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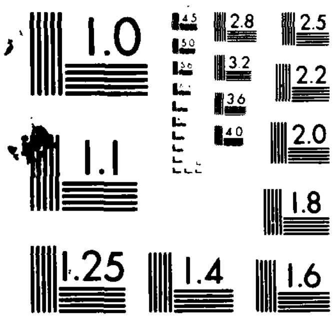

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AFWAL-TR-81-2087
Part II

AN EXPLORATORY RESEARCH AND DEVELOPMENT PROGRAM LEADING TO SPECIFICATIONS FOR AVIATION TURBINE FUEL FROM WHOLE CRUDE SHALE OIL

AD A112682

Part II. Process Variable Analyses and Laboratory Sample Production

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P.O. Box 1135
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September 1981

INTERIM REPORT FOR PERIOD 1 JULY 1979 - 1 NOVEMBER 1980

AERO PROPULSION LABORATORY
AIR FORCE WRIGHT AERONAUTICAL LABORATORIES
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This technical report has been reviewed and is approved for publication.

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Shale oil	Hydrocracking	Jet Fuel Specifications
Paraho Shale Oil	Hydrorefining	Processing Schemes
Occidental Shale Oil	Hydrogen Chloride Treatment	Investment Costs
JP-4 Jet Fuel	Solvent Extraction	Operating Costs
JP-8 Jet Fuel	Dimethylformamide	Product Costs

20. ABSTRACT (Continue on reverse side if necessary and identify by block number)

Part II - Pilot plant process data have been incorporated in three design bases for manufacturing military fuels from raw Occidental shale oil. Processing schemes for 90,000 BPCD refineries to maximize either JP-4, JP-8 or to produce JP-4 plus other military fuels are presented. The processing sequence comprises moderate severity hydrotreating, fractionation, anhydrous HCl extraction and hydrocracking. Plant capacities and product yields were not optimized. Investments for the three refinery options considered are 1.5 to 2.0 times as much as a comparable size petroleum fuels refinery. At maximum — next page

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JP-4 or JP-8 production the yields are about 87 and 53 volume % of total refinery energy input, respectively. Overall, refinery thermal efficiency is $\geq$75%. Inspection data are presented for five samples of specification aviation turbine fuels prepared from pilot plant operations.

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FOREWORD

This interim report details the results of SUN TECH'S studies in Phase II of this contract.

Process Variable Analyses and Laboratory Sample Production was carried out under Contract F33615-78-C-2024. The program is sponsored by the Aero Propulsion Laboratory, Air Force Wright Aeronautical Laboratory, Wright-Patterson AFB, Ohio under project 2480, Task 00 and work unit 01 with Ms. Eva Conley/AFWAL/POSF as the assigned Project Engineer.

Phase II work reported herein was performed during the period of 1 July 1979 to 1 November 1980 under the direction of Dr. Abraham Schneider, Scientific Advisor, SUN TECH, INC. This report was released by the authors in September 1981.

SUN TECH'S program manager wishes to express his appreciation to Dr Herbert Lander and Ms. Eva Conley for their assistance in overcoming administrative and logistical problems associated with this project.

The author gratefully acknowledge the contributions of E. J. Janoski in developing the HCl extraction process, A. Macris for assistance in hydrocracker model verification, and J. J. vanVenrooy for pilot plant operations.

This report is Part II of five planned parts of an exploratory research and development program leading to specifications for aviation turbine fuel from whole crude shale oil. Part I, Preliminary Process Analyses, evaluated three different technically feasible processing schemes proposed by SUN TECH, INC., for converting 100,000 BPCD of raw Paraho shale oil into military turbine fuels. Other parts will follow as the different phases of the program are completed.

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LIST OF SYMBOLS AND ABBREVIATIONS

SYMBOLS

Bbl/SD	Barrels per Stream Day
¢/Gal	Cents per Gallon
$/B	Dollars per Barrel
$/CD	Dollars per Calendar Day
LTSD	Long Tons per Stream Day
# PSD	Pounds per Stream Day
SCF H_2/SD	Standard Cubic Feet Hydrogen per Stream Day
STSD	Short Tons per Stream Day
Vol. %	Volume percent
Wt. %	Weight percent

ABBREVIATIONS

AGO	Atmospheric Gas Oil Fraction
API	American Petroleum Institute
BPCD	Barrels per Calendar Day
BPSD	Barrels per Stream Day
BR	Boiling Range
BTU's	British Thermal Units
CS	Centistokes
DCF	Discounted Cash Flow
DMF	n,n-Dimethylformamide

LIST OF SYMBOLS AND ABBREVIATIONS (Cont'd.)

FOE	Fuel Oil Equivalent
H_2	Hydrogen Gas
HCl	Anhydrous Hydrogen Chloride
HP Sep	High Pressure Separator
H_2S	Hydrogen Sulfide Gas
KV	Kinematic Viscosity
LHSV	Liquid Hourly Space Velocity
LP Sep	Low Pressure Separator
N_2	Nitrogen
NA	Not Available
NH_3	Ammonia Gas
O_2	Oxygen Gas
ppm	Parts per Million by Weight
pp	Partial Pressure
psig	Pounds per Square Inch Gage Pressure
R-1	First Reactor
R-2	Second Reactor
RSO	Raw Shale Oil
S	Sulfur
TBP	True Boiling Point Distillation
TPO	Texaco Partial Oxidation Process
VGO	Vacuum Gas Oil Fraction
WTD	Weighted
WWT Plant	Waste Water Treating Plant

SECTION I

SUMMARY

This interim report covers work performed by Sun Tech, Inc. in Phase II of our contract with the United States Air Force. The Phase II work incorporates laboratory and pilot plant data generated to prepare design bases for manufacturing military fuels from raw Occidental shale oil. Three different processing schemes were developed and are compared with estimates made in Phase I.[1]

The high nitrogen, oxygen, and arsenic contents of raw shale oil present special problems not encountered in refining conventional petroleum. Considerable effort was expended in selecting and evaluating non-proprietary catalysts for use in the various catalytic processing units. A six month main hydrotreater and guard case catalyst aging run was made using both Occidental and Paraho shale oils. An additional run of one month's duration followed at high operating severity with Occidental shale oil. Based on these results we estimate the main hydrotreater catalyst life to be one year and the guard case life to be 6 months. HCl treating was selected as the most effective of three extraction processes evaluated for removing organic nitrogen from hydrotreated shale oil distillates. Hydrogenation severity was varied to yield sufficient HCl extract to balance overall refinery hydrogen requirements. Hydrocracking was incorporated into the processing scheme to maximize yields of military fuels. Modification of Sun Tech's Hydrocracking Model was required to fit the

non-proprietary catalyst's denitrogenation, hydrogenation, and cracking activity parameters to this shale oil derived feedstock.

Using material produced in our pilot plant program, five 500-ml. samples of military turbine fuels of varying characteristics were prepared for laboratory testing.

Improved processing information, the use of a different feedstock, and increasing the total nitrogen content in the main hydrotreater effluent from 2000 to 5000 ppm resulted in lower plant investments than predicted in Phase I. Total plant investments ranged from $841 million for the JP-4 plus other fuels case to $859 million for maximum JP-4 production. Direct plus indirect manufacturing costs varied from $3.91 to $3.99 per bbl of liquid product. Total product costs including the adjusted crude costs were $1.00/gal of product for maximum JP-4; $1.02/gal of product for maximum JP-8; and $1.03/gal of product for the JP-4 plus other fuels case. Based on total energy input to the refinery, 86.8 volume % jet fuel is produced when maximizing JP-4; 52.8 volume % jet fuel when maximizing JP-8; and 65.3 volume % jet fuel in the JP-4 plus other fuels case.

Plant investments for the three shale oil refineries were between $7643 to $7809 per SDB of raw shale oil. Compared to a conventional petroleum refinery, the higher costs result from the need to hydrotreat 100% of the crude to the processing units, plus the need to manufacture all of the hydrogen required. The major portion of the hydrogen required is produced by partial oxidation, which is considerably more expensive than steam reforming.

SECTION II

INTRODUCTION

The purpose of the Phase II program is to demonstrate Sun Tech's concept for processing raw shale oil into high yields of aviation turbine fuels. We have been working on this program since early 1979. In July, 1979, Sun Tech completed Phase I of this program having evaluated on paper three different processing schemes for converting 100,000 barrels per calendar day of Paraho shale oil into aviation turbine fuels. In Phase II, the Phase I processing schemes were evaluated in the pilot plant using Occidental shale oil.

Sun Tech's processing concept for economically refining raw shale oil into aviation turbine fuels consists of six distinct steps: (1) hydrotreating the whole crude shale oil to partially reduce the high total nitrogen content (and convert some neutral nitrogen to basic nitrogen), while minimizing hydrogen consumption; (2) distilling the hydrotreated product into appropriate fractions for additional processing; (3) hydrotreating the light distillate fraction to meet product specifications; (4) treating the wide boiling distillate fraction with anhydrous hydrogen chloride which yields a raffinate and extract phase--the nitrogen content in the HCl raffinate is lowered and concentrated in the extract phase; (5) thermally decomposing the HCl extract to recover anhydrous hydrogen chloride and the recovered HCl-free nitrogen rich extract fraction is used for generating hydrogen by partial oxidation; and (6) hydrocracking the raffinate fraction to maximize the yield of aviation turbine fuels. This processing scheme is shown schematically in Figure 1. The slate of military fuels is optional and they can be produced to meet or exceed current military specifications.

SECTION III

PROCESS DETAILS

1. SHALE OIL CHARACTERIZATIONS

Sun Tech has evaluated two different shale oils during the course of Phase II. The predominent feedstock used was Occidental (modified in-situ) shale oil. Paraho shale oil obtained from a directly heated surface retort was also evaluated. Table 1 presents inspections and analyses for both Occidental and Paraho shale oils. Occidental shale oil can be processed using less severe conditions than those required for Paraho shale oil because of its lower boiling range, lower nitrogen and sulfur contents, and higher hydrogen content. Both shale oils contain significant quantities of arsenic not found in conventional petroleum; the nitrogen and oxygen contents of raw shale oil are also higher than those found in conventional petroleum.

2. RAW SHALE OIL HYDROTREATER

A simplified flow diagram of the raw shale oil hydrotreater and distillation plants is shown in Figure 2. The use of guard reactors is necessary to remove arsenic and iron, as well as to saturate olefins in the feed. A vacuum still is used to produce a gas oil fraction with a 1000°F end point. The waxy nature of the 1000°F+ bottoms precludes its use in the HCl treating step due to the formation of emulsions. Operating conditions used in the raw shale oil hydrotreater are given in Table 2.

Less severe conditions were used in Phase II, with whole Occidental shale oil, than used with Paraho shale oil in Phase I which increased the nitrogen content from 2000 to 5000 ppm in the effluent. A total nitrogen content of 5000 ppm in the hydrotreated product was chosen in order to produce sufficient HCl extract for hydrogen manufacture by partial oxidation. The less severe operating conditions resulted in lower hydrogen consumption and a lower C_4+ product yield. Two additional levels of hydrogenation severity, producing 2200 and 6400 ppm total nitrogen in the reactor effluent, were also evaluated and will be incorporated in Sun Tech's math model for process optimization.

a. Catalyst Life Studies

A two reactor isothermal pilot plant was employed to determine catalyst aging characteristics in the R-1 guard reactor and the R-2 hydrotreater reactor. The catalyst aging curve, Figure 3, shows that after the loss of the initial high activity characteristic of fresh catalysts, the temperature required in the R-2 catalyst bed to hydrotreat whole Occidental shale oil to 5000 ppm total nitrogen in the reactor effluent remained essentially constant. Almost four months of successful life-testing was accumulated with Occidental shale oil. Catalyst activity tests were run periodically to determine the average catalyst temperature required to produce 5000 ppm total nitrogen in the reactor effluent. Most of the on-stream time employed more severe operating conditions producing 2200 ppm total nitrogen. A minor portion of the time produced material containing 6400 ppm total nitrogen. The R-1 guard reactor catalyst bed was kept at an average temperature of 650°F.

Using the same catalyst loading that had accumulated almost four months of life with Occidental shale oil, an additional two month life test with Paraho shale oil was completed. Since the Paraho feed contained 2.13 wt.% total nitrogen as opposed to the 1.46 wt.% total nitrogen content found in Occidental shale oil, a 50°F increase in R-2 average catalyst bed temperature was required to yield a hydrotreated product containing 5000 ppm total nitrogen (see Figure 4). At this point the feed was changed back to Occidental shale and the activity checked. During the two months the unit was operated on Paraho shale oil, the catalyst activity aged 10°F. Based on the stable aging characteristics of the catalyst in R-2, a life expectancy of 1 year is projected; for R-1 we project a 6-month catalyst life. Arsenic content in the R-1 effluent varied between 0 and 1 ppm. Finally, an additional one-month long run was made employing severe operating conditions producing less than 5 ppm total nitrogen in the reactor effluent. During this period of severe operation, some catalyst activity loss was apparent.

b. Material Balance Summaries

Material balance summaries for the main hydrotreater and distillation units are given in Table 3. Significant quantities of ammonia, water, and hydrogen sulfide are produced during hydrogenation. Cut points for the distillation unit are varied depending on the type of operation, JP-4 production or JP-8 production. Product inspections on the streams from the main hydrotreater distillation units are shown in Table 4. Nitrogen, sulfur, and aromatic contents increase with increasing boiling range. Very little material is found boiling below 250°F in the hydrotreated product.

3. NAPHTHA HYDROTREATER

The purpose of the naphtha hydrotreater, shown schematically in Figure 5, is to clean up the light distillate from the atmospheric distillation unit in order to meet product specifications. The effluent is passed through a product stripper (not shown) before blending into final products. Operating conditions used in the naphtha hydrotreater are given in Table 5. Due to the higher levels of nitrogen present in our feed to the unit, hydrogen consumption is up considerably from Phase I predictions (Paraho @ 2000 ppm total nitrogen). Material balance summaries for the JP-4 and JP-8 operations are presented in Tables 6 and 7 respectively. In the JP-4 case, feedstock and product boiling ranges are 180-490°F. In the JP-8 case, the feedstock boiling range is 180-550°F; however, the hydrotreated products consisted of a C_4-290°F gasoline blendstock and the 290-550°F JP-8 product.

4. EXTRACTION PROCESSES

Three alternate processes for removal of nitrogen compounds remaining in mildly hydrotreated shale oil were evaluated. DMF and methanol appear to be about equal for extracting nitrogen compounds from light distillates (700°F end point) derived from mildly hydrotreated Occidental shale oil. These solvents would be useful for removing nitrogen compounds in the

JP-4 through diesel fuel #2 (DF-2) boiling range. Above 700°F, these solvents were only marginally effective exhibiting poor selectivity for nitrogen removal. HCl treatment of the 450-1000°F distillate fractions of hydrotreated shale oil was more effective for removal of nitrogen containing compounds than either DMF or methanol extraction. Therefore, HCl extraction was the process chosen for Phase II to remove nitrogen compounds from high boiling fractions of mildly hydrotreated Occidental shale oil.

a. HCl Treating

Pilot plant HCl treating was carried out batchwise. Due to the smooth operation of these runs, we feel that the process can be readily adapted to continuous operation and achieve similar results. A schematic flow diagram of a continuous HCl extraction plant is shown in Figure 6.

Operating conditions for HCl treating and material balance summaries for the JP-4 operation are presented in Tables 8 and 9. Here the gas oil feedstock has a 490-1000°F boiling range and a total nitrogen content of 4800 ppm. Tables 10 and 11 give the operating conditions and material balance summaries for the JP-8 operation. In this case, the gas oil feedstock has a 550-1000°F boiling range and a total nitrogen content of 5600 ppm. For both cases, higher levels of nitrogen present in the gas oils required greater HCl addition than called for in Phase I. Raffinate yields for Occidental shale oil are down, while the extract yield is considerably higher than for the more severely hydrotreated Paraho operation in Phase I. Considerable amounts of chlorides remain in both the raffinate and decomposed adduct. There is a 0.1 volume % loss of raffinate and a 5 weight % loss of anhydrous HCl in the water washing step.

5. RAFFINATE HYDROCRACKING

A single stage hydrocracker is shown in Figure 7. Reactor R-1 is used to clean up the raffinate feed before it enters the main hydrocracking reactor R-2 where most of the hydrocracking takes place. The products taken off the distillation tower can be varied. Extinction recycle of the distillation bottoms is optional.

Table 12 presents the hydrocracker operating conditions for maximum production of JP-4 jet fuel. Originally, we intended to use a proprietary hydrocracking catalyst with which we have had experience. We were barred from using this catalyst for shale oil applcations. After screening three non-proprietary catalysts, a nickel tungsten catalyst designated "B" was selected for this operation. Operating temperatures and pressures are up slightly from Phase I predictions; however, the liquid hourly space velocity in R-2 hydrocracker reactor was increased from 1.0 to 2.0. A material balance summary for the maximum JP-4 operation is given in Table 13. Ammonium chloride formed during the R-1 hydrotreating reaction is removed by the injection of water before the high pressure separator. High yields of JP-4 jet fuel are obtained with the 490°F+ bottoms being recycled to extinction.

Hydrocracker operating conditions and a material balance summary for maximizing JP-8 production are presented in Tables 14 and 15. In this case, 44% of the total liquid product (extinction recycle basis) is C_4-290°F gasoline blendstock. The remainder of the liquid product (56 volume %) is JP-8 jet fuel having a 290-550°F boiling range. Here the 550°F+ bottoms is recycled to extinction. Again, ammonium chloride is removed by injecting water after the R-1 hydrotreater.

Tables 16 and 17 summarize the Phase II hydrocracker operation for production of JP-4 and other fuels. In this operation there is no recycle stream to the R-2 hydrocracker (once-through operation). In addition to JP-4 jet fuel, diesel fuel #2 (DF-2), and a 675°F+ bottoms fuel oil are produced. Since there is no recycle oil to the R-2 hydrocracker, chemical consumption of hydrogen is significantly lower than in the maximum JP-4 case.

6. PRODUCT INSPECTIONS

Specification quality JP-4, JP-8, DF-2, and C_4-290°F gasoline blendstock can be produced by Sun Tech's process to upgrade raw Occidental shale oil. As shown in Table 17, essentially complete removal of nitrogen and sulfur is obtained. Product inspections are presented in Table 18. The blended heavy fuel consists of the 1000°F+ bottoms from the vacuum distillation unit blended with the 675°F+ fuel oil derived from the JP-4 plus other fuels operation. Some nitrogen and sulfur remain in the blended heavy fuel.

7. MATERIAL BALANCE SUMMARIES

Material balance summaries for the three alternate cases are presented in Table 19. Refinery fuel, electricity, and steam were converted to a Fuel Oil Equivalent (FOE) basis. Based on total energy input to the refinery, 86.8 volume % jet fuel is produced when maximizing JP-4; 52.8 volume % jet fuel when maximizing JP-8; and 65.3 volume % jet fuel in the JP-4 plus other fuels case.

8. LABORATORY SAMPLE PRODUCTION

During the course of the Phase II program, five 500 ml samples of prototype jet fuels derived from raw Occidental shale oil were produced. Table 20 presents inspections and analyses of three JP-4 samples: conventional, low aromatics, and high aromatics. The very low freeze point of the 100% hydrocrackate sample demonstrates that catalytic hydrocracking of the HCl raffinate provides a means for meeting jet fuel freeze point specifications. The hydrocrackate can be used by itself or as a blending component for depressing freeze points of other marginal components. Inspections and analyses of two JP-8 samples is given in Table 21. The 100% hydrocrackate had a low flash point of 75°F. This is the result of a distillation error in which too low an initial cut point was taken. This is not a serious problem, as it can be easily corrected by taking a higher initial cut point during the final product distillation.

9. PHASE II ECONOMIC EVALUATION

Guidelines for developing Sun Tech's Phase II economics are given in Table 22. A first quarter 1980 cost base was used instead of the September 1978 base used in preparing Phase I economics. Crude shale oil is valued at \$30/Bbl (vs. \$16/Bbl in Phase I) and all product fuels are equally valued at \$40/Bbl (vs. \$21/Bbl in Phase I). These prices were used for calculating interest charges for working capital.

Plant capacities and investments are presented in Table 23. The main hydrotreater consists of two parallel trains with the effluents fed to a single atmospheric and vacuum distillation plant. The gas oil hydrocracker also consists of two parallel trains with the effluents distilled in a single fractionator. Improved processing information, the use of Occidental instead of Paraho shale oil, and increasing the total nitrogen content in the main hydrotreater effluent from 2000 to 5000 ppm result in lower plant investments than predicted in Phase I. As in Phase I, the main hydrotreater and the Texaco Partial Oxidation (TPO) hydrogen plant account for the majority of the processing facility costs (over 50%). Total capital investments range from \$749.0 million for the JP-4 plus other fuels case to \$763.4 million for maximum JP-4 production. As shown in Table 24, total product costs including the adjusted crude costs are \$1.00/gallon of product for maximum JP-4 production, \$1.02/gallon of product for maximum JP-8 production, and \$1.03/gallon of product for the production of JP-4 plus other fuels. Optimization studies were not performed for this evaluation.

Results of the three processing routes are summarized in Table 25. Plant investments for the three shale oil refineries varied betweeen $7643 and $7809 per SDB of raw shale oil feed. The plant investment for a conventional petroleum fuels refinery of similar capacity is approximately $4000/SDB of crude. The higher costs of the shale oil refineries result from the need to hydrotreat 100% of the crude to the processing units and manufacture all of the hydrogen required. The majority of the required hydrogen is produced by partial oxidation, which is considerably more expensive than steam reforming.

SECTION IV

CONCLUSIONS

1. Phase II pilot plant work has been completed. Results confirm or are more favorable than the estimates made in Phase I. High yields of military fuels that meet current specifications can be produced from raw shale oil using Sun Tech's hydrogen chloride extraction process. Based on total energy input to the refinery, 86.8 volume % jet fuel is produced when maximizing JP-4; 52.8 volume % jet fuel when maximizing JP-8; and 65.3 volume % jet fuel when producing JP-4 plus other transportation fuels. Non-proprietary catalysts have been incorporated in all catalytic units. Overall thermal efficiencies of 75-76% have beeen attained based on total energy input of crude, fuel, and utilities converted to FOE.

2. Economics have been developed for a 100,000 BPSD shale oil refinery using a first quarter 1980 cost base and a $30 per barrel price for raw shale oil. A tctal product cost of $1.00 to $1.03 per gallon has been estimated which varies with the refinery product slate. Process unit capacities and product distribution have <u>not</u> been optimized in Phase II.

3. Shale oil fuels refineries are much more capital intensive than petroleum refineries of comparable size by a factor of 1.5 to 2.

SECTION V

RECOMMENDATIONS

1. Arsenic disposal may be a problem. The spent guard reactor catalyst may contain up to 15 weight % arsenic accumulated during operation. Further work is recommended in order to find an acceptable method of disposal.

2. The chloride content in the HCl raffinate has varied from 70 to 750 ppm. Since chlorides in the raffinate feed complicate the subsequent hydrocracking operation and reduce by-product ammonia yield, additional effort is recommended to minimize the chloride content.

3. Pilot plant HCl extraction was carried out batchwise. Continuous HCl treating is recommended to confirm or modify Phase II data.

4. Extinction recycle of the hydrocracker bottoms has not been demonstrated in the pilot plant. This type of operation may not be feasible due to buildup of polynuclear aromatics or wax in the recycle oil and a drag stream may be required. It is recommended that pilot plant hydrocracking with extinction recycle be carried out to firm up Phase II data generated by Suntech's Hydrocracking Math Model.

5. A two-year catalyst life has been estimated for the R-2 hydrocracking catalyst. A catalyst life study is recommended in order to confirm or modify this estimate. This type of study would also be useful in revealing possible corrosion problems encountered due to the presence of chlorides in the HCl raffinate.

REFERENCES

1. H. E. Reif, J. P. Schwedock, and A. Schneider, "An Exploratory Research and Development Program Leading to Specifications for Aviation Turbine Fuels from Whole Crude Shale Oil, Phase I - Part I AFWAL-TR-81-2087 - Preliminary Process Analyses", Report prepared for the Department of Defense U.S. Air Force by Sun Tech, Inc., under contract No. F33615-78-C-2024, 1981.

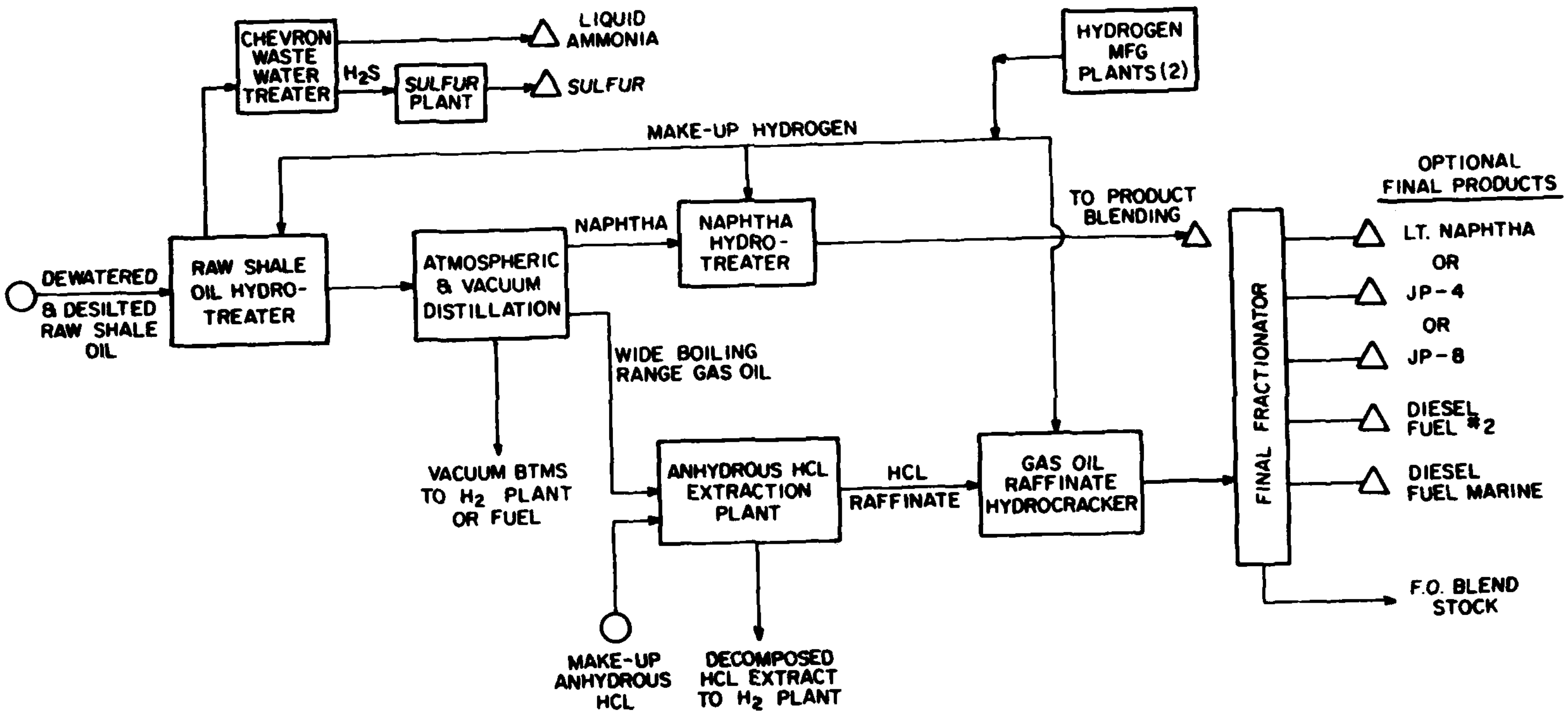

Fig. 1 SCHEMATIC FLOW DIAGRAM FOR REFINING RAW SHALE OIL USING ANHYDROUS HCL EXTRACTION

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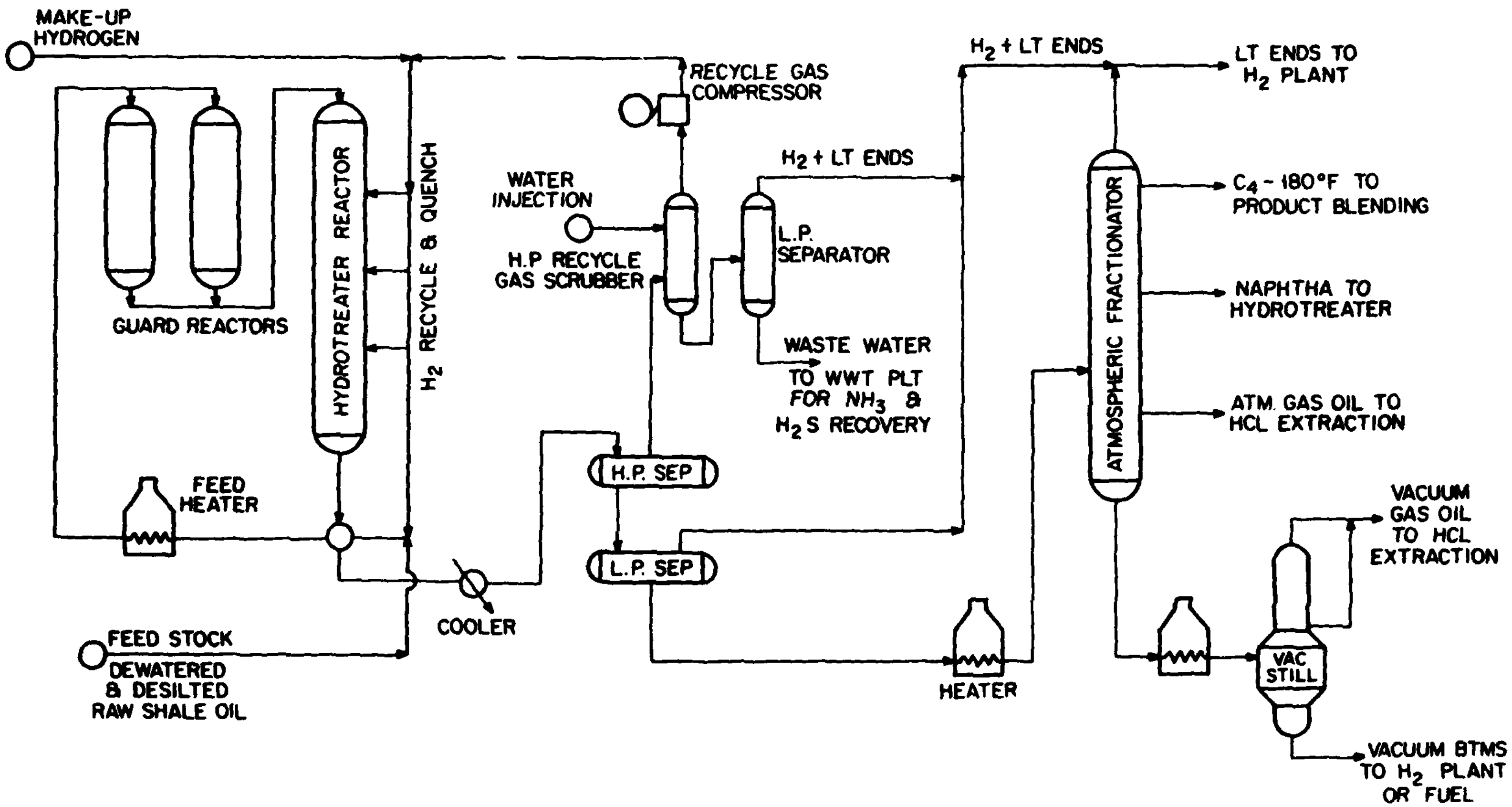

Fig. 2 SIMPLIFIED FLOW DIAGRAM OF RAW SHALE OIL HYDROTREATER AND DISTILLATION PLANTS

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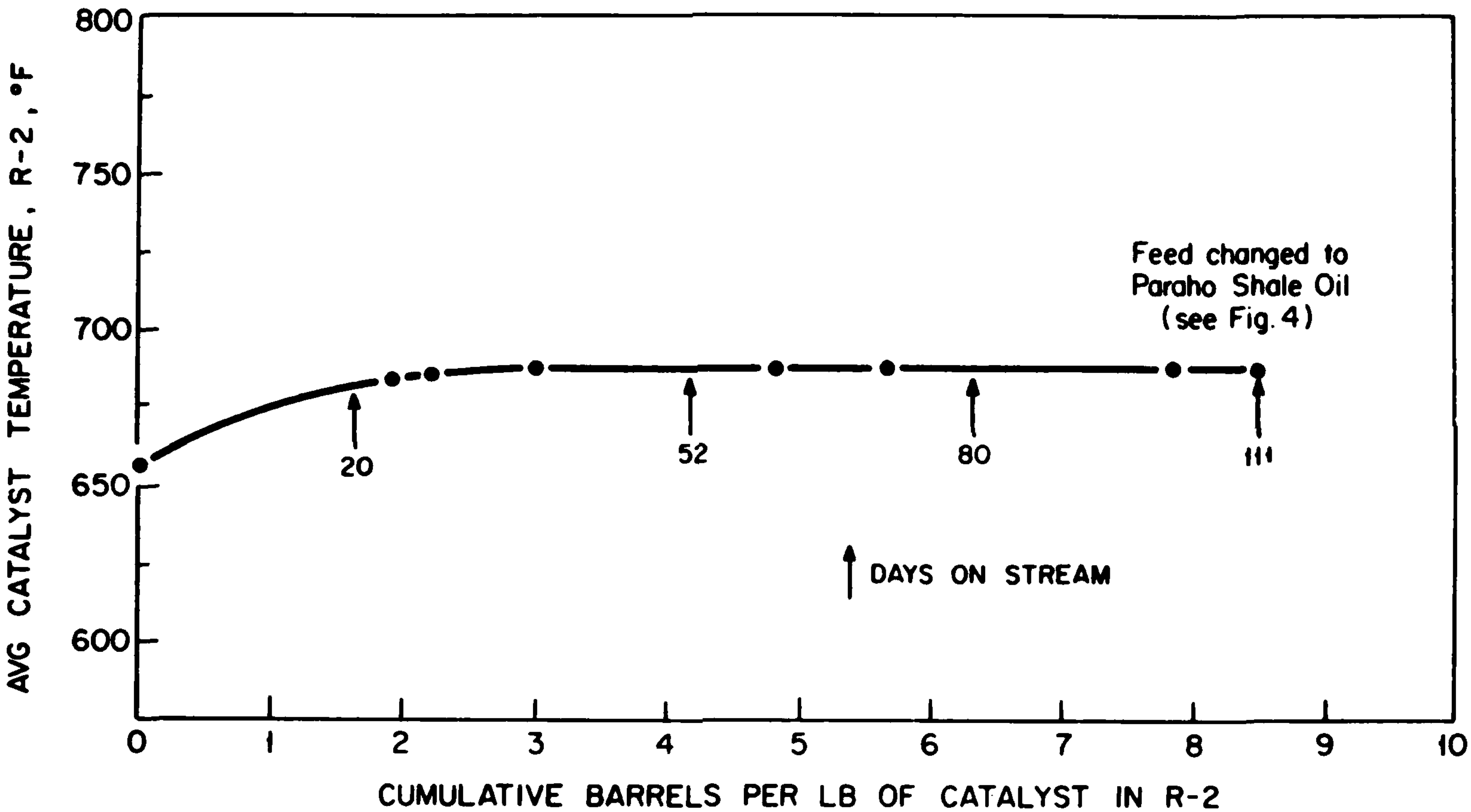

Fig. 3 CATALYST LIFE TEST FOR HYDROTREATING WHOLE OCCIDENTAL SHALE OIL

(TO 5,000 ppm N_T PRODUCT)

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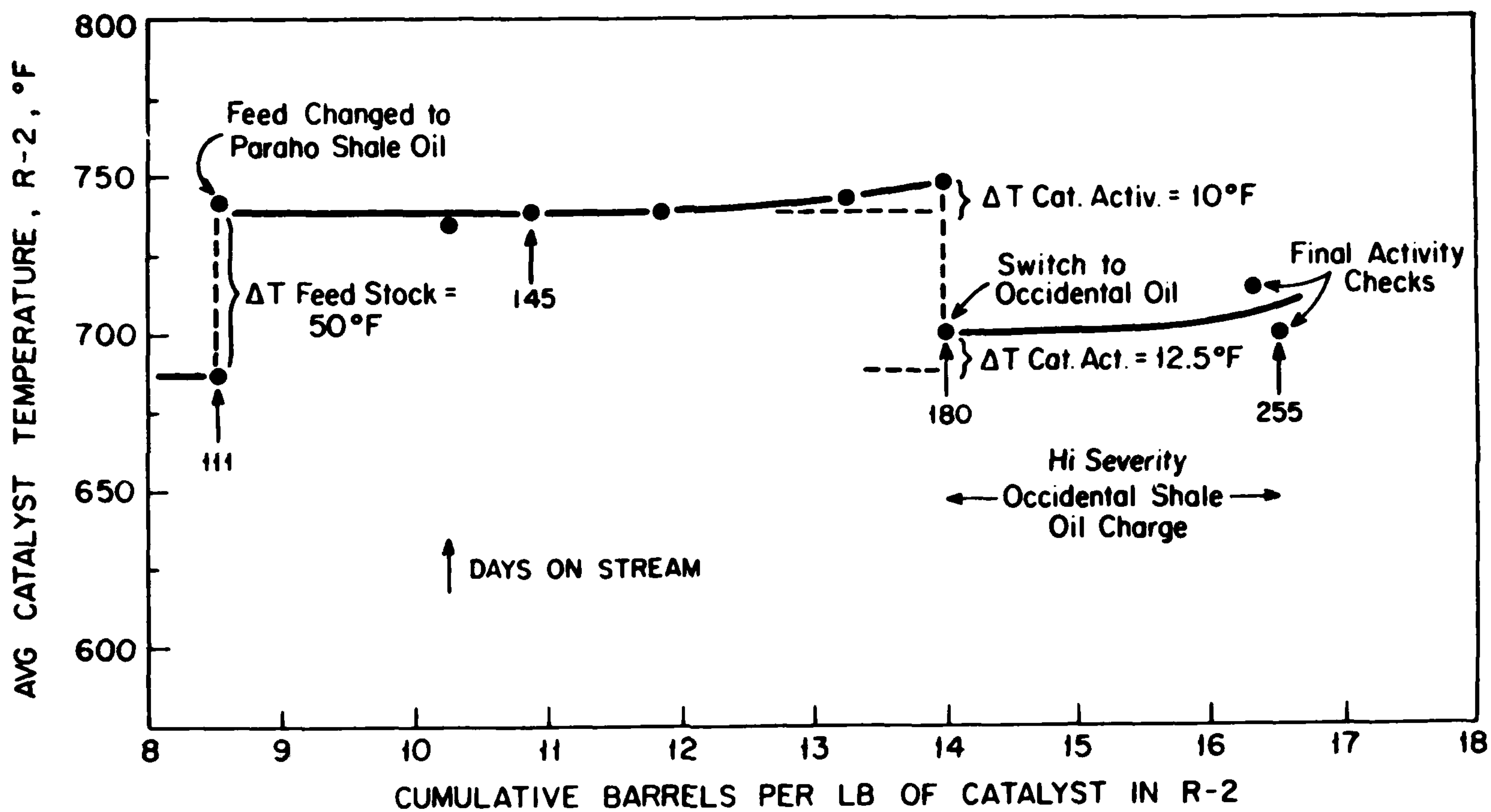

Fig. 4 CATALYST LIFE TEST FOR HYDROTREATING WHOLE PARAHO SHALE OIL (TO 5,000 ppm N_T PRODUCT)

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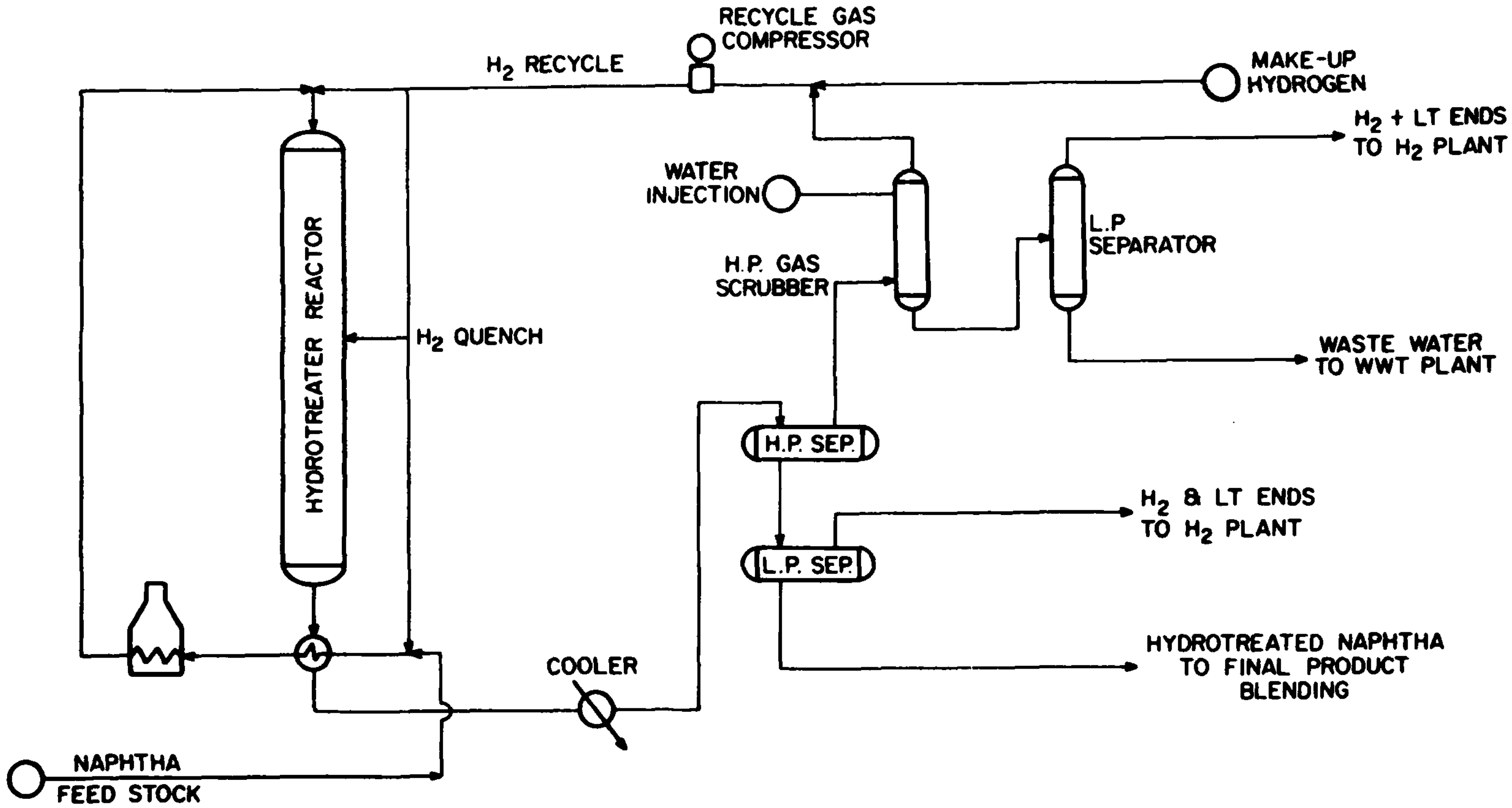

Fig. 5 SCHEMATIC FLOW DIAGRAM OF NAPHTHA HYDROTREATER

10/10/80

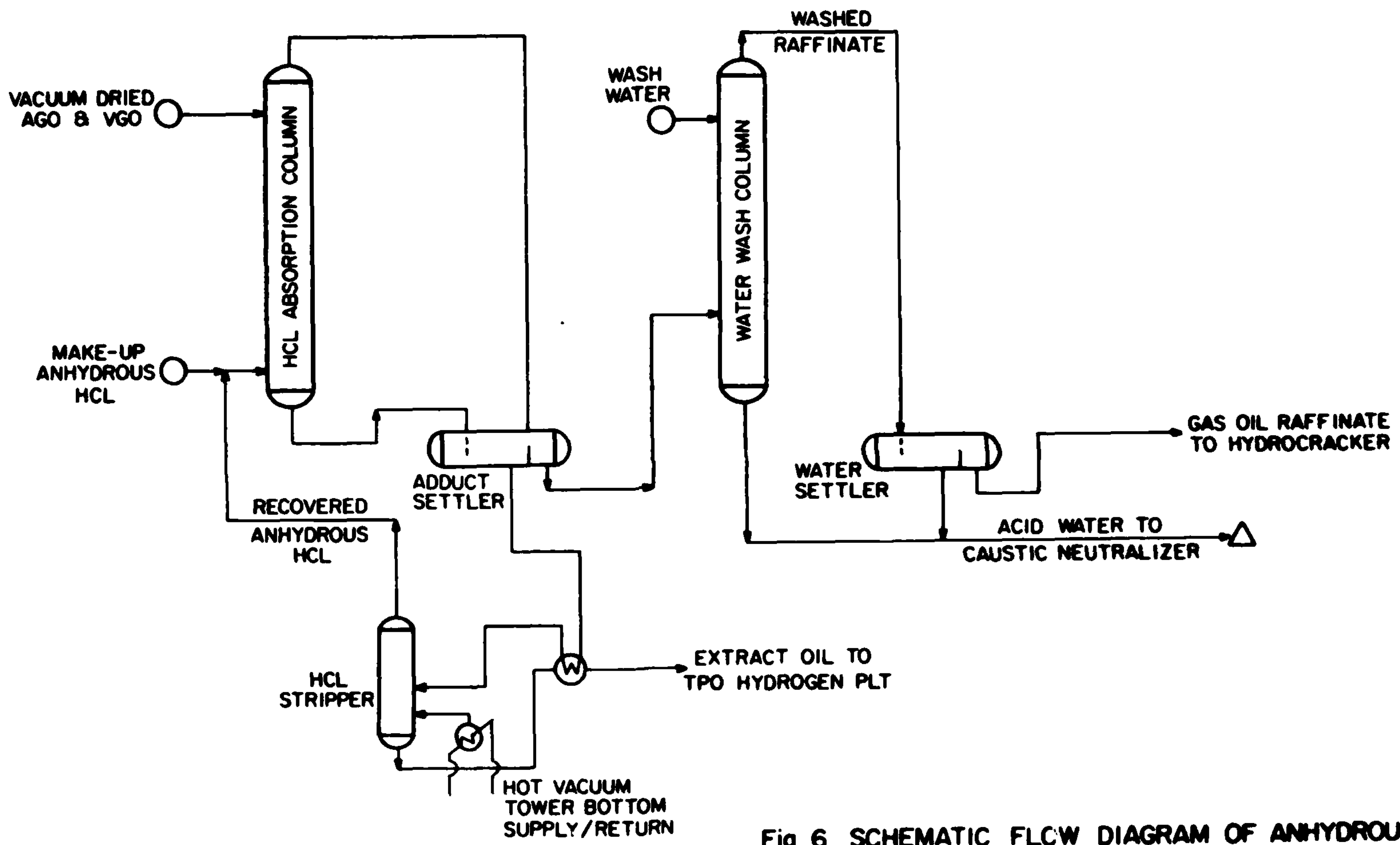

Fig. 6 SCHEMATIC FLOW DIAGRAM OF ANHYDROUS HCL EXTRACTION PLANT

10/10/80

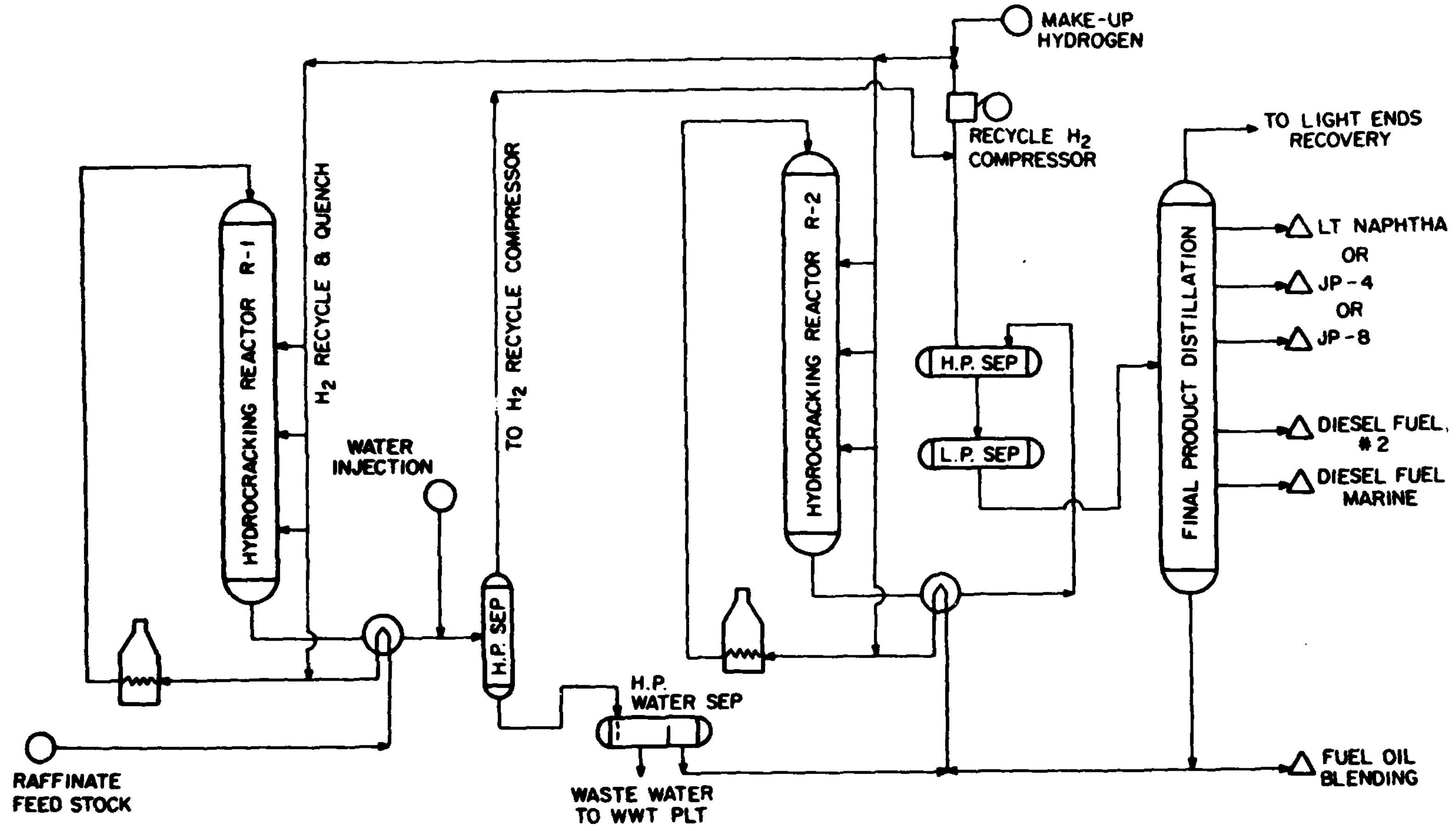

Fig. 7 SCHEMATIC FLOW DIAGRAM OF SINGLE STAGE HYDROCRACKER FOR MANUFACTURING MILITARY FUELS FROM SHALE OIL

10/10/80

TABLE 1

INSPECTIONS AND ANALYSES OF RAW SHALE OIL

Raw Shale Oil	Paraho	Occidental
Inspection Data		
API @ 60°F	20.6	23.0
Specific Gravity 60/60	0.9303	0.9160
Viscosities, KV		
@ 100°F, cs	60	32.3
@ 210°F, cs	5.38	4.82
Distillation, °F	D1160	D2887
IBP	133	296
10 Vol. %	508	459
30 "	687	558
50 "	798	649
70 "	918	768
90 "	1057	876
FBP	1065/95%	1071
Ramsbottom Carbon Res., Wt.%	1.4	-
Asphaltenes, Wt.%	-	2.4
Chemical Composition Data, Wt.%		
Carbon	83.83	84.82
Hydrogen	11.72	12.04
Oxygen	1.31	1.18
Nitrogen (Total)	2.13	1.46
(Basic)	1.31	0.81
Sulfur	0.75	0.62
Iron, ppm	90	NA
Arsenic, ppm	34	33

TABLE 2

OPERATING CONDITIONS FOR PROCESSING WHOLE OCCIDENTAL SHALE OIL

BASIS:

Charge Rate:	100,000 BPSD (90,00 BPCD)
Operating Factor:	0.90
Catalysts:	NiMo on Spherical Alumina (R-1) NiMo on Alumina (R-2)
Catalyst Life:	6 Months (R-1) 12 Months (R-2)

REACTOR OPERATING CONDITIONS

CASE	PHASE I (PARAHO)	PHASE II (OCCIDENTAL)
LHSV	0.55	1.0 R-1
		1.0 R-2
Avg. Catalyst Temp, °F		
R-1	-	625
R-2	750 (1)	690
Pressure, Total psia	1650	1615
H_2 PP	1400	1520
Recycle Gas Rate, SCF/B	4100	4000
Hydrogen Consumption, SCF/B		
Chemical	1760	1100
Dissolved	150	150
Bleed	75	100
Total to Hydrotreater	1985	1350
PRODUCT DATA		
Total Nitrogen, ppm	2000	5000
Sulfur, ppm	50	140
C_4+ Yield, Vol.% Feed	106.96	103.55

(1) Cat. Life assumed to be 6 months.

TABLE 3

MATERIAL BALANCE SUMMARY FOR MAIN HYDROTREATER AND DISTILLATION UNITS

BASIS:

100,000 BPSD Raw Occidental Shale Oil

135 x 10^6 SCF Hydrogen PSD (110 x 10^6 SCF H_2 Chemically Consumed PSD)

Liquid Effluent Treated to 5000 ppm Total Nitrogen

PRODUCTS, TBP CUT POINTS	JP-4	JP-8
Ammonia, STSD	187	187
Hydrogen Sulfide, Sulfur Eq. STSD	110	110
Unreacted H_2, SCF x 10^6 SCF PSD	25.0	25.0
C_1-C_3 Gases, Lbs. PSD	385,294	385,294
C_4-180°F, BPSD	2,116	
180-490°F, BPSD	24,141	
490-1000°F, BPSD	73,133	
C_4-290°F, BPSD		4,550
290-550°F, BPSD		25,561
550-1000°F, BPSD		69,279
1000°F+ Bottoms, BPSD	4,159	4,159
TOTAL LIQUIDS, BPSD	103,549	103,549

TABLE 4

PRODUCT INSPECTIONS ON STREAMS FROM MAIN HYDRTOREATER DISTILLATION UNIT

FRACTION	180-490°F	490-1000°F	C_4-290°F	290-550°F	550-1000°F	1000°F+ BTMS.
API Gravity @ 60°F	41.5	28.9	71.0	40.6	28.6	16.0
Distillation, °F						
I/10	180/290	490/605	50/145	290/360	550/606	NA
30/50	405/441	661/734	183/202	433/458	671/744	--
70/95	468/486	817/965	217/252	480/526	820/970	--
EP	490	1000	290	550	1000	--
Aromatics, Wt.%	24 Vol.%	42	3 Vol.%	25	45	50
Olefins, Vol.%	3	--	--	3	--	--
Total Nitrogen, ppm	3260	4800	--	3480	5600	7900
Sulfur, ppm	65	140	--	80	150	1220

TABLE 5

OPERATING CONDITIONS FOR NAPHTHA HYDROTREATER

Operator Factor: 0.91 Phase I, 0.90 Phase II

Catalyst: NiMo on Alumina
Catalyst Life: 2 Years

REACTOR OPERATING CONDITIONS:

CASE	PHASE I MAX. JP-4	PHASE II MAX. JP-4	PHASE I MAX. JP-8	PHASE II MAX. JP-8
Feedstock TBP Boiling				
Range, °F	180-450	180-490	180-535	180-550
Total Nitrogen, ppm	620	3260	690	3480
LHSV, V/Hr/V	∿2.0	2.0	∿2.0	2.0
Avg. Catalyst Temp., °F	∿725	750	∿725	750
Total Pressure, psia	1215	1500	1215	1500
H_2 PP	1100	1400	1100	1400
Recycle Gas Rate, SCF/B	4500	4000	4500	4000
Hydrogen Consumption, SCF/B				
Chemical	78	350	96	400
Dissolved	40	50	40	50
Total to Hydrotreater	118	400	136	450
Product				
Total Nitrogen, ppm	4	8	4	8
Sulfur	Nil	2	Nil	2
C_4+ Yield, Vol.%				
Feed	101.28	101.66	100.97	102.24

TABLE 6

MATERIAL BALANCE SUMMARY FOR NAPHTHA HYDROTREATER

JP-4 OPERATION

Basis: 24,141 BPSD of 180-490°F Naphtha Fraction
9.66 x 10^6 SCF Hydrogen PSD (8.45 x 10^6 SCH H_2 Chemically Consumed)

PRODUCTS

Ammonia, STSD	13.7
Hydrogen Sulfide, Sulfur, Eq, STSD	0.2
Unreacted H_2 x 10^6 SCF PSD	1.21
C_1-C_3 Gases, Lbs PSD	13,956
C_4-180°F, BPSD	1,207
180-490°F, BPSD	23,335
TOTAL CUT	24,542

INSPECTIONS ON 180-490°F. CUT

	FEED	PRODUCT
API Gravity @ 60°F	41.5	42.3
Aromatics, Vol. %	24.3	15.0
Olefins, Vol. %	3.0	1.4
Total Nitrogen, ppm	3260	8.0
Sulfur, ppm	65	2.0

TABLE 7

MATERIAL BALANCE SUMMARY FOR NAPHTHA HYDROTREATER

JP-8 OPERATION

BASIS: 27,995 BPSD of 180-550°F Kerosine Fraction
12.60 x 10^6 SCF Hydrogen PSD (11.20 x 10^6 SCF H_2 Chemically Consumed)

PRODUCTS

Ammonia, STSD	17.0
Hydrogen Sulfide, Sulfur Eq, STSD	0.3
Unreacted H_2 x 10^6 SCF PSD	1.40
C_1-C_3 Gases, Lbs PSD	16,185
C_4-290°F, BPSD	4,937
290-550°F, BPSD	23,685

INSPECTIONS ON FEED AND PRODUCT

	180-550°F FEED	290-550°F PRODUCT
API Gravity @ 60°F	40.6	41.6
Aromatics, Vol.%	25.0	15.0
Olefins, Vol.%	3.2	1.4
Total Nitrogen, ppm	3480	8.0
Sulfur, ppm	80	2.0

TABLE 8

MAXIMUM JP-4 - HCl TREATING FOR REMOVING NITROGEN FROM HYDROTREATED SHALE OIL (5000 PPM)

CASE	PHASE I	PHASE II
Hydrotreated Feed		
TBP Boiling Range, °F	450-1000	490-1000
API Gravity @ 60°F	31.9	28.9
Total Nitrogen, ppm	2000	4800
Sulfur, ppm	55	140
Aromatics & Polars, wt.%	26	42
Reactor Conditions (HCl Treatment)		
Residence Time, Minutes	∿10	30
Inlet Temp., °F	100	100
Outlet Temp., °F	105	110
Total Pressure, psig	∿1	1
HCl Addition, lbs/100 lbs. Feed	0.98	2.68
Settling Time, Minutes	∿15	30
Raffinate Data		
Yield, wt.% Oil Charged	97.0	86.2
API Gravity @ 60°F	32.5	30.7
Total Nitrogen, ppm	720	700
Sulfur, ppm	30	17
Aromatics & Polars, wt.%	24.0	34
Chloride. ppm	200	700
HCl Adduct Decomposition Conditions		
Residence Time, Minutes	∿5	30
Temperature, °F	575	575
Total Pressure, psig	1	1
HCl Recovery, wt.% Reacted	96.2	95.93
Decomposed Adduct (HCl-Free Basis)		
Yield, wt.% Oil Charged	3.0	13.9
API Gravity @ 60°F	12.5	16.8
Total Nitrogen, wt.%	4.34	3.02
Sulfur, ppm	860	900
Aromatics & Polars, wt.%	∿90	89
Chlorine, ppm	5000	1500

TABLE 9

MATERIAL BALANCE SUMMARY OF ANHYDROUS
HYDROGEN CHLORIDE EXTRACTION UNITS - JP-4 OPERATION

BASIS: 490-1000°F Hydrotreat Gas Oil Feed

	FEED	RECOVERED HCl-FREE RAFFINATE	RECOVERED EXTRACT
Yields			
Wt.%	100	86.2	13.9
Vol.%	100	87.1	12.8
Bbl/SD	73,133	63,681	9,388
Inspections & Analyses			
API/Sp. Grav. @ 60°F	28.9/0.8823	30.7/0.8725	18.4/0.9542
Aromatics, wt.%	42	34	89
Total Nitrogen, ppm	4,800	700	3.02 wt.%
Sulfur, ppm	140	17	900
Chlorine, ppm	0	700	1,500

LOSSES: Raffinate = 64 BPSD
Anhydrous HCl = 24,690 lbs/SD

TABLE 10

MAXIMUM JP-8 - HCl TREATING FOR REMOVING NITROGEN FROM HYDROTREATED SHALE OIL (5000 PPM)

CASE	PHASE I	PHASE II
Hydrotreated Feed		
TBP Boiling Range, °F	535-1000	550-1000
API Gravity @ 60°F	31.0	28.5
Total Nitrogen, ppm	2210	5600
Sulfur, ppm	60	150
Aromatics & Polars, wt.%	26.7	45
Reactor Conditions (HCl Treatment)		
Residence Time, Minutes	∿10	30
Inlet Temp., °F	100	100
Outlet Temp., °F	105	110
Total Pressure, psig	∿1	1
HCl Addition, lbs/100 lbs. Feed	1.08	2.70
Settling Time, Minutes	∿15	30
Raffinate Data		
Yield, wt.% Oil Charged	96.6	86.2
API Gravity @ 60°F	31.7	30.3
Total Nitrogen, ppm	770	750
Sulfur, ppm	40	28
Aromatics & Polars, wt.%	24.5	35
Chlorine, ppm	220	750
HCl Adduct Decomposition Conditions		
Residence Time, Minutes	∿5	30
Temperature, °F	575	575
Total Pressure, PSIG	1	1
HCl Recovery, wt.% Reacted	96.55	95.63
Decomposed Adduct (HCl-Free Basis)		
Yield, wt.% Oil Charged	3.4	13.9
API Gravity @ 60°F	11.3	16.3
Total Nitrogen, wt.%	4.34	3.02
Sulfur, ppm	630	930
Aromatics & Polars, wt.%	∿90	89
Chlorine, ppm	5000	1800

TABLE 11

MATERIAL BALANCE SUMMARY OF ANHYDROUS
HYDROGEN CHLORIDE EXTRACTION UNITS - JP-8 OPERATION

BASIS: 550-1000°F Hydrotreat Gas Oil Feed

	FEED	RECOVERED HCl-FREE RAFFINATE	RECOVERED EXTRACT
Yields			
Wt.%	100	86.2	13.9
Vol.%	100	87.1	12.8
Bbl/SD	62,279	60,329	8,890
Inspections & Analyses			
API/Sp. Grav. @ 60°F	28.6/0.8842	30.3/0.8744	16.3/0.9573
Aromatics, wt.%	45	35	89
Total Nitrogen, ppm	5,600	750	3.57 wt.%
Sulfur, ppm	150	28	930
Chlorine, ppm	0	750	1,800

LOSSES: Raffinate = 60 BPSD
Anhydrous HCl = 25,340 lbs/SD

TABLE 12

MAXIMUM JP-4 OPERATING CONDITIONS FOR GAS OIL HYDROCRACKER

BASIS	PHASE I R-1	PHASE I R-2	PHASE II R-1	PHASE II R-2
Operating Factor	0.91		0.90	
Catalyst	Proprietary	Proprietary	NiMo	"B"
Catalyst Life, Years	3	6	2	2
REACTOR OPERATING CONDITIONS:				
LHSV, vol/hr/vol	1		1	2
Average Catalyst Temp., °F	690		710	710
Total Pressure, psig	1600		1700	
Recycle Gas Rate, SCF/B	6000		6000	
Hydrogen Consumption, SCF/B				
Chemical	1250		1450	
Dissolved	105		115	
Total	1355		1565	
Conversion, vol.% Fresh Feed	42		70	
FEEDSTOCK CHARACTERIZATION:				
TBP Boiling Range, °F	450-1000		490-1000	
API Gravity	32.5		30.7	
Total Nitrogen, ppm	720		700	
PRODUCTS, VOL.% FRESH FEED:				
C_4+ Yield	115.0		114.1	
JP-4	115.0		114.1	

TABLE 13

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER
JP-4 OPERATION

BASIS: 63,681 BPSD of HCl Raffinate

99.66 x 10^6 SCF Hydrogen PSD (92.34 x 10^6 SCF H_2 PSD Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	10.3
Ammonia, STSD	5.0
Hydrogen Sulfide, Sulfur Eq. STSD	0.6
Unreacted H_2 x 10^6 SCF PSD	7.32
C_1-C_3 Gases, lbs. PSD	408,529
C_4-180°F, BPSD	10,250
180-490°F, BPSD	64,420

INSPECTIONS ON FEED AND PRODUCT

	FEED	JP-4 PRODUCT
API Gravity @ 60°F	30.7	54.7
Aromatics, %	34 wt.	15 vol.
Olefins, vol.%	-	1.4
Total Nitrogen, ppm	700	1
Sulfur, ppm	17	1

TABLE 14

MAXIMUM JP-8 OPERATING CONDITIONS
FOR GAS OIL HYDROCRACKER

BASIS	PHASE I		PHASE II	
	R-1	R-2	R-1	R-2
Operating Factor	0.91		0.90	
Catalyst	Proprietary	Proprietary	NiMo	"B"
Catalyst Life, Years	3	6	2	2
REACTOR OPERATING CONDITIONS:				
LHSV, vol/hr/vol	1		1	2
Average Catalyst Temp., °F	690		710	740
Total Pressure, psig	1600		1700	
Recycle Gas Rate, SCF/B	6000		6000	
Hydrogen Consumption, SCF/B				
Chemical	1250		1450	
Dissolved	105		125	
Total	1355		1575	
Conversion, vol.% Fresh Feed	42		70	
FEEDSTOCK CHARACTERIZATION:				
TBP Boiling Range, °F	535-1000		550-1000	
API Gravity	32.5		30.3	
Total Nitrogen, ppm	720		750	
PRODUCTS, VOL.% FRESH FEED:				
C_4+ Yield	114.6		111.3	
C_4-290°F	45.9		49.1	
JP-8 (290-550°F B.R.)	68.7		62.2	

TABLE 15

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER
JP-8 OPERATION

BASIS: 60,329 BPSD of HCl Raffinate

92.02 x 10^6 SCF Hydrogen PSD (87.48 x 10^6 SCF H_2 PSD Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	10.4
Ammonia, STSD	5.1
Hydrogen Sulfide, Sulfur Eq. STSD	0.7
Unreacted H_2 x 10^6 SCF PSD	7.54
C_1-C_3 Gases, lbs. PSD	661,589
C_4-290°F, BPSD	29,555
290-550°F, BPSD	37,573

INSPECTIONS ON FEED AND PRODUCT	FEED	JP-8 PRODUCT
API Gravity @ 60°F	30.3	43.0
Aromatics, %	35 wt.	16 vol.
Olefins, vol.%	-	1.6
Total Nitrogen, ppm	750	1
Sulfur, ppm	28	1

TABLE 16

JP-4 AND OTHER FUELS - OPERATING CONDITIONS FOR GAS OIL HYDROCRACKER

	PHASE II		
BASIS:	R-1		R-2
Operating Factor		0.90	
Catalyst	NiMo		"B"
Catalyst Life, Years	2		2
REACTOR OPERATING CONDITIONS:			
LHSV, vol/hr/v	1		2
Average Catalyst Temp., °F	710		710
Total Pressure, psig		1700	
Recycle Gas Rate, SCF/B		6000	
Hydrogen Consumption, SCF/B			
Chemical		1150	
Dissolved		105	
Total		1255	
Conversion, vol.% Fresh Feed		70	
FEEDSTOCK CHARACTERIZATION			
TBP Boiling Range, °F		490-1000	
API Gravity		30.7	
Total Nitrogen, ppm		700	
PRODUCTS, VOL.% FRESH FEED			
C_4+ Yield		109.8	
JP-4		76.1	
DF-2 (490-675°F B.R.)		25.0	
Fuel Oil (675°F+ Bottoms)		8.7	

TABLE 17

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER
JP-4 PLUS OTHER FUELS

BASIS: 63,681 BPSD of HCl Raffinate

79.92×10^6 SCF Hydrogen PSD (73.23×10^6 SCF H_2 PSD Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	10.3
Ammonia, STSD	5.0
Hydrogen Sulfide, Sulfur Eq. STSD	0.6
Unreacted $H_2 \times 10^6$ SCF PSD	6.69
C_1-C_3 Gases, lbs. PSD	312,950
C_4-180°F, BPSD	6,725
180-490°F, BPSD	41,711
490-675°F, BPSD	15,920
675°F+ Bottoms, BPSD	5,540

INSPECTIONS ON FEED AND PRODUCT

	FEED	JP-4	DF-2	675°F+ BOTTOMS
API Gravity @ 60°F	30.7	54.7	36.0	33.0
Aromatics, %	34 wt.	15 vol.	23 wt.	30 wt.
Olefins, vol.%	--	1.4	--	--
Total Nitrogen, ppm	700	1	1	3
Sulfur, ppm	17	1	1	2

TABLE 18

PRODUCT INSPECTIONS AND ANALYSES

CHEMICAL AND PHYSICAL TEST DATA	JP-4	JP-8	I-290°F	DF-2	BLENDED HEAVY FUEL
API @ 60°F	50.9	42.9	71.0	38.0	25.7
Distillation, °F	D-2887	D-2887	D-86	D-86	---
I	158	210	50	450	---
10 vol.%	---	310	145	480	---
20 vol.%	260	352	170	495	---
50 vol.%	342	410	202	535	---
90 vol.%	459	510	240	590	---
EP	527	560	290	650	---
Aromatics, %	15	16	3	24	36
Olefins, vol.%	1	2	1	---	---
Mercaptans, wt.%	0.0001	0.0003	---	---	---
Sulfur, wt.%	0.0003	0.0002	NA	0.0011	0.06
Nitrogen (Total), ppm	5	3	1	3	3600
Flash Pt., °F	---	100	---	210	---
Freeze Pt., °F	-72	-70	-76	-10	---
Net Ht. of Comb., BTU/lb	18,764	18,610	19,050	18,730	18,500
H_2 Content, wt.%	14.16	13.85	NA	NA	NA

TABLE 19

MATERIAL BALANCE SUMMARY

BASIS: 100,000 BPSD Raw Occidental Shale Oil Refinery

CASE	MAX. JP-4	MAX. JP-8	JP-4 PLUS OTHER FUELS
Net Products, BPSD (TBP Cuts)			
C_4-290°F B.R. Naphtha	---	36,608	---
C_4-490°F B.R. JP-4	101,337	---	75,094
290-550°F B.R. JP-8	---	61,258	---
490-675°F B.R. DF-2	---	---	15,920
675-1000°F B.R. Heavy Fuel	---	---	5,540
1000°F+ Bottoms Heavy Fuel	4,159	4,159	4,159
TOTAL FUELS	105,496	102,025	100,713
Other Products, STSD			
Liquid Ammonia	203	207	203
Sulfur	111	111	111
Ammonium Chloride	14	14	14
Liquid Fuel Yields			
Total Products as vol.% Process Feeds	105.1	102.0	100.7
Naphtha	---	36.5	---
JP-4	101.0	---	75.1
JP-8	---	61.3	---
DF-2	---	---	15.9
Heavy Fuel	4.1	4.2	9.7
TOTAL REFINERY INPUT (CRUDE, FUEL & UTILITIES CONVERTED TO FOE), BPSD	116,776	116,105	115,143
TOTAL PRODUCTS AS VOL. % REFINERY INPUT	90.3	87.9	87.5
Naphtha	---	31.5	---
JP-4	86.8	---	65.3
JP-8	---	52.8	---
DF-2	---	---	13.8
Heavy Fuel	3.5	3.6	8.4
Overall Refinery Thermal Energy Efficiency, %	76	75	76

TABLE 20

INSPECTIONS AND ANALYSES OF LABORATORY PRODUCTION SAMPLES OF JP-4 FROM RAW OCCIDENTAL SHALE OIL

	SPEC.	CONVENTIONAL	LOW AROM.	HIGH AROM.
COMPOSITION, VOL.%				
Hydrotreated Naphtha	---	15	0	60
Hydrocrackate	---	85	100	40
CHEMICAL & PHYSICAL TEST DATA				
Aromatics, vol.% max.	25.0	14.6	4.4	17.0
Olefins, vol.% max.	5.0	1.3	1.4	2.6
Mercaptans, wt.% max.	0.001	0.0001*	< 0.0001*	0.0001*
Sulfur, wt.% max.	0.40	0.0003	0.0003	0.0005
Nitrogen (Total), ppm	---	5	< 1	27
Simulated Distillation, °F (ASTM D-2887)				
IBP	Report	158	100	105
20 vol.% (max. temp.)	266	260	213	216
50 vol.% (max. temp.)	365	342	304	356
90 vol.% (max. temp.)	482	459	440	479
EP (max. temp.)	608	527	540	598
Density @ 15°C, min/max.	0.751/0.802	0.776	0.753	0.779
RVP (100°F) psi min/max.	2.0/3.0	1.6*	2.2*	2.6*
Freeze Pt., °F, max.	-72	-72	-89	-72
Net Ht. of Comb., min.	18,400	18,674*	18,824*	18,706*
H_2 Content, wt.% min.	13.6	14.16*	14.80*	14.24*
Copper Strip, Corrosion, max.	1B	1-	1-	1-

* Data from Fuels Lab, WPAFB

TABLE 21

INSPECTIONS AND ANALYSES OF LABORATORY PRODUCTION SAMPLES OF JP-8 FROM RAW OCCIDENTAL SHALE OIL

	SPEC.	100% HYDRO. NAPHTHA	100% HYDROCRACKATE
CHEMICAL AND PHYSICAL TEST DATA			
Aromatics, vol.% max.	25.0	13.2	17.4
Olefins, vol.% max.	5.0	1.4	2.3
Mercaptans, wt.% max.	0.001	0.0001*	0.0003*
Sulfur, wt.% max.	0.4	0.0006	0.0003
Nitrogen (Total), ppm	NA	7.3	1.4
Simulated Distillation, °F (ASTM D-2887)			
IBP	Report	210	187
10 vol.% (max. temp.)	367	310	271
20 vol.% (max. temp.)	Report	352	313
50 vol.% (max. temp.)	Report	403	416
90 vol.% (max. temp.)	Report	464	534
EP	626	516	(586)
Flash, °F, min.	100	100	75**
Density @ 15°C, min.	0.775	0.796	0.814
max.	0.840		
Freeze Pt., °F	-58	-62	<-76
Net Ht. of Comb., BTU/lb.	18,400	18,639*	18,588*
H_2 Content, wt.% min.	13.6	13.98*	13.75
Copper Strip Corrosion, max.	1B	1-	1-

* Data from Fuels Lab, WPAFB

** Does not meet Spec. due to low IBP

TABLE 22

BASIS FOR DEVELOPING PAHSE II
PRELIMINARY ECONOMIC EVALUATION

CAPITAL INVESTMENT:

- Midwest Plant Location Adjacent to Existing Refinery
- 100,000 BPSD Crude Shale Oil Capacity
- 1st Quarter 1980 Cost Base
- 100% Equity Financing
- Investment Timing over Three-Year Construction Period
  - 25% First Year
  - 50% Second Year
  - 25% Third Yeat
- 10% Investment Tax Credit

WORKING CAPITAL:

- 21 Days Crude Storage Capacity/14 Day Crude Inventory
- 14 Days Product Storage Capacity/7 Day Product Inventory
- Crude Valued at $30.00/Bbl
- Product Valued at $40.00/Bbl
- Debt Financed at 10%

CAPITAL RETURN:

- 15% DCF Rate
- Zero Salvage Value
- 13 Years Sum of Years Digets Depreciation

OPERATING BASES:

- 16 Year Plant Operating Life
- 50% Operating Capacity 1st Year - 100% Thereafter
- 90% On-Stream Factor

OPERATING COST:

- Crude Shale Oil - $30/Bbl
- All Process Heat Requirements are Generated Internally
- Cooling Water 3¢/1000 gal.
- Electricity 3.5¢/KwH
- Operators* $12.00/Manhour
- Helpers* $10.50/Manhour   Wtd. Avg. @ $10.95
- Supervision 25% of Direct Labor
- Overhead 100% of Direct Labor
- Federal & State Taxes - 50%
- Maintenance, Local Taxes & Insurance - 4.5% of Fixed Investment
- Product Values - All Fuels are Equal Value
- By-Product Values - Ammonia ($120/ST)
  Sulfur ($ 53/LT)

* 4.2 Shift Positions plus 10% Relief Required for Continuous Operation

TABLE 23

PLANT CAPACITIES AND ESTIMATED FIRST QUARTER 1980 INVESTMENTS (PHASE II)

	MAX. JP-4		MAX. JP-8		JP-4 + OTHER FUELS	
	CAPACITY PSD	$ X 10^6	CAPACITY PSD	$ X 10^6	CAPACITY PSD	$ X 10^6
H_2 Plant (TPO), MMSCF/D (100% H_2 Basis)	151	107.4	134	100.0	145	105.0
H_2 Plant (Steam Reforming), MMSCF/SD	94	33.9	114	34.5	80	30.8
Sulfur Recovery, STSD	111	10.0	111	10.0	111	10.0
Waste Water Treating, STSD NH_3	206	12.9	209	12.9	206	12.9
Main Hydrotreater & H_2S Recovery, MBPSD	100	166.3	100	166.3	100	166.3
Atm. & Vac. Distn. MBPSD	104	45.0	104	45.0	104	45.0
Dist. Hydrotreater, MBPSD	25	31.5	28	34.4	25	31.5
HCl Treater, MBPSD	74	2.4	61	2.3	74	2.4
Hydrocracker & Atm. Distn., MBPSD Fresh Feed	64	89.5	61	87.3	64	86.1
SUB TOTAL		498.9		492.7		490.0
Tankage, MM Bbls.	5.1	40.0	5.0	39.0	5.0	39.0
TOTAL ON-SITES		538.9		531.7		529.0
Off-Sites (45% On-Sites Less Tankage)		224.5		221.7		220.0
TOTAL CAPITAL INVESTMENT		763.4		753.4		749.0

TABLE 24

PHASE II PRELIMINARY COST COMPARISON FOR MANUFACTURING MILITARY FUELS FROM RAW OCCIDENTAL SHALE OIL

BASIS: 100,000 BPSD Refinery Crude Capacity (90,000 BPCD)

	MAX. JP-4	MAX. JP-8	JP-4 AND OTHER FUELS
TOTAL PLANT INVESTMENT, \$ x 10^6			
Plant	763.4	753.4	749.0
Catalysts	17.5	17.3	15.3
Working Capital	78.3	77.1	76.3
TOTAL	859.2	847.8	840.6
MANUFACTURING COSTS - \$/CD			
Direct Labor	15,538	15,538	15,538
Purchased Power and Cooling Water	70,454	68,454	60,090
Catalyst, Chemicals & Royalties	42,775	42,873	40,035
Overhead @ 100% Direct Labor	15,538	15,538	15,538
Maint., Local Taxes & Insurance	66,440	65,552	65,219
Sub Total	210,775	207,955	196,420
Less NH_3 & S (Credit)	29,612	30,092	27,612
Direct Costs	181,163	177,863	168,808
Per Bbl Liquid Product	\$1.91	\$1.94	\$1.86
TOTAL LIQUID FUELS, BPCD	94,946	91,823	90,641
TOTAL MANUFACTURING COSTS, \$/Bbl Product*	\$8.97	\$9.03	\$9.17
Adjusted Crude Cost, \$/Bbl Product	\$32.86	\$33.80	\$34.00
TOTAL PRODUCT COST			
\$/Bbl	\$41.83	\$42.83	\$43.17
¢/Gal	100	102	103

* Total Manufacturing Costs Computed on the Basis Shown in Table 22 for Developing Phase II Preliminary Economics

TABLE 25

SUMMARY

PROCESSING ROUTE FOR	MAX. JP-4	MAX. JP-8	JP-4 AND OTHER FUELS
PRODUCTS AS VOL. % CRUDE PROCESSED			
Jet Fuel	101.0	61.3	75.1
Total Liquid Products	105.1	102.0	100.7
PRODUCTS AS VOL. % TOTAL ENERGY INPUT (CRUDE + FUEL + UTILITIES CONVERTED TO FOE)			
Jet Fuel	86.8	52.8	65.3
Total Liquid Products	90.3	87.9	87.5
Total Product Cost, \$/Bbl	41.83	42.83	43.17
¢/Gal	100	102	103
Overall Thermal Efficiency, %	76	75	76
Plant Investment, \$/SDB	7809	7707	7643