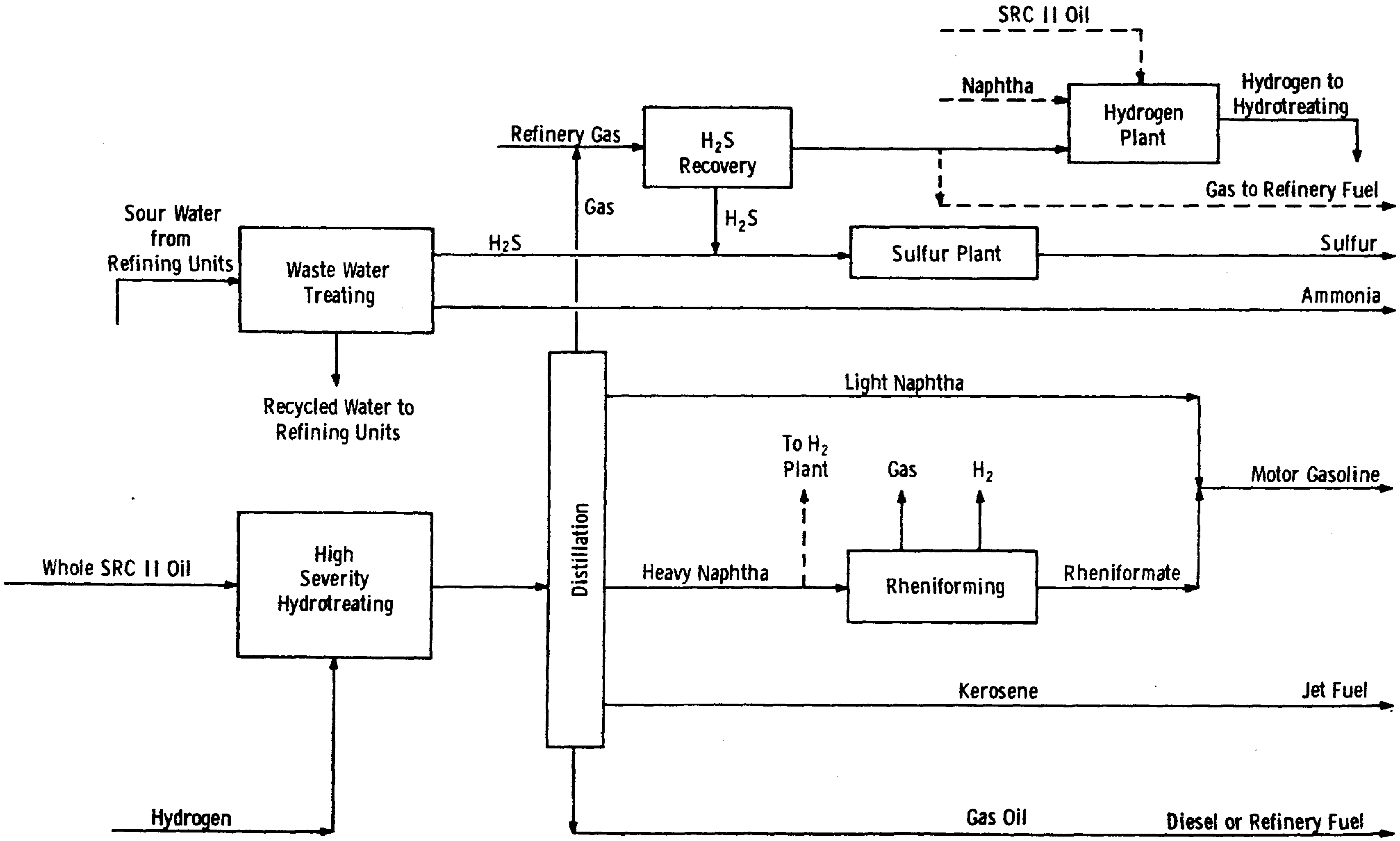
FIGURE 5

SIMPLIFIED FLOW DIAGRAM
REFINING OF SRC II OIL BY HIGH SEVERITY HYDROTREATING



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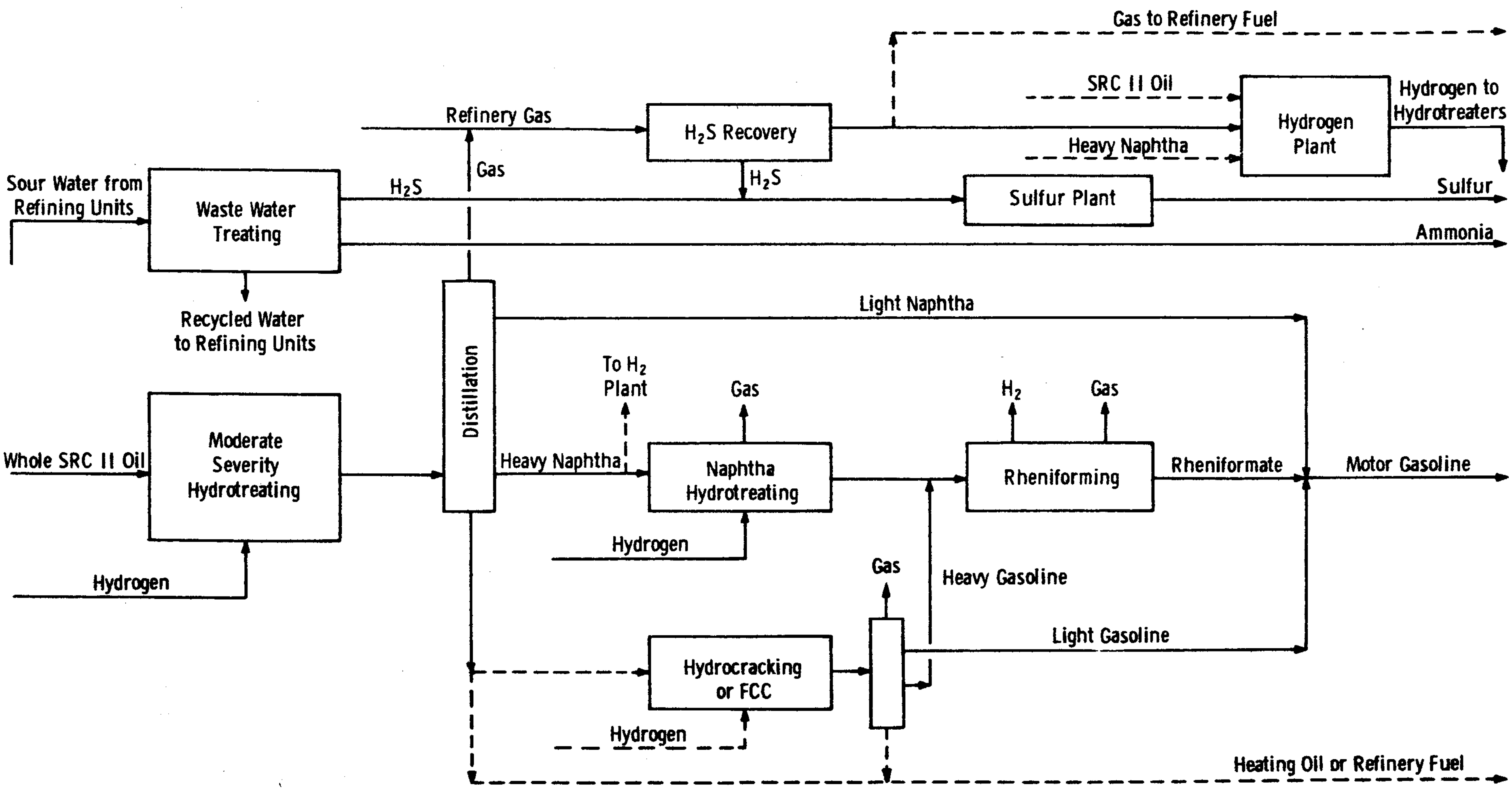
PROCESS ENGINEERING DEPARTMENT
PROCESS DESIGN DIVISION
HAF RD 792200

FIGURE 6

SIMPLIFIED FLOW DIAGRAM
REFINING OF SRC II OIL BY MODERATE SEVERITY HYDROTREATING



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INDIRECT SUPPLY OF TRANSPORTATION FUELS FROM SRC–II PRODUCTS

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CHEMICALS AND MINERALS DIVISION
REPORT NO. 627RK010
JULY, 1979

INDIRECT SUPPLY OF TRANSPORTATION FUELS RESULTING FROM THE USE OF SRC-II FUELS

This volume discusses how SRC-II products could be converted to transportation fuels and how the use of SRC-II products would result in supplying equivalent amounts of transportation fuels indirectly.

As mentioned in the Battelle study in Volume 4 of this series, the use of energy in the United States is divided almost equally among the major sectors: residential-commercial, utility, industrial, and transportation. Of these sectors, transportation is the most inflexible and most demanding in terms of the quality and specifications of the products. Further, the widespread and heavy importance of the auto and truck industry, as well as their usage, to the U.S. economy makes gasoline and diesel fuel of major importance to the energy scene. Hence, transportation products are among the highest priority application of hydrocarbon fuels.

1.0 MARKET DIRECTIONS FOR SRC-II COAL LIQUIDS

The conceptual commercial SRC-II plant is expected to produce a slate of products with fairly clearly indicated applications (Figure 1). An objective of the development of the SRC-II process is to make products which can effectively compete in the refined petroleum products marketplace. These products and their specific applications, as discussed in the previous volumes, especially Volume 1, Deliverable 9, are expected to be more than competitive with other means of utilizing coal in certain applications and, consequently, have a predicted market opportunity. The SRC-II fuel oil products are expected to be used in applications that are fundamentally oil or liquid fuel applications and for which direct application of coal is not economically or otherwise feasible.

The projected market directions that are expected to be initially most logical for SRC-II coal liquids (shown in Figure 2) are thus toward refined petroleum product applications and not toward coal applications, such as some large utility or process industry boilers.

Residual petroleum is not a necessary product from petroleum refining. Displacement of residual petroleum by direct application of coal in large utility or process industry boilers, where environmental and coal handling factors permit, and by SRC-II fuel oil in the East Coast urban areas, is probably a less expensive method of making more refined petroleum products available. Refineries would convert the displaced residual fuel oils to lower boiling products. This alternative, of course, requires expansion and modification of the current domestic petroleum refining capacity to provide for residual processing capability.

2.0 HIGH VALUE PETROLEUM PRODUCTS BY DISPLACEMENT AND SUBSTITUTION

SRC-II coal liquefaction has the attractive potential for making available additional higher value of higher priority refined petroleum products, such as gasoline, diesel fuel, jet fuel, and home heating oil. The SRC-II process has this potential by supplying the products indicated in Figure 1, which could penetrate selected petroleum markets, as shown in Figure 2.

Figure 2 also indicates that it is not technically appropriate to use raw SRC-II products as petroleum refinery feedstock. Broad conversion of coal liquids to transportation quality products requires a substantially different technology than exists today at petroleum refineries, although the technology is available.

The potential for making available higher value petroleum distillate products from the SRC-II process is depicted in further detail in Figure 3; specifically:

METHANE

Methane from the SRC-II process could be used directly in the existing natural gas distribution network to alleviate demand for home heating oil, or its equivalent No. 2 oil in industrial applications, thereby making it more available elsewhere. The related petroleum middle distillate products, diesel fuel and jet fuel, would also be made more available.

Figure 1

PRODUCTS FROM CONCEPTUAL COMMERCIAL SRC-II PLANT

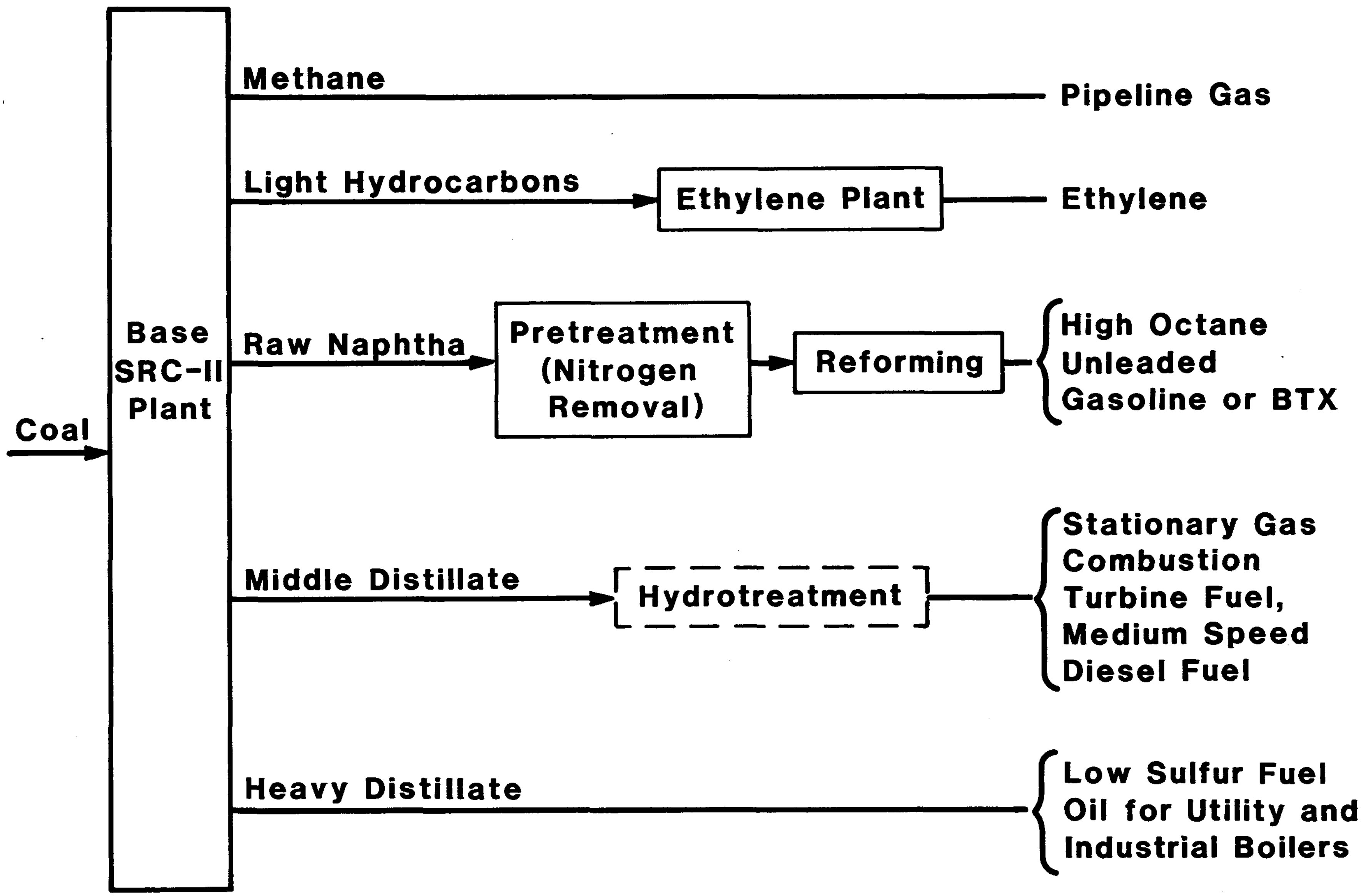


Figure 2

MARKET DIRECTIONS FOR SRC-II COAL LIQUIDS

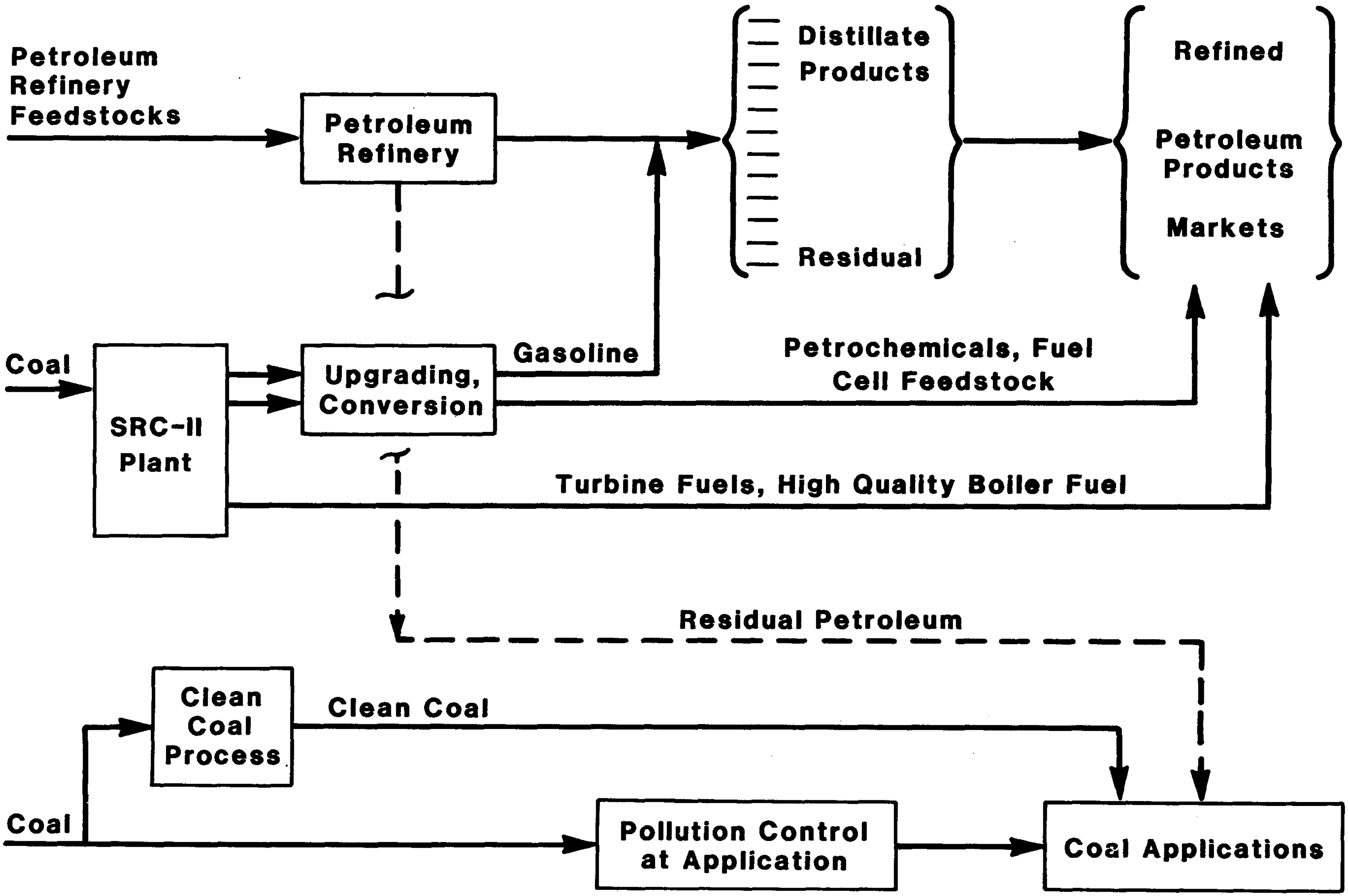
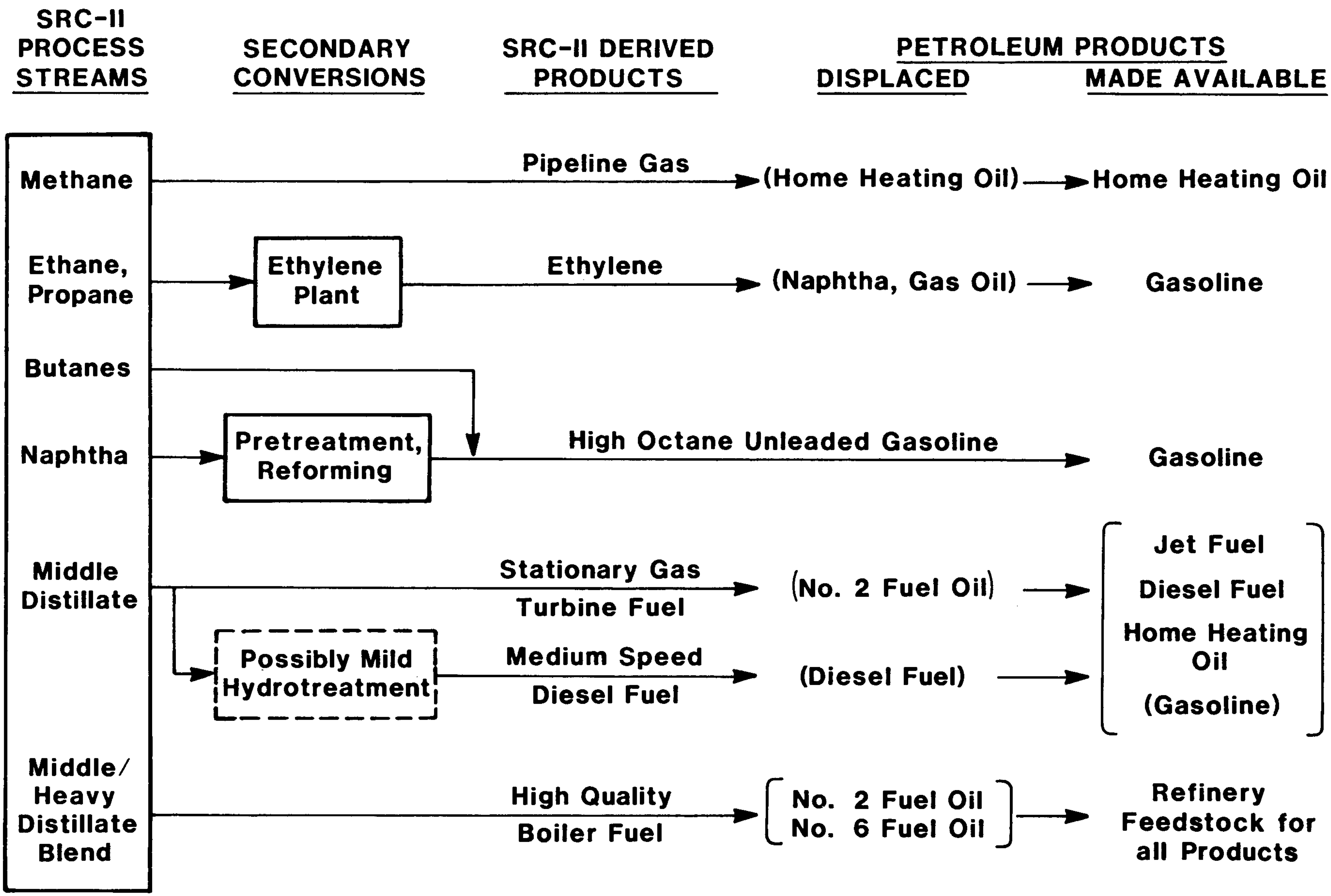


Figure 3

SRC-II COAL LIQUEFACTION SUPPLYING PETROLEUM DISTILLATE PRODUCTS, NEAR TERM ECONOMIC CASE



ETHANE AND PROPANE

Ethane and propane could most effectively be used for conversion to ethylene, the principal building block of the petrochemical industry, at potentially attractive economics, which would displace the petroleum naphtha and gas oil feedstocks and in turn, through petroleum refining, make available more gasoline and middle distillate products (No. 2 oil, jet ful, diesel fuel) from the otherwise same amount of petroleum feedstocks.

RAW SRC-II NAPHTHA

Raw SRC-II naphtha is excellent gasoline reformer feedstock once pretreated to remove nitrogen and the remaining sulfur. These steps could be accomplished commercially at competitive economics with regard to crude petroleum refining, based on current estimated costs for SRC-II naphtha. Thus the SRC-II naphtha could directly augment the supply of high octane unleaded gasoline as well as potentially enable higher yields of gasoline per unit crude charge at the petroleum refinery.

SRC-II FUEL OIL, MIDDLE DISTILLATE

SRC-II fuel oil in the middle distillate range could potentially substitute for No. 2 fuel oil in electric utility combustion turbines, thereby making available more middle distillate petroleum products, such as jet fuel, diesel fuel, and home heating oil. Also, the SRC-II middle distillate, possibly with mild hydrotreating, could be a medium-speed diesel fuel for railroad locomotive and domestic marine applications, thereby potentially releasing directly an equivalent amount of higher quality petroleum-derived diesel fuel necessary for truck application.

SRC-II FUEL OIL, HEAVY DISTILLATE

SRC-II middle and heavy distillate fuel oil blend can be used as an environmentally acceptable industrial and utility boiler fuel, meaning that No. 2 through No. 6 oil could be displaced for refining into a full slate of petroleum distillate products.