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ADVANCED ENERGY RESEARCH CORPORATION

REFINING AND UPGRADING OF SYNFUELS FROM COAL AND OIL SHALES  
BY ADVANCED CATALYTIC PROCESSES

First Interim Report, Processing of Paraho Shale Oil

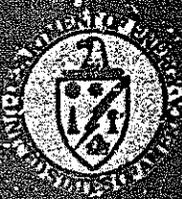
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MASTER

April 1978  
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Chevron Research Company  
Richmond, California



U. S. DEPARTMENT OF ENERGY

REFINING AND UPGRADING OF  
SYNFUELS FROM COAL AND OIL SHALES  
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First Interim Report  
Processing of Paraho Shale Oil

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DEPARTMENT OF ENERGY

PROCESSING OF PARAH0 SHALE OIL  
DOE CONTRACT EF-76-C-01-2325

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PROCESSING OF PARAHIO SHALE OIL  
DOE CONTRACT EF-76-C-01-2315

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## A B S T R A C T

Advanced commercial petroleum processing technology was employed in pilot plant facilities to produce transportation fuels (gasoline, kerosene, jet fuel, and diesel) from Paraho shale oil. Three likely refining routes were identified and demonstrated sufficiently to permit preparation of screening-type process designs and cost estimates: (1) hydrotreating followed by hydrocracking, (2) hydrotreating followed by catalytic cracking, and (3) coking followed by hydrotreating. In a fourth feasible alternative, raw shale oil was hydrotreated to produce a synthetic crude suitable for processing in an existing refinery.

The key to successful shale oil refining is an effective initial hydrotreating step. This removes contaminants and permits the use of conventional hydrocracking or fluid catalytic cracking. Distillates from coking of raw shale oil also require subsequent hydrotreating to remove residual impurities and meet final product specifications.

Refining costs to convert raw shale oil to transportation fuels were estimated to be about \$8-\$10/barrel, depending upon the processing route and based on producing 100,000 BPCD of transportation fuels from a "grass roots" refinery located near a Rocky Mountain of Mid-Continent urban center. Costs are higher for smaller, remotely located refineries. Hydroprocessing of raw shale oil to produce a synthetic crude would cost about \$6.50/barrel.

## EXECUTIVE SUMMARY

A Colorado shale oil produced in a surface retort at the Paraho test facility was refined in pilot plant facilities at Chevron Research Company. Advanced commercial petroleum processing technology was employed to produce transportation fuels (primarily gasoline, kerosene jet fuel, and diesel). Three likely refining routes were identified and demonstrated sufficiently to permit preparation of screening-type process designs and cost estimates. These routes are: (1) hydrotreating followed by hydrocracking, (2) hydrotreating followed by catalytic cracking, and (3) coking followed by hydrotreating. In addition, a fourth alternative was evaluated where raw shale oil was hydrotreated to produce a synthetic crude suitable for processing in an existing refinery.

Shale oil presents unusual refining problems. It is notable for high nitrogen content and for large amounts of unsaturated and metallic contaminants. These prevent its being mixed with crude oils for normal refinery processing.

The key to successful shale oil refining is an effective initial hydrotreating step. This removes the contaminants and permits the use of conventional hydrocracking or fluid catalytic cracking. Distillates from coking of raw shale oil also require subsequent hydrotreating to remove residual impurities and meet final product specifications.

Refining costs were estimated for each of the three process routes based on producing (1) 50,000 BPCD of transportation fuels from a "grass roots" refinery in a remote Rocky Mountain location, and (2) 100,000 BPCD of transportation fuels from a "grass roots" refinery located near a Rocky Mountain or Mid-Continent urban center. The costs for hydroprocessing to produce a synthetic crude also were estimated both for a "grass roots" site near the shale oil retorting facilities and for location at a typical existing Mid-Continent refinery. All refineries were designed with facilities to meet present environmental requirements and anticipated near-term environmental and energy conservation regulations.

Refining costs to convert raw shale oil to transportation fuels were estimated to be about \$8-10/barrel for the 100,000 BPCD refineries, depending upon the processing route. Costs are higher for the smaller, remotely located refineries.

It was estimated that hydroprocessing of raw shale oil to produce a synthetic crude would cost about \$6.50/barrel at either of the sites considered.

The hydrocracking and fluid catalytic cracking cases show a much higher yield of light products (91-95 vol %) from raw shale oil than from the coking case (78 vol %), which is the lowest cost plan. The fluid

catalytic cracking case has a slight advantage in cost and yield over hydrocracking, but the latter provides more flexibility for varying individual product yields and is the only scheme capable of efficiently maximizing jet fuel production.

The total processing costs given above include capital charges based on a discounted cash flow (DCF) rate of return on investment of 15%. In the report, total processing costs are presented covering DCF rates of return in the range of 10% to 20%. For evaluating new projects, a 15% DCF rate of return is often used in industry as a checkpoint below which the justification for investment becomes increasingly questionable. Taking into account that the production of oil from shale is an unproven venture with significant internal and external uncertainties and risks, the appropriate rate of return to be used in economic analysis of this new technology should be at least this high.

In order to obtain the total cost of producing finished light products from shale, it would be necessary to add to the above processing costs, the costs for resource acquisition, shale mining, handling, retorting, and waste disposal, and the transportation of raw shale oil. In the case of the hydroprocessed synthetic crude, the downstream refining costs to produce finished products also need to be included.

It is concluded that a Paraho shale oil can be refined to high quality transportation fuels using advanced commercial petroleum processing technology. Reasonable solutions appear to be at hand to accommodate the special problems posed by high contaminant levels in the crude shale oil. Technical areas were identified where future study would be desirable. None of these appears to present a major problem. However, further pilot plant and process design studies will be warranted to support evaluation of specific commercial projects.

## TECHNICAL SUMMARY

Chevron Research Company conducted extensive laboratory, pilot plant, and engineering studies to demonstrate the feasibility and estimate costs for refining of surface-retorted Paraho shale oil. These studies showed that raw shale oil can be conventionally refined using advanced, commercialized petroleum technology to produce finished transportation fuels, primarily motor gasoline, kerosene jet fuel, and diesel fuel.

Three alternative refining routes were evaluated: (1) hydrotreating followed by hydrocracking, (2) hydrotreating followed by fluid catalytic cracking, and (3) coking followed by hydrotreating. In addition, a fourth alternative was evaluated where raw shale oil was hydrotreated to produce a synthetic crude suitable for processing in an existing refinery.

Shale oil is sufficiently different from most petroleum crudes that refining presents unusual problems. The shale oil used in this study contained over 2 wt % nitrogen compared to less than 0.3% for most petroleum crudes. It also contained sulfur, arsenic, and oxygenated, unsaturated, and metallic contaminants or impurities. These prevent its being mixed with crude oils for normal refinery processing.

The key to successful refining of whole shale oil is an effective initial hydrotreating process. This removes the contaminants and permits subsequent conversion to proceed through established hydrocracking or fluid catalytic cracking units. Distillates from coking of raw shale oil also require subsequent hydrotreating to remove residual impurities and meet final product specifications.

Pilot plant tests showed that the nitrogen in shale oil can be reduced to levels as low as 1 ppm in a single processing step. However, evaluation of the results indicated that it is adequate and economically advantageous to hydrotreat to an intermediate nitrogen level of about 500-1500 ppm in the whole liquid product. This product can be refined to high quality transportation fuels using conventional refining technology.

The key hydrotreating step was demonstrated in a 3500-hour pilot plant test in which whole shale oil was hydroprocessed to about 500 ppm product nitrogen over Chevron ICR 106 catalyst. Products from the hydrotreated shale oil are primarily (1) a 650°F+ gas oil that is similar in most respects to normal gas oil feeds to hydrocrackers and fluid catalytic cracking units, (2) a middle distillate fraction that can be converted to salable diesel and jet fuel by additional hydrotreatment, and (3) naphtha that is a suitable feed to a two-stage catalytic reformer.

If hydrocracking is the major conversion process after hydrotreating, pilot plant studies show that the gas oil fraction can be converted to a stable, low nitrogen, high smoke point jet fuel plus additional naphtha for reforming to high octane gasoline. Alternatively,

hydrocracking conditions could be adjusted to produce primarily either gasoline or diesel fuel. If fluid catalytic cracking is the major conversion process, the main products are high octane gasoline and light cycle oil that can be further hydrotreated to diesel fuel. In the coking alternative, raw shale oil is fed directly to the coker; and the distillate product is hydrotreated to diesel fuel and a naphtha suitable for catalytic reforming to motor gasoline.

The overall yields of finished transportation fuels for the cases studied ranged between 78 vol % of shale oil feed for the coking case to 91-95 vol % for the hydrocracking and fluid catalytic cracking cases, respectively. These yields represent the net amount of light products after subtracting refinery fuel and hydrogen plant feed, which were internally supplied. The net yields of transportation fuels for each case are as follows:

Major Conversion Process	Hydrocracking	Fluid Catalytic Cracking	Coking
<u>Product Type, LV % Net Yield from Dewatered Raw Shale Oil Feed</u>			
Propane-Butane (LPG)	0	0	2
Motor Gasoline	17	38	17
Kerosene Jet Fuel	20	0	0
Diesel Fuel	<u>54</u>	<u>57</u>	<u>59</u>
Total Transportation Fuel	91	95	78

Product yields can be varied depending on demand. For example, in the hydrocracking case, the quantity of kerosene jet fuel could be increased to perhaps 50-60% of the total product plus additional naphtha. If military jet fuel such as JP-4 is also a target product, it would include much of the naphtha, giving an even higher yield of total jet fuel. Additional pilot plant and engineering studies would be required to more clearly define the processing conditions and costs for these cases.

Screening-type process designs and estimated refining costs were developed for each of the three process routes listed above based on producing (1) 50,000 BPCD of transportation products from a "grass roots" refinery in a remote Rocky Mountain location and (2) 100,000 BPCD of transportation products from a "grass roots" refinery located near a Rocky Mountain or Mid-Continent urban center. The costs for hydroprocessing to produce a synthetic crude were estimated both for a "grass roots" site near the shale oil retorting facilities and for

location at a typical existing Mid-Continent refinery. All refineries were designed with facilities to meet present environmental requirements and anticipated near-term environmental and energy conservation regulations.

Results of the cost studies showed that refining of crude Parano shale oil to finished transportation fuels in new "grass roots" facilities would cost about \$8-\$10 per barrel for 100,000 BPCD and about \$10-\$12 per barrel for 50,000 BPCD of light products at today's (First Quarter 1978) costs. Refining to a synthetic crude would cost about \$6.50 per barrel, not including costs for downstream processing. The synthetic crude would be suitable for further upgrading into finished products in a large percentage of existing refineries.

As outlined above, the hydrocracking and fluid catalytic cracking cases show a much higher yield of light products from raw shale oil than the coking case, which is the lowest cost plan. The fluid catalytic cracking case has a slight advantage in cost and yield over hydrocracking, but the latter provides greater flexibility for varying individual product yields. Hydrocracking is the only scheme capable of efficiently maximizing jet fuel production.

The total processing costs given above include capital charges based on a 15% discounted cash flow (DCF) rate of return on investment. In the report, total processing costs are presented covering DCF rates of return in the range of 10% to 20%. For evaluating new projects, a 15% DCF rate of return is often used in industry as a checkpoint below which the justification for investment becomes increasingly questionable. Taking into account that the production of oil from shale is an unproven venture with significant internal and external uncertainties and risks, the appropriate rate of return to be used in economic analysis of this new technology should be at least this high.

In order to obtain the total cost of producing finished light products from shale, it would be necessary to add to the above processing costs, the cost of resource acquisition, shale mining, handling, retorting, and waste disposal, and the transportation of raw shale oil. In the case of hydroprocessed synthetic crude, the downstream refining costs to produce finished products also need to be included.

Areas for future study also were identified. None of these appears to represent a major problem. Rather, most involve second-order process refinements that will be incorporated in future commercial projects which are based on specific product slates.

## CONTRACT OBJECTIVES AND SCOPE OF WORK

The objective of the program is to determine the feasibility and estimate the costs of hydroprocessing four synthetic crude feedstocks to distillate fuels, including high octane gasoline, using presently available technology. To date, two feedstocks have been studied:

- (1) Paraho shale oil, prepared in the indirectly heated mode and
- (2) solvent refined coal (SRC) of the type produced at the DOE pilot plant in Tacoma, Washington. This report summarizes the pilot plant work with the first feedstock, Paraho shale oil.

The feasibilities of hydroprocessing each of the synthetic liquids are to be compared through catalyst tests and evaluations, whereby commercial plant yields, hydrogen consumption, product distribution, and product inspection will be estimated. The necessary tests and evaluations for each feedstock are to be done to support "process comparison"-type estimates for each of the major refining steps. The results of the contract, insofar as hydroprocessing is concerned, will be obtained with the proprietary commercial Chevron catalysts identified in the contract.

Catalyst activity and stability information for each feedstock are to be obtained as needed to define commercial operating conditions. These data will provide the basis for the overall refining plan, plant cost estimates, utility and hydrogen requirements, etc. If tests show that refining a particular feedstock using presently existing information is not feasible, it is not intended under this program to conduct any research or development work to solve the problems encountered.

Tests are to be conducted only to the extent needed to enable making reasonable estimates of commercial plant performance and only to the extent a commercial plant is feasible using presently existing technology, subject to the mutual agreement of DOE and Chevron Research. Tests are to be made for each whole synthetic oil and, where appropriate, for the fractions derived therefrom. Tests will not be carried out for processes which can be reliably estimated.

Part I  
Laboratory and Pilot Plant Studies

By

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DOE Contract EF-76-C-01-2315

## 1. Summary of Shale Oil Pilot Plant Studies

### A. Schedules

According to the original timing estimate for an individual feedstock, the preliminary feed analyses and pilot plant program for each feedstock will require one year, followed by an additional 14 weeks for completion of the product analyses, pilot plant report, and final process design.

A general schedule was set up for each feedstock to serve as a guideline for work. It is recognized, however, that there is an exploratory nature to processing these feedstocks and that it may be desirable to alter the schedule as information on any single feedstock becomes available, either to emphasize or deemphasize a certain part of the program.

Figure 1 shows how the general schedule applied to our work on Paraho shale oil.

With the permission of the DOE technical representative, the program was modified somewhat from the original proposal, which was to evaluate only one basic refining scheme for each feedstock. This first refinery plan involved hydrotreating the whole shale oil followed by hydrocracking as a major conversion step. As the program progressed, it became obvious that two other processing schemes could also be attractive: (1) hydrotreating the whole shale oil followed by FCC as a major conversion step and (2) delayed coking of the shale oil followed by hydrotreating of the products. Sufficient pilot plant work was included so that these alternate processing schemes could also be evaluated. As a result of the work described in this report, we are able to propose three different technically feasible schemes for refining shale oil. Choice of the most appropriate scheme will depend on the desired products and particular refinery situation. (No additional funding was required; however, the program took two to three months longer than originally estimated as a result of the modifications.)

An eighth task was also added. Professor Victor Yesavage of the Colorado School of Mines asked that several shale oil distillate fractions be hydrotreated and sent to him for his research studies. This work was done under Contract Modification No. A004, and additional funding was provided for the task.

### B. Summary of Experimental Studies

The laboratory and pilot plant studies described in this report demonstrate that three technically feasible processing routes for processing shale oil are (1) hydrotreating followed by hydrocracking, (2) hydrotreating followed by FCC, and (3) severe coking followed by hydrotreating. Schematic flow diagrams for the three refineries proposed as a

result of this program are shown in Figures 2, 3, and 4. Figure 5 is a schematic diagram showing the processing steps applied in our pilot plant program classified according to task number as summarized below.

Task 1 - Feed Analysis  
and Feed Preparation

The shale oil was dewatered by heating to 170°F and allowing it to stand. The water layer separated, and most of the fines were removed with the water. Inspections of the dewatered feed are presented.

Task 2 - Whole Oil Hydrotreating  
(First-Stage Hydrotreating)

If the whole shale oil is hydrotreated before distillation, the entire stock is upgraded in this first catalytic step. Whole shale oil can be hydrotreated to nitrogen levels as low as 1 ppm in a single hydrotreating stage. However, evaluation of the results indicated that it is economically advantageous to hydrotreat to about 500-1500 ppm nitrogen in the whole liquid product, then distill and, as necessary, process the various product fractions.

A pilot plant test was undertaken to demonstrate operation at this intermediate nitrogen level. In a 3500-hour pilot plant test using Chevron ICR 106 catalyst, whole Paraho shale oil was hydrotreated to about 500 ppm product nitrogen. Catalyst deactivation rate was very low at processing conditions of 0.6 LHSV and 2000 psia hydrogen pressure. The run was shut down when the feed supply was exhausted; the catalyst was still active; and, clearly, a commercial run length was demonstrated. At 500 ppm product nitrogen, the hydrogen consumption was around 2000 SCF/bbl.

Alumina of a moderate surface area was used as a guard bed to remove arsenic and iron upstream of the catalyst. The guard bed was designed to protect the downstream catalyst from these potential catalyst poisons.

At 500-1000 ppm product nitrogen, the 650°F+ product is similar in most respects to normal gas oil feeds to existing hydrocrackers and FCC units. Some cracking of shale oil occurs during hydrotreating, and the amount of 1000°F+ material is reduced from 8-9% to about 3% of the whole product. This end point reduction is desirable for downstream conversion processes. The sulfur content of the hydrotreated shale oil is very low compared to the petroleum-derived vacuum gas oils (VGO's).

Task 2A - Coking of Whole  
Shale Oil, Hydrofinishing  
of Coker Distillate

As an alternate to the whole oil hydrotreating, coking has been proposed as the first processing step in the conversion of raw shale oil into distillate fuels. It has the advantage of removing any shale

solids suspended in the raw oil, most of the nonfilterable iron, and about 80% of the arsenic. The coker distillate is more easily hydro-treated than raw shale oil, and the hydrogen consumption is lower for product of a given nitrogen level. A disadvantage of coking is that one of the products is coke which is, based on current prices, of relatively low value compared to transportation fuels. Another disadvantage is that the coking refinery is less efficient in producing transportation fuels; about 17-22% more shale oil is required in making the same light products as the other cases described here.

Our results indicate that stable diesel was produced from shale oil coker distillate at relatively high nitrogen levels. It, therefore, may be possible to produce diesel fuel in the first-stage hydrotreater without further processing. It is somewhat unexpected that such high nitrogen content diesel fuel is stable, and coker operating conditions might affect this result. In order to ensure product stability over a typical range of operating conditions, the proposed coking refinery also includes a middle distillate hydrotreater. In addition to providing the insurance of stable diesel fuel product, this hydrotreater allows more refinery flexibility, including the possibility of making jet fuel by more severely hydrotreating a portion of the middle distillate. (No pilot plant work was done to study the production of jet fuel from shale oil coker distillate.)

#### Task 3 - Middle Distillate Hydrotreating

The 350-650°F product after hydrotreating whole shale oil to about 500 ppm nitrogen did not consistently pass the stability test for diesel fuel; therefore, further hydrotreating was required by means of a middle distillate hydrotreater. At 2.0 LHSV, 1200 psig total pressure, it was possible to produce stable diesel with an additional hydrogen input of 100 SCF/bbl or lower. A stable jet fuel can be produced by hydrotreating a kerosene fraction of between about 475°F and 500°F end point to approximately 5 ppm nitrogen or less. (The middle distillate fraction is waxy; and, therefore, higher end point kerosenes will not pass the freezing point specifications for jet fuel.)

#### Task 4 - Recycle Hydrocracking of Hydrotreated Shale Oil

Two hydrotreated VGO's from shale oil were pilot plant tested for extinction recycle hydrocracking using Chevron's Isocracking process in a single stage to produce jet fuel and naphtha. The first feed tested was the entire 650°F+ hydrotreated shale oil; the second was the 650-850°F fraction. High smoke point, low freeze point, nitrogen-free jet fuel is produced in both cases at a recycle cut point of 525-535°F. The hydrotreated shale oil is somewhat harder to process than a Middle Eastern gas oil of the same boiling range. The 850°F end point hydro-treated shale oil feed is easier to hydrocrack than the entire 650°F+ fraction of the hydrotreated shale oil.

Task 4A - Alternate Processing of  
650°F+ Hydrotreated Shale Oil: FCC

In a series of catalytic cracking tests, 650°F+ hydrotreated shale oil of 1500 ppm nitrogen or lower was shown to be an excellent feed to a catalytic cracker. Yields and product properties are similar to those obtained from Middle Eastern feedstocks. At a given nitrogen level, the coke make is lower for the hydrotreated shale oil than for a hydro-fined Arabian gas oil, probably as a result of the large hydrogen input for the shale oil.

Task 5 - Catalytic Reforming

According to the contract, catalytic reforming performance was to be estimated based on correlations unless the feedstock appears sufficiently unique to justify catalyst testing. The hydrotreated naphtha products obtained in this program were judged to be good feedstocks for catalytic reforming using Chevron's Rheniforming process, provided a naphtha hydrotreater is included as a first stage. Therefore, no pilot plant work on catalytic reforming is included.

Task 6 - Process Design Studies

Results of these studies are presented in Part II of this report.

Task 7 - Reports

The 24 reports to date include monthly, quarterly, and annual reports. The present report is first interim report for Paraho shale oil, presenting detailed laboratory and pilot plant results in Part I and process engineering studies and cost estimates in Part II.

Task 8 - Distillate Shale  
Oil Hydrotreating

This task was added at the request of Professor Yesavage. The purpose was to hydrotreat two shale oil distillates to several different nitrogen levels for Professor Yesavage's research studies.

## II. Description of Technical Accomplishments

### Task 1 - Feedstock Analysis and Feed Preparation

The purpose of the first task in this study was to perform preliminary analyses of the Paraho shale oil feed and then to carry out any feed pretreatment or feed preparation steps necessary prior to the first catalytic hydroprocessing steps described in Task 2. Tables I-VI, inclusive, summarize the results of analyses of the shale oil feedstocks used in this study.

The analytical methods used throughout this study are for the most part either standard ASTM methods used throughout the petroleum industry or routine chemical analyses by conventional techniques. However, several analytical methods are used that are proprietary techniques developed at Chevron (for example, the group-type mass spectrometric methods) or Chevron variations on methods used industrywide (such as the diesel thermal stability test described later in this report). The asphaltene tests (which fall into the second category) are referred to as "hot heptane asphaltenes" which determines the material insoluble in a standard amount of normal heptane and "C<sub>5</sub> asphaltenes" which determines the amount of material insoluble in a standard amount of normal pentane.

Three hundred and thirty gallons of Paraho crude shale oil, produced in the indirectly heated mode, were shipped from the Laramie Energy Research Center to Chevron Research Company. These drums arrived on September 10, 1976.

The Paraho shale oil samples, as received, were nitrogen blanketed; all of the experiments described in this report were performed with nitrogen blanketing of the shale oil samples to avoid any contact with air.

The Paraho shale oil contains a considerable amount of water as an emulsion. Representative samples were separated from water by equilibrium flash vaporization (EFV). Results compare to those obtained by standard ASTM Method D 96 (centrifugal separation) as follows:

Separation Method	Water Phase, % of Total Emulsion		Fines, Vol % of Total Emulsion
	Wt %	Vol %	
EFV	6.4	5.9	0.7
EFV	5.3	5.0	No Analysis
D 96	6.0	5.6	0.4

Table I gives inspections of the EFV dewatered whole shale oil, together with inspections for the first pilot plant feed, described later in this section.

The EFV dewatered shale oil was distilled in a spinning band column into the following fractions: Start-400°F, 400-650°F, 650-800°F, and 835°F+ bottoms. Distribution by weight and volume is given in Table IV.

Tables V and VI show properties of fractions obtained in the spinning band distillation.

The water layer from the EFV separation was a dark color and contained dissolved oil. The pH of this layer was 9. When acidified to a pH of 5 using HCl, an oil layer separated immediately from the water. This oil layer was approximately 6 vol % of the water layer (0.35 vol % of the original emulsion).

Preliminary laboratory experiments showed that a reasonably good separation of water from the shale oil was obtained by simply heating the emulsion and allowing it to stand. Therefore, this technique was used to remove the water to prepare the whole shale oil pilot plant feed.

The eleven 30-gallon drums of shale oil (except for retain samples) were combined and heated in a large kettle under nitrogen blanketing to between 160°F and 170°F. After about six hours, a water phase was drawn off from the bottom of the kettle. This layer represented approximately 7 wt % of the emulsion and appeared to contain most of the fines.

The oil phase was filtered through a 15-micron filter, Honeycomb viscose filter tube Type 17R10S. The filter tube appeared clean after filtration, and we have no evidence that any appreciable quantities of additional fines were removed by the filter. The analysis for sediment and water by ASTM Standard Method D 96 was 0.1 vol %. (No water could be seen; the entire deposit appeared to be sediment.) Further filtration through a 1-micron Honeycomb viscose filter tube, Type E39R10SV, caused no change in the analysis for sediment. Again the filter was clean. The ash analysis by ASTM Method D 486 was 0.03% by weight for both samples of filtered material.

The filtered shale oil identified as WOW 3394 was further analyzed by means of a filter residue ash test. Results of using two millipore filter sizes were as follows:

Filter Size, Micron	0.45	0.8
Total Solids, ppm	252	234
Ash, ppm	194	168

All pilot plant tests for the whole shale oil hydrotreater were made with feed containing fines.

The filtered shale oil feed was further analyzed to compare other properties with the preliminary inspections obtained for the EFV dried sample. Results are given in Table I. More detailed analyses are given in the first column of Table II.

From the filter residue ash tests using X-ray diffraction (XRD), the ash was shown to consist of  $\text{Ca}_2\text{Al}_2\text{SiO}_7$ ,  $\text{CaMgSi}_2\text{O}_7$ , and  $\text{Fe}_3\text{O}_4$ . The filtrate from this test was analyzed for metals by emission spectrochemical analysis. Results are given in Table III.

In particular, it should be noted that the filtrate contains 70 ppm of iron that is not removed by the 0.45- $\mu$  filter. In addition to the metals listed in Table III, the shale oil contains 28 ppm of arsenic that is also not removed by filtration.

A second shipment of 300 gallons of Paraho shale oil was received from Laramie during the week of May 20, 1977. Again, the blend was heated to 170°F in a large kettle under nitrogen blanketing. The water and most of the fines formed a separate layer which was drained from the kettle. The remaining oil was filtered through a 15- $\mu$  filter. The material balance for the separation is as follows:

	Identification	Wt %
Dewatered Shale Oil	WOW 3511	93.6
Water Plus Fines	WOW 3512	5.4
Residue at Bottom of Kettle	SGQ 6160	0.7
Loss	-	0.3

Table II shows that the second dewatered oil, WOW 3511, is essentially the same as the first dewatered whole shale oil sample, WOW 3394. (The properties of WOW 3394 were used in the process engineering estimates to be described in Part II of this report.) The water and residue samples from the second shipment were returned to the Laramie Energy Research Center.

Task 2 - Whole Oil  
Hydrotreating

A. Screening Tests

Test Conditions--Pilot plant screening tests were made with the whole shale oil to determine the most appropriate of three commercial hydro-treating catalysts. Test conditions were:

745°F  
2200 psig Total Pressure (Approximately 1850 psia H<sub>2</sub>  
Partial Pressure)  
0.6 LHSV  
5000 SCF/bbl Recycle Gas Rate  
Recycle Gas Scrubbed With Water to Remove Ammonia and H<sub>2</sub>S

Each test is run downflow with a fixed bed of 130 cc of catalyst. The catalysts were sulfided in situ according to a procedure simulating commercial sulfiding. Feed was introduced at 500°F, and the temperature gradually increased over a period of eight hours to 745°F.

Catalysts--Three catalysts were tested: ICR 104, ICR 106, and ICR 109. ICR 104 and ICR 106 are nickel-tungsten-silica-alumina catalysts; ICR 109 is a nickel-molybdenum-silica-alumina catalyst. In Figure 6, whole liquid product nitrogen is plotted as a function of time onstream for the three tests.

1. ICR 106

Whole liquid product nitrogen for ICR 106 was less than 1000 ppm at 100 hours onstream and about 2000 ppm at 500 hours onstream (Run 81-4).

2. ICR 104

Product nitrogen for ICR 104 was 1600 ppm at 100 hours onstream and about 3000 ppm at 480 hours. (Run 75-142, shut down at 484 hours onstream.)

3. ICR 109

Product nitrogen was 2200 ppm at 100 hours onstream. (Run 75-143, shut down at 150 hours onstream.)

Because of the activity advantage of ICR 106, it is concluded that of the three catalysts tested, ICR 106 is the catalyst of choice for whole shale oil hydrotreating. Run 81-4 with ICR 106 was continued to study the effect of recycle gas rate and to obtain preliminary catalyst deactivation rate information (fouling rate), described later in this report.

This screening program by no means exhausts the possible Chevron catalysts and catalyst combinations for hydrotreating. However, this

program is a feasibility study rather than a process optimization study; and further catalyst testing is beyond the scope of the program.

#### B. Two-Stage Hydrotreating with ICR 106

Because of the large amount of nitrogen in the feed and the resulting high ammonia partial pressure in the gas phase after hydrotreating, it has been suggested that ammonia is acting as a catalyst poison, thereby decreasing catalyst activity. Therefore, an improvement in catalyst activity might be obtained if the hydrotreating is performed in two stages with intermediate removal of nitrogen from the recycle gas.

To quantify the effect, two pilot plant tests were made on ICR 106: (1) A first-stage hydrotreating test at 1.2 LHSV and other conditions the same as those for the previously discussed screening tests (745°F, 2200 psig total pressure, 5000 SCF/B recycle gas rate) and (2) a second-stage test at the same conditions using the product from the first stage as feed. The overall LHSV for the two stages was 0.6, the same as that for the single-stage catalyst screening test.

The results of the two-stage experiments, shown in Figure 7, can be compared with the single-stage results for ICR 106 shown in Figure 6. After 150 hours onstream, the whole liquid product nitrogen from both two-stage processing and single-stage processing was about 1000 ppm. Therefore, it is concluded that no significant benefit is obtained by two-stage hydrotreating at these nitrogen levels and at this particular set of conditions compared to single-stage hydrotreating with ICR 106 catalyst.

#### C. Effect of Gas Rate

After the single-stage screening test (Run 81-4) with ICR 106, a brief test was made to determine the effect of recycle gas rate on catalyst activity. Figure 6 shows that the gas rate was increased to 10,000 SCF/bbl from 5,000 SCF/bbl for a short period of time. A modest improvement in rate of nitrogen removal was observed. We estimate this to be equivalent to roughly 10°F improvement in catalyst activity.

#### D. Yields and Product Properties

During the catalyst screening tests, yield information and product inspections were obtained. The gross hydrogen consumption for all three catalysts tested in the single-stage screening tests is approximately 1800-2000 SCF/bbl.

Tables VII-X, inclusive, summarize liquid distribution and product properties for ICR 106 and ICR 104 after about 200 hours onstream.

Table VII shows that there is substantial conversion of 650°F+ feed to product boiling below 650°F. Most of this product boils between 400°F and 650°F range.

The product inspections indicate that the catalysts are effective for sulfur removal as well as nitrogen removal. The asphaltene content of the product is also low.

Table XI shows the product properties of the whole liquid product from whole shale oil hydrotreating with ICR 106 at 1.2 LHSV. This product was the feed to the second stage in the two-stage test.

During these tests, we had some problems with the liquid product let-down valves, probably as a result of some of the fines from the feed passing through the catalyst bed and coming out with the product.

#### E. Demonstration Runs

##### Run 81-4

Typical hydroprocessing catalysts may undergo a transient activity loss early in the run that can be attributed to "equilibration" with a particular feed. Following this transient period, the catalyst reaches a "lined out" or stable activity and deactivates at a much slower rate. To determine whether the activity change observed with ICR 106 in the screening run is the result of this type of equilibration or a true catalyst fouling, Run 81-4 was continued at the conditions of the screening test following the completion of the study of effect of gas rate.

Figure 8 summarizes the whole liquid product nitrogen for the first 1200 hours onstream. The catalyst undergoes a transient activity loss early in the run; the transient period is about 750 hours. Figure 8 shows that between 750 and 1200 hours onstream, no catalyst deactivation or fouling can be detected within the limits of the experimental data.

Figure 9 shows additional tests with ICR 106 in pilot plant Run 81-4. After about 1200 hours onstream, the temperature was increased from 745°F to 767°F. Product nitrogen decreased to approximately 700 ppm.

Figure 9 also shows that about 100 hours after the temperature increase, the LHSV was lowered to 0.3, half of its previous value. As a result, the product nitrogen was lowered to 30 ppm.

At 1440 hours onstream, the LHSV was decreased to 0.2. After 150 hours at this low LHSV, the product nitrogen decreased to 1.3 ppm. It should be pointed out, however, that the pilot plant is not ideally suited for operation at 0.2 LHSV. The total catalyst volume is 130 ml, and the feed rate is only 26 ml/hour. Because of the relatively large holdup in the product high pressure separator and low pressure product handling system, a large volume of higher nitrogen product from the previous

period of operation remains to contaminate the low nitrogen product for a relatively long period. Also, with the high level of denitrification required to convert 2.18% nitrogen shale oil feed to 1-2 ppm nitrogen product, any effects of channeling or catalyst bypassing is emphasized in this single-reactor system. Maintaining the desired recycle gas rate is also difficult at the low gas throughput required at 0.2 LHSV. Despite these operation problems, the results of this experiment are encouraging and demonstrate that essentially all of the nitrogen can be removed from shale oil in a single hydrotreating step.

Figure 10 compares the liquid product boiling range at three product nitrogen levels with whole shale oil. As expected, more hydrocracking occurs at the higher levels of nitrogen removal; the less nitrogen in the product, the lower the boiling range of the product. Most of the 1000°F+ residuum is hydrocracked to lower boiling product at product nitrogen levels of 1000 ppm and lower.

Figure 11 shows a plot of denitrification kinetics developed during these tests. Activation energy for nitrogen removal is approximately 35 kcal/g mole.

Preliminary process design studies indicated that rather than denitrifying to 1 ppm, it is more economically attractive to reduce the product nitrogen to 500-1500 ppm in the whole shale oil hydrotreater. The lighter products can then be separated from the higher boiling fraction before further hydrocracking. Therefore, after the brief 0.2 LHSV study described above, the LHSV was changed back to 0.6 in Run 81-4. Figure 9 shows that within experimental error, no activity was lost during the periods at the lower space velocities.

Figures 8 and 9 also show little catalyst deactivation can be detected for about 1000 hours between 750 and 1750 hours onstream. However, after about 1800 hours onstream, a rapid catalyst deactivation occurred. This was accompanied by problems in maintaining gas flow across the catalyst bed and plugging within the reactor. Because of pressure drop problems and the apparent loss of catalyst activity, the run was shut down after about 2000 hours onstream.

The catalyst was removed from the reactor in layers for analysis. The catalyst itself was free flowing, and there appeared to be no plugging in the catalyst bed itself. Rather, the plugging was due to a large deposit of black material in the preheat section of the reactor. This deposit appears to be the result of thermal decomposition of constituents of the shale oil in the presence of hydrogen prior to any contact of the feed with the catalyst. This deposit was analyzed by a combination of energy-dispersive and wavelength-dispersive X-ray fluorescence as follows: 30% iron, 11.4% arsenic, 2.8% zinc, 1% calcium, and 0.23% selenium. The deposit also contained 14.7% carbon and 15.4% sulfur.

Table XII shows the arsenic and iron distribution throughout the catalyst bed as determined by X-ray fluorescence.

### Run 86-50

A large-scale, whole shale oil hydrotreating pilot plant run was made with 650 ml of ICR 106 catalyst. This unit contains six reactors in series. The first reactor was charged with 130 ml of alumina to serve as a guard bed for removal of a large fraction of the feed arsenic and iron. The remaining five reactors each contained 130 cc of ICR 106 catalyst. This run was made at 0.6 LHSV, based on the ICR 106 catalyst volume. Other conditions were 2200 psig total pressure, 5000 SCF/bbl recycle gas rate, and 500 ppm target product nitrogen.

The purpose of Run 86-50 was to serve both as a demonstration run for whole shale oil hydrofining and as a feed preparation run for the scheduled downstream processing studies.

Figure 12 is a summary of average catalyst temperature and observed product nitrogen. After 250 hours onstream, the run was interrupted in order to reinsulate the furnace. Upon restarting, a plug developed in a 1/4-in. external diameter Aminco preheat line (0.081-in. internal diameter) carrying the feed to the guard bed. The material that caused the plug was similar to the powdery black material that caused the preheat plugging problem in the 2000-hour pilot plant test, Run 81-4. The temperature of the line that plugged in Run 86-50 was approximately 750°F.

### Run 86-51

In order to avoid the plugging problem that occurred in Run 86-50, the preheat and guard bed section of the pilot plant was altered so that the feed and hydrogen mixture is first heated in the guard bed itself rather than the very narrow preheat line. The guard bed is a reactor with an 0.951-in. internal diameter filled with alumina to provide a surface for deposition of iron, arsenic, and other materials that might otherwise act as catalyst poisons. The shale oil-hydrogen mixture moves upward through the 352-ml guard bed, then through a short segment of 0.359-in. internal diameter tubing, then downflow through an additional 83 ml of guard bed before reaching the first catalyst zone.

Run 86-51 is another large-scale, whole shale oil hydrotreating pilot plant run made with 650 ml of ICR 106 catalyst. Conditions were as follows: (a) LHSV, based on ICR 106 catalyst, is 0.6; (b) target product nitrogen was 500 ppm in the whole product, with manual temperature adjustment to control product nitrogen; (c) recycle gas rate is 8000 SCF/bbl; (d) during the first 225 hours onstream, total pressure was 2200 psig; at 225 hours onstream, the total pressure was increased to 2350 psig to adjust the hydrogen pressure to the target of 1750-1800 psia.

In contrast to the smaller-scale pilot plant tests which were essentially isothermal, Runs 86-50 and 86-51 showed an exothermic reaction that initially occurred in the first catalyst bed. The catalyst in the first bed reached a temperature about 30°F higher than the overall

average catalyst temperature. Also, the second catalyst bed operated at a temperature about 10°F higher than the average. Figure 13 shows the temperature profile across each of the five catalyst beds after about 500 hours onstream in Run 86-51. Changes in temperature gradients as the run progresses provided us with a means of following catalyst deactivation. Figures 14 through 17 show temperature profiles taken later in the run.

More light gases (methane and ethane) are produced in the runs with the temperature profile than the small-scale isothermal run. As a result, the hydrogen content of the recycle gas is lower. Therefore, in order to maintain the target hydrogen pressure of about 1750-1800 psia, the total pressure was increased after 225 hours onstream.

The time history of the run to 3450 hours is shown in Figure 18. After a 300-hour line-out period, the product nitrogen settled down in the 500-1000 ppm range. Hydrogen partial pressure slowly drifted down as the catalyst temperature was increased and methane production rate rose. At about 960 hours, a leak developed in the recycle gas system. By the time it was corrected, at 1200 hours, a substantial increase in catalyst activity had appeared. The leak had bled off enough methane to increase the hydrogen partial pressure from 1650 psia to 2100 psia. At 2100 psia, the catalyst is roughly 25°F more active than at 1625 psia. With the leak repaired, hydrogen partial pressure dropped again; and the catalyst lost activity. Because of the dramatic effect of hydrogen partial pressure on catalyst activity, an intentional gas bleed of about 500 SCF/bbl was introduced at 1440 hours. By 1510 hours, hydrogen pressure had increased by about 300 psia; and the catalyst activity, again, showed improvement.

At this point, our original shipment of raw shale oil ran out. While waiting for the next shipment from Laramie, we reran some partially hydrotreated shale oil collected during offtest periods on other runs. During the next 400 hours of operation, the partially hydrotreated feed contained 2000 ppm of nitrogen. The space velocity was increased to 1.0 LHSV to maintain the temperature in the same range used for whole oil hydroprocessing.

Shortly before the whole shale oil feed ran out, it was noted that a pressure drop had developed across the guard bed due to the deposits of iron and arsenic. By the time the new feed was ready, the pressure drop had increased to 800 psia. Estimated hydrogen partial pressure had decreased to about 1200 psia. The guard bed was recharged with fresh material, and the run continued on the new dewatered shale oil, WOW 3511, at 0.6 LHSV. This new shale oil was virtually identical to the first one. Inspections are given in Table II. A recycle bleed of 500 SCF/bbl was used to maintain the hydrogen pressure at about 2000 psi. At this pressure, catalyst deactivation rate is low; the catalyst fouled at an estimated average rate of roughly 0.005°F/hour during the last 2000 hours of operation. The guard bed was replaced for a second time at 3260 hours on stream when a pressure drop across the bed was again noted.

The run was terminated at 3509 hours when the feed supply was again exhausted. (A portion of the second batch of feed was coked, as described in a later section of this report.)

About 60 hours before the end of the run, the space velocity was changed to 0.3 LHSV and the temperature to 772°F. The recycle gas bleed was discontinued during this period. Whole product nitrogen was 200 ppm.

Analytical information of the composition of the guard bed that was replaced at 1950 hours onstream is shown in Table XIII. Figures 19-21 show a profile of the composition of the aged guard bed and a typical temperature profile across the guard bed. In studying these figures, remember that the guard bed is operated in an upflow mode.

The figures show that arsenic and iron are found throughout the guard bed, even the portion with a temperature of 200-300°F. The largest amount was deposited in the part of the bed with temperatures ranging from 400-700°F. Plugging occurred when the iron content was 15 wt % and the arsenic content was 7 wt %. The result suggests that if a more gradual temperature is imposed upon the guard bed, more efficient use of the guard bed can be obtained. With more uniform distribution of these metals, a longer period of operation is expected before it is necessary to replace the guard bed.

The shale oil feed contains 70 ppm of iron and 28 ppm of arsenic for an iron/arsenic ratio of 2.5/1. The profile shows that the iron comes out more easily than arsenic. At the bottom of the bed (inlet) which operates at about 200°F, the iron/arsenic ratio is 6/1. At the top of the bed (outlet), which operates at over 750°F, the iron/arsenic ratio is slightly less than one.

Assuming the analyses are correct, virtually all of the arsenic from the feed can be accounted for in the guard bed; 70% of the iron from the feed is accounted for by the iron found in the guard bed. The guard bed is also effective for removal of zinc and selenium from the feed.

Table XIV shows the composition of the guard bed removed at 3260 hours. This guard bed had been in service for somewhat less time than the first guard bed that was replaced; therefore, the total quantity of metals deposited is somewhat lower although the metals profile is generally smaller.

Tables XV-XXIX, inclusive, summarize yields, hydrogen consumptions, and product properties obtained during the demonstration tests with ICR 106 catalyst.

Figure 22 summarizes the hydrogen consumption results as a function of product nitrogen. Results obtained with the larger pilot plant appear to be slightly lower at a given pressure and are probably more reliable.

Tables XXX-XXXVI show properties of large-scale product blends and fractions from these blends. Some of these fractions were used as feeds in the downstream processing studies described in later sections of this report. Note the comparison between the properties of Arabian VGO, Sumatran VGO, and the hydrotreated shale oil shown in Table XXX.

Tables XXXVII-XL compare the properties of the hydrotreated shale oil at 500-700 ppm with product nitrogen to the properties at 1.3 ppm and 30 ppm nitrogen.

Task 2A - Alternate  
to Whole Oil Hydrotreating  
Coking of Whole Shale Oil;  
Hydrofinishing of Coker  
Distillate

A. Coking of Paraho Shale Oil

Coking has been proposed as a first processing step in the conversion of raw shale oil into distillate fuels. It has the advantage of removing any shale solids suspended in the raw shale oil with the coke. Furthermore, the coker distillate may be more easily hydrotreated than the raw shale oil.

Two drums of Paraho shale oil (110 gallons) were coked to provide sufficient coker distillate for HDN processing studies. The oil was coked in a batch destructive distillation carried out in a 500-gallon shell still fitted with an 8-ft packed column to provide reflux. A column top temperature of 450-550°F was maintained by an air-quench coil in the top section of the column. Still pot temperature was raised gradually by controlled gas firing of the still to a maximum temperature of 975°F. This maximum temperature was maintained for four hours after the last liquid came overhead, at which time gas evolution had ceased. Gas generated in the coking was passed through a caustic scrubber, a gas meter, and was periodically sampled. The coke was chipped from the cooled still pot.

The same Paraho shale oil was also coked in smaller quantities in an all-glass laboratory batch coking still. The still pot was fitted with a short-jacketed Vigreux column, which was heated by an electric tape. Column top temperature was further controlled with an air quench stream to the jacket. Charge for a coking run in this apparatus was 250 grams. A total of three runs was made, two at a nominal 650°F and one at 450°F column top temperature.

Inspections for the whole shale oil coking feed (WOW 3511) and for the coker distillate liquid product from the large coking run (WOW 3524) are given in Table XLI. Table XLII gives properties of narrow boiling fractions of the coker distillate. Table XLIII shows the yields from the large-scale run and the three laboratory runs. Coker distillate yields are given as both weight and liquid volume percent based on feed, as actually produced, and on a C<sub>5</sub>+ liquid basis. The latter yields are slightly higher due to transfer of C<sub>5</sub>+ hydrocarbons from the gas to the coker distillate.

Gas yields were determined by difference and by calculation using the metered gas volumes and the chromatographic analyses of the gas. The latter method of calculating gas yields consistently gave smaller yields than the difference method (100% - % coker distillate - % coke). This difference in gas yields for the large-scale coking run was 4.63% on feed. About 0.75% of this can be accounted for by the cracking of S, N, and O compounds in the shale oil to H<sub>2</sub>S, NH<sub>3</sub>, and H<sub>2</sub>O, which were

removed from the gas stream by the caustic scrubber. The probable explanation for the remaining 3.88% is that one of the drums of shale oil that was coked contained some water. (This drum included the first material removed after the separation of the water layer; the technician who made the separation believes that some water was taken into this drum.) The difference between gas yields by the two calculation methods is much less in the case of the laboratory coking runs where the coker feed was withdrawn from a 1-gallon container. Here, most of the difference can be accounted for by the evolution of  $H_2S$ ,  $NH_3$ , and  $H_2O$  into the gas stream.

Of greater interest is the explanation for the seemingly wide range of coke yields obtained in the four coking runs. Coke yield is determined by the manner in which the coking is done, i.e., by the temperature maintained at the top of the reflux column. Figure 23 shows the coke yield is a function of the coker distillate end point, the higher the end point, the lower the coke yield. End point temperatures were obtained from simulated TBP distillations (temperature-programmed chromatograph) of the coker distillates. The end point temperatures were obtained by a linear extrapolation of the 60% and 90% overhead temperatures to 100%. The correlation shown in Figure 23 holds for both the laboratory and the large-scale coking runs. In a commercial delayed coker, the coker distillate end point is controlled by the operation of the fractionating column, which returns cycle oil to the coker.

The properties of the coke made in the large-scale coking of the Paraho shale oil are shown in Table XLIV. Nitrogen, arsenic, and metals concentrate in the coke.

Average composition for the gas produced in the large-scale coking run is shown in Table XLV. The main components of the gas are hydrogen, methane, and ethane; these three compounds constitute almost 80 mole % of the gas. The composition is typical of a gas from a thermal decomposition process.

It is concluded that coking of shale oil is a viable first processing step in the conversion of raw shale oil to distillate fuel products. The coke yield can be as low as 5-10% if the coking process is controlled to produce a high end point coker distillate.

#### B. Hydrofinishing of Coker Distillate

The coker distillate (WOW 3524, Table XLI) described above was hydro-treated in Pilot Plant Run 81-10, a 1440-hour test with 130 cc of ICR 106 catalyst. Test conditions were as follows: 1700 psig total pressure, approximately 1500 psia hydrogen partial pressure, 1.0 LHSV, 8000 SCF/bbl recycle gas rate, and average catalyst temperatures ranging from 715-735°F.

Purpose of the run was to determine catalyst activity and to obtain yields, hydrogen consumptions, and product inspections at several levels of denitrification. Although it was not our intention to accurately determine the catalyst fouling rate, the run was continued long enough to show that the catalyst is stable. Total run length was 1440 hours. Figure 24 summarizes the run history and shows the nitrogen in the St-350°F product and the 350°F+ product.

After an initial 300-hour transient period of catalyst stabilization, the catalyst activity appears constant. However, at 630 hours, the feed was stopped and the unit depressured for a minor equipment modification; when the feed was restarted, the catalyst appeared to have lost 10-15°F in activity. (Other unit upsets such as low recycle gas rates may have contributed to the activity loss.) During the following 700 hours of operation, the catalyst activity again appears stable; and, within limits of measurement, no catalyst fouling is observed.

Figure 25 shows the effect of operating severity (as indicated by product nitrogen) on chemical hydrogen consumption. At 1500 ppm nitrogen in the 350°F+ product, hydrogen consumption is 1150 SCF/bbl. At 100-200 ppm nitrogen in the 350°F+ product, hydrogen consumption is 1250 SCF/bbl.

Table XLVI summarizes yield periods from the run. Tables XLVII and XLVIII summarize properties of the 350°F+ product (diesel boiling range). Tables XLIX and L show properties of various fractions distilled from the 350°F+ product. Table LI shows that the 350°F+ product contains about 8% of 650°F+ material of high pour point. Therefore, the entire 350°F+ product has a high pour point (~25°F) and would not meet typical diesel specifications. A lower end point coker distillate feed would be required as feed for the entire 350°F+ fraction to meet typical diesel pour point specifications. With a coker distillate feed of the end point used in these studies, the higher boiling material would have to be removed by distillation to make an acceptable diesel. For example, the 300-600°F diesel has a pour point of -15°F to -20°F. The 300-650°F diesel has a pour point of 5°F. The actual specification will vary, depending on the area, time of year, etc.

One of the analyses of particular importance for the diesel product is the thermal stability test. The test used in these studies is a Chevron adaption of Du Pont Test Method No. 21-61.\* In Du Pont's method, a test blotter is used to rate the stability. However, because evaluation of the result is visual and subjective, Chevron has found that better repeatability is obtained by evaluating the test filters using a reflectometer. A reflectance of 80% or higher is recommended as a passing rating for automotive diesels. (This is comparable to a test blotter rating of No. 6 or below by the Du Pont method.) For some

\*Du Pont Petroleum Laboratory Test Methods, E. I. du Pont de Nemours and Company, Wilmington, Delaware, June 1967.

railroad diesels, a reflectance of 70% or above is acceptable. (The latter rating compares to a test blotter rating of No. 7 or below by the Du Pont method.)

Figure 26 shows that the thermal stability of the diesel is excellent based on this accelerated test, even at nitrogen contents as high as 2000 ppm.

Table LI gives a summary of inspections of the 180-350°F naphtha produced. It is likely that the desired nitrogen content of the naphtha (to be fed to the first stage of a reformer) will dictate the selection of target product nitrogen for the coker distillate hydrotreater.

Table LII gives the analysis of the light naphtha (C<sub>8</sub>-180°F).

Table LIII summarizes the composition of deposits on the aged catalyst following Run 81-10. The run was performed in a single-tube downflow reactor with 118 ml, 75.6 g of alumina ahead of the catalyst to act as a guard bed to remove arsenic from the feed. The results show that much of the arsenic was removed by the guard bed and that remaining was deposited in the first 25 ml of the catalyst bed. (The arsenic content of the feed is 6 ppm.)

### Task 3 - Middle Distillate Hydrotreating

The 350-650°F fraction of the product from hydrotreating whole shale oil to about 500 ppm nitrogen did not pass the stability test for diesel described under Task 2 Alternate, Part B. This contrasts with the hydrotreated diesel from the coker distillate which passed the stability test at 500 ppm nitrogen. (See also Task 2 Alternate, Part B.)

Two approaches were tried for making salable diesel from the hydro-treated shale oil. The first was to distill to a lower diesel cut point. The second was to further hydrotreat the diesel fraction. The results are described below.

#### A. Effect of Cut Point on Diesel Stability

A blend of product from the whole oil hydrotreater (WOW 3535, see Table LXV) was fed to a two-in-series continuous distillation setup. The temperatures were adjusted such that the overhead from the second still had an initial point of 350-375°F. The diesel cut point was varied from 585-650°F.

Figure 27 shows a plot of the diesel stability versus diesel cut point. In this figure, diesel cut point is defined as that temperature at 95 LV % distilled as determined by simulated chromatographic distillation. Figure 28 shows a plot of the pour point and freeze point versus diesel cut point. At the cut point of 590°F, a low pour, fairly stable diesel can be made. At higher cut points, the stability is marginal or unacceptable at this nitrogen level.

#### B. Diesel Hydrotreating

The diesel fraction, nominally 350-650°F, from whole oil hydrotreating was further hydrotreated in Run 66-185 using ICR 106 catalyst. The inspections of the feed, SGQ 6198, are shown in Table LIV. The conditions for this run were 2.0 LHSV, 1200 psig total pressure, approximately 1100 psia hydrogen pressure, and 5000 SCF/bbl recycle gas rate. A history of the temperature and nitrogen level obtained is shown in Figure 29. Table LV shows a typical yield period. The hydrogen consumption is 100 SCF/bbl or lower (about the limit of accuracy of measurement).

The major reason for further hydrotreating of this diesel was to improve the stability as measured by the standard thermal stability test. Table LVI shows some inspections on whole liquid product from Run 66-185. The thermal stability is markedly improved over that of the feed even at a very low level of denitrification.

During Run 66-185, there were several periods of severe pressure fluctuations of several hundred pounds resulting from a faulty pressure

controller. As a result, recycle gas flow was erratic and probably stopped completely for appreciable lengths of time. We believe this resulted in catalyst deactivation and presumably accounts for the unusual response of product nitrogen to increases in temperature. Unfortunately, sufficient quantities of feed are not available to repeat this run.

In a second run, Run 66-187, SGQ 6239 was used as the feed. Inspections are shown in Table LVII. This feed was obtained by combining product from a continuous distillation of a blend of whole hydrotreated shale oil.

Figure 30 shows the history of gravity, product nitrogen, and average catalyst temperature for Run 66-187. The limited amount of feed precluded a long run. Following an initial transient, the product nitrogen level was fairly constant for ~100 hours. (Two nitrogen analyses appear to be in error and are labeled "questionable.")

### C. Jet Hydrotreating

The quality of the jet cut that could be obtained by additional hydrotreating was evaluated both indirectly and directly. The indirect method was to take product from the diesel hydrotreating run, Run 66-185, and distill to a jet end point. Results are shown in Table LVIII. Note, especially, that at a 500°F jet end point the gum level was 4 mg; although the nitrogen is high for stable jet fuel. A typical jet specification for gum is 5 mg maximum. The smoke point is marginally below specification. Therefore, a slightly higher degree of aromatics saturation is required for specification jet fuel; the additional hydrogenation would also result in more nitrogen removal. Further work would be necessary to define precisely the limits in nitrogen content for stable jet fuel.

The direct method for evaluating jet quality was to prepare a jet boiling range fraction from our continuous distillation of hydrotreated shale oil. The inspections on this jet cut are shown in Table LIX. This feed was then processed in Run 66-187. The conditions, except for a higher catalyst temperature, were the same as those used for hydrotreating the diesel; namely, 2.0 LHSV, 1200 psig, and 5000 SCF/bbl once-through hydrogen. Because of the limited amount of feed, we were only able to run for two days. The temperature, product nitrogen, and gravity are shown in Figure 30. Other product inspections are shown in Table LX. The gum level is acceptable (2-5 ppm), and the nitrogen level is low (2-5 ppm). At this low nitrogen level, the jet fuel is expected to meet stability specifications.

Task 4 - Extinction  
Recycle Hydrocracking

A. Extinction Recycle  
Hydrocracking of 650°F+  
Hydrofined Shale Oil

The 650°F+ fraction of the hydrotreated shale oil (SQQ 6152, Table XXXII) was hydrocracked to extinction in a single stage in pilot plant catalyst screening run, 81-8, with 130 ml of ICR 106 as the catalyst.

Target conditions were as follows: 0.6 LHSV, 2300 psig total, ~2000 psia H<sub>2</sub>, 5000 SCF/bbl recycle gas rate, 525°F recycle cut point. Lined out per pass conversion was 40% at 795°F.

Yields and product inspections are given in Tables LXI-LXIV, inclusive.

The catalyst is less active than expected, based on correlation with petroleum stocks. It is likely that the small amount of high boiling material (900°F+) is quite refractory and is not converted readily. It appears desirable to remove this material by distillation and use it as refinery fuel rather than crack the entire 650°F+ hydrotreated shale oil.

B. Extinction Recycle  
Hydrocracking of 625-850°F  
Hydrotreated Shale Oil

Pilot plant Run 81-11 is a test to determine the activity of ICR 106 for single-stage extinction recycle hydrocracking of the 625-850°F fraction of hydrotreated shale oil containing about 500 ppm of nitrogen and to obtain yields, hydrogen consumptions, and product inspections when operating at about 535°F recycle cut point. Feed inspections are given in Table LXV.

Due to the shortage of feed, the pilot plant was lined out using Arabian Light VGO as feed. During this period some equipment problems were corrected.

After 650 hours onstream, the feed was switched to 625-850°F hydro-treated shale oil feed. As in the previous test in which a 650°F+ hydrotreated shale oil was hydrocracked, the catalyst activity is somewhat lower than expected, based on results with the Arabian feed-stocks. However, the 850°F end point feed is considerably easier to process than the entire 650°F+ fraction of the hydrotreated shale oil. Further work would be required to determine the optimum feed boiling range for this process.

Run conditions were 1.0 LHSV, 2350 psig, ~2125 psia hydrogen, 8000 SCF/bbl recycle gas rate, and a recycle cut point of approximately 550°F. Conversion was about 52% per pass below 535°F at 785-790°F.

Results of four yield periods are given in Tables LXVI-LXIX, inclusive. Product inspections are given in tables. Note that the nitrogen content is less than 1 ppm, the smoke point is high (over 30), and the freeze point is low (less than  $-60^{\circ}\text{F}$ ).

A possible alternate to the single-stage hydrocracker is a two-stage hydrocracker. The feed is further hydrotreated in the first stage to remove most of the remaining nitrogen. The denitrified product is cracked to extinction in the second stage. Insufficient feed was available to pilot test this type of hydrocracker; further experimental work would be required to determine which type of hydrocracker is preferable for processing shale oil.

Task 4A - Alternate Processing  
of 650°F+ Hydrofined Shale Oil: FCC

A. Small-Scale Catalytic  
Cracking Tests

Catalytic cracking is a possible downstream processing step for the refining of the hydrofined shale oil into transportation fuels.

Feed--Preliminary small-scale catalytic cracking tests were made on the 650°F+ fractions of two hydrofined Paraho shale oils. The two oils differ in the hydrofining severity, i.e., the extent of denitrification employed in their preparation. Nitrogen contents of the 650°F+ hydrofined shale oil fractions were 385 ppm for the more severely hydrofined and 870 ppm for the less severely hydrofined products used in these tests. Lower boiling material was removed from the hydrofined shale oils by distillation. The inspections for these two oils are given in the first two columns of Table XXXII.

The hydrofined shale oils were cracked in a fixed fluidized bed test unit using a equilibrium catalyst withdrawn from an operating FCC unit. The quality of the catalyst is shown in Table LXX. It is a moderately active, moderately metal-contaminated zeolite catalyst.

Pilot Plant Description--The fixed fluidized bed test unit is operated in a cyclic manner. Preheated feed enters the bottom of a conical-shaped reactor where it vaporizes and fluidizes the catalyst in the reactor. Steam is added with the feed to aid in its vaporization. Oil is fed for a five-minute period after which a nitrogen purge sweeps the hydrocarbon vapors from the reactor. During the nitrogen purge the temperature of the reactor is raised to 1050°F; and with completion of the purge, air is introduced into the reactor to burn the accumulated coke from the catalyst. At the end of the regeneration period, a second nitrogen purge removes the air from the reactor, thus completing the cracking-regeneration cycle. Multiple cycles are usually made to provide sufficient liquid product for distillation and inspection of the distillate fractions.

The hydrocarbon gas from each cycle is collected in a gas holder, mixed, and sampled. Gas composition is determined chromatographically. The coke deposited on the catalyst is determined for each cycle by passing an aliquot of the regeneration gases over hot cupric oxide to convert carbon monoxide to carbon dioxide and then through an Ascarite absorption bulb.

Conditions--Nominal reactor conditions for the present study were:

Reactor Temperature, °F	975
Feed Rate, ml/Min.	60
Water (Steam) Rate, ml/Min.	4
Feed Period, Min.	5

The operating conditions chosen were those used at an earlier date to crack a different "Paraho-type" raw shale oil over the same catalyst and in the same test unit. This permits comparison of crackability and yields between a raw shale oil and hydrofined shale oils. The limited quantities of feeds available permitted only a single six-cycle run to be made with each of the two 650°F+ hydrofined shale oils. Yields and product qualities are discussed below.

Yields--The operating conditions and product yields are presented in Table LXXI. Yields are on a no-loss basis without correction for sulfur and nitrogen compounds in the reaction and regeneration gases. Conversion is defined as 100% minus the percent of 430°F+ liquid product based on feed volume.

The crackability of the hydrofined shale oil is dependent upon its nitrogen content because nitrogen compounds act as temporary catalyst poisons. The 385 ppm nitrogen feed gives 80.1 LV % conversion; whereas, the 870 ppm nitrogen oil gives 75.4 LV % conversion. The raw shale oil cracked earlier at the same conditions was only 38.9% converted. The coke yield is much higher with the raw shale oil (6.21% at 38.9% conversion) than with the hydrofined shale oils (4.41% at 80.9% conversion and 3.87% and 75.4% conversion). Other yield comparisons are not meaningful because of the widely different conversions obtained.

It is interesting to compare yields obtained in cracking hydrofined VGO's from a typical petroleum source. Two hydrofined gas oils from Arabian Light crude are selected for this comparison.<sup>1</sup>

1. CCL-4993                      27.3°API; 320 ppm Nitrogen
2. CCL-4990                      25.3°API; 860 ppm Nitrogen

The nitrogen contents of these two oils are about the same as those of the two hydrofined shale oils of the present study. These oils were cracked at 975°F over the same catalyst (CCL 4914) used in the shale oil study but in a different pilot cracking unit. The pilot plant used to study the cracking of the Arabian gas oils is a continuous catalyst circulation unit. As such, it is a closer approach to commercial FCC unit operation than the cyclic fixed fluidized bed operation of the present study. The small quantities of hydrofined shale oils available precluded use of the continuous catalyst circulation pilot unit during this preliminary stage. (However, as described in a later paragraph, runs were later made in the larger unit.) Product workup is similar for both types of units. Only minor yield differences are obtained in cracking a given oil in the two units to the same conversion.

<sup>1</sup>"A Pilot Plant Comparison of Riser and Dense Bed Cracking of Hydrofined Feedstocks," W. M. Haunschild, D. O. Chessmore, and B. G. Spars. Presented at the 79th AIChE Meeting, Houston, Texas, March 16-20, 1975.

The yields and operating conditions for the cracking runs made on the two hydrofined Arabian gas oils are shown in Table LXXII. Two runs are shown for each oil. They bracket the conversions obtained with the two hydrogenated shale oils. The hydrofined Arabian gas oils and the shale oils of about the same nitrogen content have roughly equivalent crackabilities based on the space velocities (WHSV) and catalyst/oil weight ratios employed in the two pilot plant operations. Coke yields for the shale oils were appreciably lower than those for the Arabian gas oils at constant conversion. C<sub>3</sub> yields were 1-3% on feed higher for the shale oils; C<sub>4</sub> yields were approximately 4% on feed higher for the shale oils. Gasoline yield for the more severely hydrotreated shale oil was about 3.5% lower than that for the more severely hydrofined Arabian gas oil. At the higher nitrogen level of about 860 ppm, the shale oil produced about 1% more gasoline than the Arabian gas oil at the same conversion.

Product Qualities--Product inspections for the light and heavy gasolines and for the light and heavy cycle oils produced from cracking hydrotreated shale oil are presented in Table LXXIII. The gasolines and cycle oils have properties well within the range which we experience for such fractions derived from the cracking of gas oils from petroleum. In this report the only comparison between hydrotreated shale oils and hydrofined Arabian gas oils will be that of the gasoline octane numbers. Table LXXIV makes that comparison. The F-1 clear octane numbers for the hydrofined shale oil light gasolines are 0.7 to 1.0 octane number lower than those from the Arabian gas oils. The F-2 octane numbers are, perhaps, about 0.5 octane number lower for the shale oil light gasolines. However, the shale oil heavy gasoline F-1 clear octane numbers are 0.5-0.7 higher than those of the Arabian heavy gasolines; while F-2 clear octane numbers are approximately the same for both types of heavy gasolines.

B. Larger-Scale Catalytic  
Cracking Tests (Continuous  
Catalyst Circulation  
Pilot Plant)

A third hydrotreated shale oil with a 1300 ppm nitrogen content was catalytically cracked in the continuous catalyst circulation pilot plant used for the hydrofined Arabian Light gas oil tests described in the previous section. Inspections of the feed (SGQ 6139; CCL 5420) are shown in Column 3 of Table XXXII. Three runs were made. The tabulation below shows the cracking test conditions, conversions, and coke yields.

PP Run 158-	511	512	513
Reactor Temperature, °F	930	930	975
WHSV, g/Feed/Hour/g Catalyst	7.85	7.86	7.71
C/O, g Catalyst/g Feed	6.71	4.39	7.10
Catalyst	Equilibrium	CBZ-1 (CCL 4904)	
Conversion, LV % (430°F-)	70.5	62.1	78.6
Coke, Wt %	3.61	3.01	4.87

The catalyst used in this study was essentially the same equilibrium catalyst (CBZ-1) employed in the fixed fluidized bed tests. The first two runs (511 and 512) were made at a reactor temperature of 930°F. Catalyst/oil ratio, i.e., catalyst circulation rate, was varied to give different conversions. The third run (513) was made at 975°F reactor temperature; the temperature of the tests with hydrotreated shale oil in the fixed fluidized bed pilot unit. The results of these tests support the general conclusions from the smaller-scale tests.

Comparisons of these yields are made with yields obtained in cracking two hydrofined Arabian Light VGO's (nitrogen contents were 320 ppm and 860 ppm) in the same pilot unit and employing the same equilibrium catalyst.

Table LXXV gives the cracking conditions, conversions, and yields for the three runs. Yields are on a no-loss basis without correction for sulfur and nitrogen compounds in the reaction and regeneration gases. Conversion is defined as follows: 100% - (430°F + liquid product, LV %).

The crackability of the hydrotreated shale oil is dependent upon its nitrogen content since nitrogen compounds act as temporary catalyst poisons. At 930°F, the hydrotreated Paraho shale oil with 1300 ppm N was cracked to about the same conversion as the hydrofined Arabian Light gas oil with 320 ppm N when the two stocks were cracked at constant severity. Figure 31 shows these relationships. Cracking severity is defined by the ratio: catalyst/oil + weight hourly space velocity (C/O + WHSV). The solid lines represent the hydrofined Arabian Light gas oil cracking data at 930°F for the two oils, and the dashed lines represent the hydrotreated shale oil cracking data at 975°F reactor temperature. At both temperatures, the Arabian gas oil with the lower nitrogen content gave the higher conversion at constant severity. Increasing reactor temperature, of course, increases conversion.

Coke yields for the hydrotreated shale oil are appreciably lower than those for the 860 ppm N Arabian gas oil at both 930°F and 975°F reactor temperature when compared at constant conversion. At 975°F, the coke yield for the shale oil equals that for the 320 ppm N Arabian gas oil at the same conversion. This is shown in Figure 32.

Gasoline ( $C_5$ -430°F) yields are shown in Figure 33. With the Arabian gas oils, the oil with the lower nitrogen content gives the higher gasoline yields at both 930°F and 975°F reactor temperatures at constant conversion. The higher reactor temperature gives the lower gasoline yield for both the 320 ppm and the 860 ppm N oils, primarily because of the higher  $C_3$  yield at the higher temperature. With the 1300 ppm N shale oil, the gasoline yield is intermediate between that of the two Arabian gas oils at 930°F. However, at 975°F, it is equal to that of the 320 ppm N Arabian gas oil.

The  $C_3$  and  $C_4$  yields are shown in Figures 34 and 35. The Arabian gas oils make the same amounts of propene and propane regardless of their nitrogen content. The yields of both propene and propane are higher at 975°F reactor temperature than at 930°F. The hydrotreated shale oil gives the same yields of  $C_3$ 's as the Arabian gas oils. Neither temperature nor nitrogen content affect the total  $C_4$  yield in the case of the Arabian gas oils. The hydrotreated shale oil gives the same yields of total  $C_4$ 's as the Arabian gas oils. Butene yields from the shale oil are closer to that of the 320 ppm N Arabian gas oil at 930°F and closer to that of the 860 ppm N gas oil at 975°F.

Product inspections for the light and heavy gasolines and for the light and heavy cycle oils from the cracking of the hydrotreated Paraho shale oil are shown in Table LXXVI. The gasolines and cycle oils have properties well within the range which we experience for such fractions derived from the catalytic cracking of gas oils from petroleum. In this report, the only comparison between the hydrotreated shale oil and the hydrofined Arabian gas oils will be that of the gasoline octane numbers. Figures 36 and 37 make this comparison for F-1 and F-2 clear octane numbers, respectively. Light gasoline ( $C_5$ -250°F) F-1 clear octane numbers for the hydrotreated shale oil are about the same as those for the light gasolines from the hydrotreated Arabian gas oils at both 930°F and 975°F reactor temperatures, at the same conversion. The heavy (250-430°F) gasoline F-1 clear octane numbers from the shale oil are one to four numbers lower than those for the heavy gasolines from the Arabian gas oils at both reactor temperatures. The light gasoline F-2 clear octane numbers of the shale oil gasolines are equal to or slightly higher than those of the Arabian gas oil light gasolines. Again, the heavy gasoline F-2 octane numbers for the gasolines obtained from the shale oil are two to three numbers lower than those of the Arabian gas oil heavy gasolines. It is concluded that Paraho shale oil makes an excellent catalytic cracking feedstock provided its nitrogen content is reduced to about 0.15% or lower by appropriate hydrotreating. Gasolines and cycle oils derived from the cracking of hydrotreated shale oils are similar to those obtained from the cracking of hydrofined petroleum gas oils.

### Task 5 - Catalytic Reforming

The naphtha products from hydrotreating whole shale oil and shale oil coker distillate have properties that indicate that they will be good reformer feeds. These properties are within the scope of the range of properties of petroleum naphthas which serve as the data base for the correlations used in Chevron's Rheniforming process. Therefore, catalytic reforming performance was estimated by correlations; and no pilot plant work was undertaken. It will be necessary to further hydroprocess these naphthas to remove the remaining nitrogen in the naphtha as the first stage of the Rheniformer. The denitrified naphtha is then catalytically reformed to high octane gasoline in the second stage over a bimetallic catalyst.

Naphtha produced in the hydrocracker is nitrogen free and can be fed directly to the second stage of the Rheniformer.

Task 6 - Final Process Design

The results of the process engineering studies and estimated processing costs referred to as the "Final Process Design" in the work statement of the contract are summarized in Part II of this report.

### Task 7 - Reports

All progress reports and quarterly reports have been issued on schedule. A list of the quarterly reports is given in Section VI, together with a list of papers summarizing portions of this work.

According to the contract, a final report will cover the work on all four feedstocks. For practical reasons, it is convenient to issue this report in several parts as interim reports. The present report covers the work on the first feedstock Paraho shale oil and is identified as the first interim report. The results complete the work on the first feedstock except for comparisons with later feeds.

Task 8 - Distillate Shale  
Oil Hydrotreating

This task was added by DOE in Modification No. A004 at the request of Professor Yesavage and includes hydrotreating of distillate fractions from both in situ and ex situ retorting. Professor Yesavage provided us with two shale oil distillates and requested that we hydrotreat each to three different levels of severity. The inspections of these feeds are listed in Table LXXVII. These feeds were hydrotreated in pilot plant Run 66-184 at the following conditions: 0.5 LHSV, 1900 psig, and 5000 SCF/bbl once-through hydrogen. The catalyst used was ICR 106. Temperature was varied to provide the different hydrotreating severities.

The target nitrogen levels and the experimental levels we obtained are as follows:

Laramie Retort SQQ 6131		TOSCO II SGQ 6132	
Target	Pilot Plant Results	Target	Pilot Plant Results
0.18 Wt % S	0.14 Wt % S	0.20 Wt % S	0.13 Wt % S
0.4-0.7 Wt % N	0.43 Wt % N	0.4-0.7 Wt % N	0.30 Wt % N
	0.73 Wt % N		0.82 Wt % N
<0.2 Wt % N	0.079 Wt % N	<0.2 Wt % N	0.014 Wt % N

We deviated somewhat from the original request for three reasons: (1) the extremely limited amount of feed which we were given (5 gallons of SQQ 6131 and approximately 8 gallons of SQQ 6132), (2) our unfamiliarity with the low severity hydroprocessing which was requested in one case, and (3) wide pressure fluctuations through critical parts of the run. Figure 38 shows a history of temperature and the product nitrogen or sulfur observed. The offspecification material from the first 209 hours of operation, when SQQ 6132 was the feed, was combined and labeled SQQ 6183. This product had a gravity of 28.9 and sulfur of 0.63 wt %. It was then processed to afford the desired product.

We supplied Professor Yesavage with products from the four levels of hydrotreating on each feed as well as all offspecification products. Inspections and quantities of the blends are listed in Table LXXVIII. Inspections on offspecification material or material produced during temperature line-out are shown in Table LXXIX.

Although the pressure variation during this run complicates the kinetics, the tremendous variation in severity of denitrification provides the opportunity for determining the energy of activation for denitrification. Figure 39 shows a plot of all the denitrification data as

first-order rate constants versus the reciprocal of the absolute temperature. The activation energy determined from this plot is approximately 35 kcal/mole. This activation energy agrees with the result for whole shale oil hydrotreating with ICR 106.

### III. Areas for Future Study

The experimental work under this contract demonstrated that a Paraho shale oil can be refined to high quality transportation fuels using advanced state-of-the-art commercial petroleum refining technology. Areas for future study are mainly related to optimization. The following examples are typical of the second-order considerations mainly related to optimization that would be addressed in implementing an actual commercial project.

A. Stability limits for diesel and jet fuel. It was demonstrated that both stable diesel and stable jet fuel can be obtained by normal middle distillate hydrotreating. However, stability limits would ordinarily be investigated to provide operating flexibility for an actual commercial project.

B. Optimum nitrogen content of FCC feed. At a given nitrogen level and boiling range, the FCC results were generally more favorable than for petroleum stocks. Therefore, even higher FCC feed nitrogen levels might be considered.

C. Variation of feed and product cut points and process conditions. The existing data and correlations can supply much of the basis for optimization of the refining schemes to most economically produce a specific product slate.

D. Heat exchanger fouling. Because of the unstable nature of shale oil and the contaminants it contains, particular attention should be paid to the possibility of heat exchanger fouling in the design of a facility for handling raw shale oil. The tendency to foul heat exchangers may be different for oils produced from shales from different parts of the oil shale deposits. It most certainly could be affected by the retorting process employed. The petroleum industry is experienced in handling unstable stocks in the processing of thermally cracked materials such as coker distillates. The technology for handling shale oil would be selected from available petroleum technology when the potential problem is more clearly defined in a specific application.

E. Choice of catalysts. Experience in petroleum processing has shown that it is often desirable to provide a relatively cheap catalyst at the inlet of the first catalytic processing zone. Such a catalyst may be less active and have a shorter life than the catalysts used in further processing. However, it would serve to eliminate any metal contaminants that on occasion might pass through the guard bed. The accumulation of metal contaminants would make regeneration of any catalyst difficult. The inlet catalyst, because of its low cost, could be discarded at the end of each cycle rather than regenerated. This would permit relatively uncomplicated regeneration of the major active catalyst. This would be a significant improvement over the replacement of the complete catalyst charge for each run as was assumed in these studies.

None of the above areas for future study represent major problems; they are the kind of second-order improvement areas that will likely require attention after the desired product distribution has been selected in order to maximize overall product values for business reasons.

There is, however, one general area deserving of more attention. This involves the provisions made to remove arsenic, iron, and other metallic contaminants from raw shale oil. The problem is different in degree but not in kind from a similar problem for which there is demonstrated technology in petroleum refining; that is, the removal of metallic constituents with a guard bed.

This is the solution adapted in the present studies. This is a practical and demonstrated technique. In petroleum refining, the use of a guard bed ahead of the primary hydroprocessing has proved to be cheaper than more sophisticated approaches. On the other hand, we note that treating processes have been suggested for arsenic removal. If a treating process rather than a guard bed is effective in removing arsenic in the primary hydroprocessing unit and its cost is not greater, it would doubtless find a place in shale oil refining.

In this connection, the unusually large amount of iron in the shale oil should be noted. It is possible that this was picked up from the small-scale, high surface-to-volume ratio pilot retorts employed in its production. It is also possible it came from the barrels in which it was stored.

It may be necessary to learn more about the metals content of shale oils from other parts of the oil-shale deposits before one can be content to choose a most appropriate means for their removal.

In conclusion, we would like to enlarge on this last point. This study was devoted to the processing of a single type of surface-retorted shale oil. Properties of shale oils appear to be similar in most ways, and we would expect the same refining routes to be applicable. However, there are possible differences depending upon either the properties of the oil shale or the retorting process used to produce the shale oil. It is likely that some optimization may be required for each oil. In particular, much current attention is being devoted to in situ retorting of oil shale. The subject of refining of in situ shale oil thus represents a significant area for possible future work.

#### IV. Conclusions

It is concluded, based on pilot plant results, that refining shale oil is feasible using current advanced state-of-the-art hydroprocessing technology. Sufficient pilot plant work was done to demonstrate the feasibility of three refining schemes: (1) hydrotreating of the whole shale oil followed by hydrocracking as the major conversion process, (2) hydrotreating of the whole shale oil followed by fluid catalytic cracking as the major conversion process, and (3) coking of shale oil followed by hydrotreating of the coker distillate. In each case, salable transportation fuels are the end products. Choice of the refining scheme in any particular application will probably depend on the product slate and flexibility desired.

The key step in the first two refining schemes is hydrotreating of the whole oil. A 3500-hour pilot plant run clearly demonstrated that a commercial length run could be expected. A guard bed can successfully remove most of the iron and arsenic upstream from the hydroprocessing catalyst bed.

The synthetic crude produced by the whole oil hydrotreater resembles a typical petroleum distillate. This hydrotreated shale oil can be converted to transportation fuels such as high octane gasoline, diesel, and jet fuel in conventional refining facilities.

Hydrotreated shale oil has an advantage as a refinery feed. In contrast to most petroleum oils, it contains essentially no residuum. Properties of the hydrotreated oil are similar to those of distillate fractions from Arabian or Sumatran crude petroleum oils. An exception is the sulfur content which is much lower for the hydrotreated shale oil than for most crudes.

The proposed refinery using hydrocracking as the major conversion step offers the most flexibility and is the preferred processing scheme if jet fuel is to be a major product. A high smoke point, low freeze point, stable jet fuel is produced by hydrocracking. It was shown, however, that hydrotreated shale oil is somewhat more difficult to process by hydrocracking than a Middle Eastern vacuum gas oil of the same boiling range.

The 650°F+ fraction of hydrotreated shale oil is an excellent feed for a fluid bed catalytic cracking unit, and catalytic cracking may be the preferred conversion process if high octane gasoline is the major product. Product yields and properties are similar to those obtained from hydrofined Arabian gas oil. Coke yields are low due to the high hydrogen content of the hydrotreated shale oil.

In either of the first two refinery plans, diesel fuel can be produced in large quantities by hydrofinishing a diesel fraction of the hydro-treated shale oil.

If the shale oil is coked prior to hydroprocessing as in the third refining scheme, the coker distillate can be hydrotreated directly to diesel fuel and a naphtha feed for a catalytic reformer. Less hydrogen is required in a coking refinery than if the whole oil is hydrotreated; however, 17-22% more raw shale oil is required to produce an equivalent volume of transportation fuel. The quantity of coke produced can be varied by changing the severity of coking.

The current work is a feasibility study rather than a process optimization study. The three refining schemes by no means exhaust the possible combinations of processes available in existing technology for a shale oil refinery. A possible scheme that is frequently suggested (but was not included in the study) is distillation of the crude shale oil before any downstream processing. It is expected that the light distillate fractions would be easier to hydrotreat than whole shale oil and that the highest boiling fraction would be more difficult to hydrotreat than the full boiling range stock. In a small-to-medium sized refinery such as those considered in Part II of this report, upgrading the whole shale oil in the first-stage hydrotreater offers the advantage of being less complex and, therefore, most likely less costly than one in which the fractions are hydroprocessed separately. In a very large refinery, the added flexibility resulting from hydrotreating the various fractions separately could offer advantages to outweigh the disadvantage of increased complexity. Further study would be required to evaluate this case.

Other areas in which future study is desirable are mainly related to process optimization. No area was identified that represents a major problem. Rather, most involve the kind of second-order process refinements that will likely come after a specific commercial project and a desirable slate of products have been selected.

V. List of Quarterly Reports and  
Papers Covering Portions of  
Work Under DOE Contract  
EF-76-C-01-2315

1. R. F. Sullivan, "Refining and Upgrading of Synfuels from Coal and Oil Shales by Advanced Catalytic Processes," Quarterly Report for the Period July-September 1976, UC-90d, Report No. FE-2315-4, October 1977.
2. R. F. Sullivan, "Refining and Upgrading of Synfuels from Coal and Oil Shales by Advanced Catalytic Processes," Quarterly Report for the Period October-December 1976, UC-90d, Report No. FE-2315-8, January 1977.
3. R. F. Sullivan, "Refining and Upgrading of Synfuels from Coal and Oil Shales by Advanced Catalytic Processes," Quarterly Report for the Period January-March 1977, UC-90d, Report No. FE-2315-12, April 1977.
4. R. F. Sullivan, H. A. Frumkin, C. E. Rudy, and H. C. Chen, "Refining and Upgrading of Synfuels from Coal and Oil Shales by Advanced Catalytic Processes," Quarterly Report for the Period April-June 1977, UC-90d, Report No. FE-2315-15, July 1977.
5. R. F. Sullivan, C. E. Rudy, D. C. Green, and H. C. Chen, "Refining and Upgrading of Synfuels from Coal and Oil Shales by Advanced Catalytic Processes," Quarterly Report for the Period July-September 1977, UC-90d, Report No. FE-2315-19, October 1977.
6. R. F. Sullivan, D. C. Green, and H. C. Chen, "Refining and Upgrading of Synfuels from Coal and Oil Shales by Advanced Catalytic Processes," Quarterly Report for the Period October-December 1977, UC-90d, Report No. FE-2315-22, January 1978.
7. R. F. Sullivan and B. E. Stangeland, "Catalytic Hydroprocessing of Shale Oil to Produce Distillate Fuels," Symposium on Refining of Synthetic Crudes, Petroleum Division, 174th National American Chemical Society Meeting, Chicago, Illinois, August 28-September 2, 1977.
8. R. F. Sullivan and B. E. Stangeland, "Converting Green River Shale Oil to Transportation Fuels," Oil Shale Symposium, Colorado School of Mines, April 12-14, 1978.
9. R. F. Sullivan, B. E. Stangeland, H. A. Frumkin, and C. W. Samuel, "Refining Shale Oil," 43rd Midyear Refining Meeting, American Petroleum Institute, Toronto, Canada, May 8-11, 1978.
10. "Process Upgrades Shale Oil to Usable Crude," Chemical and Engineering News, 56 (2), pp 33-34, January 9, 1978.

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TABLE I

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF DEWATERED PARAHO SHALE OIL  
 (PRODUCED BY INDIRECTLY HEATED MODE)

Chevron Research Identification No.	SGQ 5595 <sup>1</sup>	WOW 3394 <sup>2</sup>
Gravity, °API	19.7/18.8	20.2
Sulfur, Wt %	0.63	0.66
Total Nitrogen, Wt %	2.16	2.18
Oxygen, Wt %	1.23	1.16
Arsenic, ppm	33	28
Four Point, ASTM, °F	85	90
Carbon, Wt %		84.30
Hydrogen, Wt %	11.59	11.29
Hydrogen/Carbon Atom Ratio		1.60
Chloride, ppm	11	<0.2
Sodium, ppm	90	5
Ash, Wt % (ASTM D 486)	0.45	0.03
<u>Filter Residue Ash (0.45 µ-Filter)</u>		
Total Solids, ppm		252
Ash, ppm		194
Sediment (Plus Trace Water), Vol %	0.7	0.1
Bromine No.		51
Average Molecular Weight		326
<u>Viscosity, cSt</u>		
122°F	22.47	25.45
210°F	5.055	5.541
300°F	2.299	
Acid Neutralization No., mg KOH/g	3.69	2.3
Hot Heptane Asphaltenes (Including Any Fines), Wt %	1.26	0.17
<u>ASTM D 1160 Distillation, °F</u> (At 10 mm, Corrected to 1 Atm.)		
St/5	332/448	386/456
10/30	499/640	508/659
50	763	776
70/90	857/974	871/995
EP	/999 (90.5%)	/1022 (94%)
% Overhead (Incl. Trap)	90.5	95
% in Trap	Trace	1

<sup>1</sup>EFV dewatered sample.

<sup>2</sup>Pilot plant feed.

TABLE II  
DOE CONTRACT EF-76-C-01-2315  
PROPERTIES OF DEWATERED PARAHO SHALE OIL  
PILOT PLANT FEEDS  
(PRODUCED BY INDIRECTLY HEATED MODE)

Chevron ID No.	WOW 3394	WOW 3511
Gravity, °API	20.2	20.5
Sulfur, Wt %	0.66	0.63
Total Nitrogen, Wt %	2.18	2.15
Oxygen, Wt %	1.16	1.19
Arsenic, ppm	28	26.5
Pour Point, ASTM, °F	90	85
Carbon, Wt %	84.30	84.58
Hydrogen, Wt %	11.29	11.42
Hydrogen/Carbon Atom Ratio	1.60	1.61
Chloride, ppm	<0.2	0.7
Sodium, ppm	5	
Ash, Wt % (ASTM D 486)	0.03	0.04
Iron, ppm	70	68
<u>Filter Residue Ash</u> (0.8 μ-Filter)		
Total Solids, ppm	234	296
Ash, ppm	168	156
<u>Filter Residue Ash</u> (0.45 μ-Filter)		
Total Solids, ppm	252	461
Ash, ppm	194	156
Sediment (Plus Trace Water), Vol %	0.1	0.1
Bromine No.	51	
Average Molecular Weight	326	315
Ramsbottom Carbon, %	2.5	1.85
<u>Viscosity, cs</u>		
At 122°F	25.45	22.48
At 210°F	5.54	5.15
Acid Neutralization No., mg KOH/g	2.3	2.4
Base Neutralization No., mg KOH/g (Equivalent)	38	36
pH	9.2	9.1
Maleic Anhydride No., mg/g	40.6	47.2
Navy Heater Test	No. 1 Rating	
Hot Heptane Asphaltenes (Incl. Any Fines), Wt %	0.17	1.06
<u>ASTM D 1160 Distillation, °F</u> (Corrected to 1 Atm.)		
St/5	386/456	404/506
10/30	508/659	544/670
50	776	772
70/90	871/995	866/1010
EP	/1022	/1088
% Overhead (Excl. Trap)	94	93
% in Trap	1	2.5
% in Flask	5	4.5
<u>TBP Distillation, °F</u> (Simulated by Chromatography)		
St/5	183/347	188/396
10/30	453/610	452/607
50	963	734
70/90	842/963	840/958
95/99	(1053)/-	(1021)/-
C <sub>5</sub> Asphaltenes (Incl. Any Fines), Wt %		1.19
Nitrogen in C <sub>5</sub> Deasphalted Oil, Wt %		2.20

TABLE III

DOE CONTRACT EF-76-C-01-2315  
 ANALYSIS OF PARHO SHALE OIL  
 FOR METALS BY EMISSION SPECTROCHEMICAL METHOD<sup>1</sup>

Identification Size Filter, $\mu$	Pilot Plant Feed, WOW 3394	Filtrate from WOW 3394		Pilot Plant Feed, WOW 3511
	15	0.8	0.45	15
<u>Metals, ppm</u>				
Aluminum	5	-	1.0	13
Boron	0.3	0.2	0.2	0.3
Calcium	20	0.1	0.1	>26
Chromium	0.04	0.1	0.1	-
Cobalt	0.6	0.9	0.9	1.0
Copper	0.1	0.1	0.1	0.2
Iron	>11	>9	70	68
Magnesium	>3	1.9	1.9	>3
Nickel	1.8	2.4	2.3	2.8
Silicon	>15	0.5	0.2	>17
Tin	-	0.7	0.6	-
Titanium	0.8	0.1	0.2	1.4
Vanadium	0.2	0.2	0.2	0.3
Zinc	2.8	4.3	5.3	3.3

<sup>1</sup>These samples also contain 26-28 ppm arsenic that is not removed by filtration.

TABLE IV

DOE CONTRACT EF-76-C-01-2315  
 DISTILLATION OF DEWATERED PARAHO SHALE  
 OIL, SGQ 5996, IN SPINNING BAND COLUMN

Boiling Range, °F	Gravity, °API	Wt %	Vol %
266-400*	38.3	3.6	4.0
400-650	29.4	25.0	26.7
650-800	20.5	22.8	23.0
800-835	18.6	12.0	11.9
835+ Bottoms	10.4	36.6	34.4

\*A drop or two of water observed in the 266-400°F overhead.

TABLE V

DOE CONTRACT EF-76-C-01-2315  
 TBP DISTILLATIONS OF FRACTIONS FROM  
 PARAHO SHALE OIL, SGQ 5996<sup>1</sup>  
 (SIMULATED BY CHROMATOGRAPHY)

TBP Distillation, °F (By Chromatography)	Boiling Range, °F (by Spinning Distillation)			
	266-400	400-650	650-800	800-835
St/5	155/234	325/405	393/629	(381)/635
10/30	275/323	427/488	647/690	731/808
50	352	539	727	833
70/90	383/419	587/633	763/800	854/878
95/99	436/612	651/744	814/869	886/-

<sup>1</sup>TBP distillation of the dewatered whole shale oil is as follows:

TBP, °F (Simulated by  
Chromatography)

St/5	183/397
10/30	453/610
50	736
70/90	842/963
95/99	(1053)/-

TABLE VI  
DOE CONTRACT EF-76-C-01-2315  
PROPERTIES OF FRACTIONS FROM PARAHO SHALE  
OIL, SQG 5996

Boiling Range, °F	Gravity, °API	Sulfur, Wt %	Total Nitrogen, Wt %	Arsenic, ppm	Oxygen, Wt %
	266-400 <sup>1</sup>	38.3	1.08	1.55	6
400-650	29.4	0.74	1.58	38	1.26
650-800	20.5	0.66	2.14	21	0.97
800-835	18.6	0.54	2.07	18	0.78
835+ <sup>2</sup>	10.4	0.52	2.53	35	1.24
	Bromine No.	Pour Point, ASTM, °F	Viscosity, cSt		
			122°F	210°F	300°F
266-400	79		0.905		
400-650	60	10	3.241		
650-800	53	80	19.64	4.49	
800-835	46	110	63	9.04	
835+	Not Detected	95	2966	97.73	15.63
	Group Type, LV% <sup>3</sup>				
	Paraffins	Naphthenes	Aromatics	Sulfur Compounds	
266-400	19.8	38.3	41.9	(Footnote 4)	
400-650	46.7	28.2	21.1	3.9	
650-800	37.4	25.9	3.1	4.5	
800-835	17.0	44.7	34.3	3.9	
835+			No Analysis		

<sup>1</sup>Aniline point of the 266-400°F fraction is 88.3°F. Other fractions are too dark to determine aniline point.

<sup>2</sup>Other properties of the 835°F+ bottoms are as follows: n-C<sub>5</sub> asphaltenes, 8.35 wt %; Hot C<sub>7</sub> asphaltenes, 5.41 wt %; molecular weight, 1227; Ramsbottom carbon, 13.75%. (Note: Fines are included as asphaltenes in these analyses.)

<sup>3</sup>Olefins are not determined by the mass spectrometric group-type method. Reliability of all of the mass spectrometric group-type analyses must be regarded as questionable due to the large quantities of heterocyclic compounds.

<sup>4</sup>Sulfur compounds are not determined by the mass spectrometric method used for the 266-400°F fraction.

TABLE VII  
DOE CONTRACT EF-76-C-01-2315  
COMPARISON OF SHALE OIL FEED AND  
HYDROTREATED LIQUID PRODUCT  
NEAR START-OF-RUN

	Feed (WOW 3394)		Product From ICR 106: 81-4 220-244 Hr. <sup>1</sup>		Product From ICR 104: 75-142 216-240 Hr	
	Wt %	LV%	Wt %	LV%	Wt %	LV%
<u>Boiling Range</u>						
Start-400°F	3.6	4.0	4.0	4.4	5.4	6.0
400-650°F	25.0	26.7	39.6	40.5	38.1	39.2
650°F+	71.4	69.3	56.3	55.1	56.6	54.8
Amount Boiling Above 1000°F (From D 1160)	9.5		4		4	
<u>Inspections of Whole Liquid Product</u>						
Gravity, °API	20.2		34.4		33.9	
Aniline Point, °F			186.3		175.3	
Sulfur, ppm	6600		<10		300	
Total Nitrogen, Wt %	2.18		0.08		0.22	
Pour Point, °F	+90		+80		+80	
Bromine No.	51		2.0		4.4	
Average Molecular Weight	326		281		281	
Hot Heptane Asphaltenes, ppm	1700		82		69	
<u>Group Type, LV%<sup>2</sup> (Mass Spec. 22-Component)</u>						
Paraffins	(16.8)		35.7		36.0	
Naphthenes	(30.4)		40.3		36.1	
Aromatics	(47.5)		23.7		27.4	
Sulfur Compounds	(5.3)		0.3		0.5	
<u>Viscosity, cSt</u>						
100°F			7.719		7.136	
210°F	5.541		3.009		3.541	
<u>ASTM D 1160 Distillation, °F (At 10 mm, Corrected to 1 Atm.)</u>						
St/5	386/456		332/397		313/408	
10/30	508/659		444/575		459/580	
50	776		668		686	
70/90	871/995		776/902		809/977	
95/EP	-/1022 (94%)		960/1013 (97%)		-/1026 (92%)	
Overhead Incl. Trap, %	95		97		97	
Trap, %	1		0		5	

<sup>1</sup>Results suggest loss of some light ends from Run 81-4 at 220-244 hr.

<sup>2</sup>Reliability of the 22-component Mass Spectrometric Group-Type Analysis for material containing this many heterocyclic compounds is questionable. Also, olefins are not detected by this method.

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TABLE VIII

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF START-400°F SHALE OIL AND  
 PRODUCT FROM HYDROTREATED SHALE OIL

	Feed (WOW 3394)	Product from ICR 106: 81-4 at 220- 224 Hr	Product From ICR 104: 75-142 at 216-240 Hr
<u>Start-400°F Inspections</u>	1	2, 3	2
Gravity, °API	38.3	48.1	53.8
Aniline Point, °F		129.7	125.5
Nitrogen, ppm	15,500	470	297
Bromine No.	79	1.6	1.6
<u>Group Type, LV%</u> <u>(Low Mass)</u>			
Paraffins	(19.8) <sup>4</sup>	45.6	53.2
Naphthenes	(38.3) <sup>4</sup>	40.6	36.0
Aromatics	(41.9) <sup>4</sup>	13.8	10.8
<u>TBP Distillation, °F</u> <u>(Simulated by</u> <u>Chromatography)</u>			
St/5	155/234	210/278	82/150
10/30	275/323	287/311	178/238
50	352	336	281
70/90	383/419	366/426	326/395
95/99	436/612	466/601	438/631

<sup>1</sup>Distillation using spinning band column.

<sup>2</sup>Distillation by D 1160-type column.

<sup>3</sup>Results indicate some loss of light ends.

<sup>4</sup>Olefins and heterocyclic compounds are not detected by this analytical method.

TABLE IX

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF 400-650°F SHALE OIL AND  
 PRODUCT FROM HYDROTREATED SHALE OIL

	Feed	Product From ICR 106: 81-4 at 220- 244 Hr	Product From ICR 104: 75-142 at 216-240 Hr
<u>400-650°F Inspections</u>	1	2	2
Gravity, °API	29.4	38.8	38.6
Aniline Point, °F	88.3	152.1	149.6
Sulfur, ppm	7,400	20	700
Nitrogen, ppm	15,800	715	1,290
Bromine No.	60	1.7	3.3
Average Molecular Weight		222	204
<u>Group Type, LV%<sup>3</sup></u>	22-Component	High Mass	High Mass
Paraffins	(46.7)	34.3	34.8
Naphthenes	(28.2)	47.4	44.8
Aromatics	(21.1)	18.3	20.4
Sulfur Compounds	( 2.9)	-	-
Viscosity, cSt at 122°F	3.241	2.090	2.009
<u>TBP Distillation, °F (Simulated by Chromatography)</u>			
St/5	325/405	272/343	240/341
10/30	427/488	378/452	374/451
50	539	506	504
70/90	587/633	559/626	558/623
95/99	651/744	657/737	655/735

<sup>1</sup>Distillation using a spinning band column.

<sup>2</sup>Distillation using a D 1160-type column.

<sup>3</sup>Reliability of Mass Spectrometric Group-Type Analyses are uncertain with material containing as many heterocyclic compounds as the raw shale oil. Also, olefins are not detected by this method.

TABLE X

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF 650°F+ SHALE OIL AND  
 PRODUCT FROM HYDROTREATED SHALE OIL

	Feed	Product From ICR 106: 81-4 at 220-244 Hr	Product From ICR 104: 75-142 at 216-240 Hr
<u>650°F+ Inspections</u>			
Gravity, °API	15.0	31.3	28.4
Aniline Point, °F		208.5	195.6
Sulfur, ppm		10	70
Nitrogen, ppm		(490) (?)	2290
Pour Point, °F		+95	+95
Bromine No.		2.7	4.7
Average Molecular Weight		398	498
Hot Heptane Asphaltenes, ppm		166	110
<u>Group Type, LV% (22-Component Mass Spec.)</u>			
Paraffins		30.8	32.6
Naphthenes		44.1	37.2
Aromatics		23.4	29.9
Sulfur Compounds		1.7	0.2
<u>Viscosity, cSt</u>			
122°F		17.03	19.92
210°F		4.793	5.190
<u>ASTM D 1160 Distillation, °F</u>			
St/5		599/646	626/649
10/30		666/728	629/729
50		786	786
70/90		847/946	845/941
% Overhead Incl. Trap		1025/1041	987/1027
Trap, %		99.0	99.0

TABLE XI

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF PARTIALLY HYDROTREATED SHALE OIL PRODUCT  
 FROM RUN 75-144 PREPARED AT 1.2 LHSV, 745°F USING ICR 106 CATALYST  
 (FEED TO RUN 76-154)

Identification No. Run Hours	SGQ 6040-1	SGQ 6040-2
	42-102	102-150
<u>Inspections</u>		
Gravity, °API	31.0	
Sulfur, Wt %	0.015	0.017
Nitrogen, Wt %	0.72	0.74
Pour Point, °F	+85	
Bromine No.	11	
Hot Heptane Asphaltenes, ppm	130	
<u>ASTM D 1160 Distillation, °F</u>		
St/5	315/422	
10/30	498/605	
50	714	
70/90	826/888	
EP	/1022 (91%)	
Overhead, Incl. Trap, %	93.5	
Trap, %	2.5	
<u>TBP Distillation, °F</u> (Simulated by Chromatography)		
St/5	143/324	
10/30	394/548	
50	665	
70/90	788/913	
95	975	

TABLE XII

DOE CONTRACT EF-76-C-01-2315  
 ARSENIC AND IRON IN  
 AGED CATALYST FROM RUN 81-4  
 AFTER PROCESSING 1060 VOLUMES WHOLE SHALE OIL  
 PER VOLUME OF CATALYST<sup>1</sup>

Layer No.	Wt of Layer, g <sup>2</sup>	Cumulative (From Top), g	Iron, Wt %	Arsenic, Wt %
1	4	4	3.4	3.3
2	5	9	2.5	2.0
3	11	20	1.7	1.3
4	11	31	1.3	0.89
5	10	41		
6	11	52	0.90	0.50
7	10	62		
8	11	73	0.75	0.28
9	10	83		
10	13	96	0.20	0.062
11	12	108		
12	9	117		
13	12	129		
14	5	134	0.043	0.0083

<sup>1</sup>No guard bed was used in this run.

<sup>2</sup>In addition, a total of 5.6 g of fines (smaller than 24 mesh) was distributed throughout the bed. These fines contained 2.3% arsenic and 6.4% iron.

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TABLE XIII

DOE CONTRACT EF-76-C-01-2315  
 COMPOSITION OF GUARD BED REMOVED AT 1950 HOURS

Pilot Plant Run 86-51

210 grams of alumina charged  
 333 grams of alumina plus deposits removed  
 123 grams of deposits accumulated (by difference)  
 Upflow (Layer No. 12 was the first layer contacted)

Layer No.	12 (Bottom)	11	10	9	8	7	6	5	4	3	2	1 (Top)
Weight of Layer, g	41	38	33	42	40	31	20	24	22.5	17.0	12.0	12.5
<u>Composition, Wt %</u>												
Arsenic	1.0	2.7	7.1	7.1	7.1	5.6	4.1	3.5	2.7	2.2	1.9	1.8
Iron	5.5	7.3	14.1	15.0	12.1	5.0	2.9	2.5	1.9	1.4	1.4	1.5
Selenium	0.03	0.09	0.12	0.09	0.06	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
Zinc	0.18	0.79	2.0	2.0	1.6	0.50	0.17	0.12	0.09	0.05	0.04	0.04
Carbon	15.1	18.5	9.25	9.67	11.5	12.9	11.9	11.7	10.9	11.0	10.8	10.4
Sulfur	2.8	3.7	6.93	2.36	2.15	0.51	0.23	0.23	0.17	0.15	0.14	0.30

TABLE XIV

DOE CONTRACT EF-76-C-01-2315  
COMPOSITION OF GUARD BED REMOVED AT 3260 HOURS

Pilot Plant Run 86-51

203 grams alumina charged  
 296 grams alumina plus deposits removed  
 93 grams of deposits accumulated (by difference)  
 Upflow (Layer No. 8 was the first layer contacted)

Layer No.	8 (Bottom)	7	6	5	4	3	2	1 (Top)
Weight of Layer, g	30	41	41	40	39	32	36	37
<u>Composition, Wt %</u>								
Arsenic	1.3	4.1	6.5	5.1	3.0	1.9	1.4	0.95
Iron	2.6	9.9	13.7	10.3	5.1	3.3	2.4	1.9
Selenium	0.03	0.11	0.08	0.04	0.01	0.01	0.01	0.01
Zinc	0.08	0.72	1.0	0.65	0.19	0.07	0.04	0.03
Carbon	10.7	17.3	7.2	8.8	11.3	11.4	11.3	10.9
Sulfur	1.0	4.5	4.5	3.2	1.6	1.6	1.1	1.2

TABLE XV

DOE CONTRACT EF 76-C-01-2315  
HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

Whole Liq. Prod. N, ppm	1325	2800	645	505
RUN	81- 4	81- 4	81- 4	81- 4
FEED	404 3394	WOW 3394	WOW 3394	WOW 3394
RUN HOURS	556.0- 580.0	844.0- 860.0	1252.0-1276.0	1636.0-1660.0
AVG. CAT. TEMP., F.	744.	742.	766.	767.
LHSV	0.59	0.61	0.58	0.59
TOTAL PRESSURE, PSIG	2202.	2203.	2206.	2199.
H2 MEAN PRESS., PSIA	1874.	1824.	1858.	1846.
TOTAL GAS IN, SCF/B	12304.	6522.	7370.	7226.
RECYCLE GAS, SCF/B	10308.	4906.	5195.	5132.
NO LOSS PROD. YIELDS	WT.PC. VOL.PC.	WT.PC. VOL.PC.	WT.PC. VOL.PC.	WT.PC. VOL.PC.
C1	0.42	0.39	0.32	0.27
C2	0.78	0.66	0.80	0.69
C3	0.70	0.59	0.85	0.76
I-C4	0.15 0.25	0.13 0.21	0.19 0.31	0.17 0.28
N-C4	0.44 0.77	0.38 0.61	0.57 0.91	0.53 0.85
C5-400 F.	10.56 13.01	11.19 14.29	12.36 15.23	13.24 16.32
400-650 F.	36.61 39.84	35.08 38.99	39.32 43.78	38.72 43.04
650-800 F.	22.95 24.53	23.18 24.64	22.61 24.27	22.56 24.24
800- EP F.	25.82 26.94	26.29 27.19	21.76 22.93	21.72 22.91
TOTAL C5+	95.96 104.33	95.75 105.03	96.06 106.22	96.26 106.57
ACT./NO LOSS RECOV.	102.16/103.13	100.94/102.54	102.05/103.44	99.84/103.32
H2 CONS(GROSS), SCF/B	1996.	1616.	2175.	2094.
H2 CONS(HCSON), SCF/B <sup>2</sup>	1923.	1558.	2110.	2058.

## ....LIQUID PRODUCT INSPECTIONS....

C5-400 F. PRODUCT<sup>1</sup>

GRAVITY, API	50.6	60.5	51.4	52.2
ANILINE PT., F.	128.2	127.3	128.2	129.0
SULFUR, PPM.	30.0		14.0	25.0
NITROGEN, PPM.	513.00	506.00	125.00	67.00
LOW MASS, LV.PC.				
PARAFFINS	48.7	49.4	47.3	49.4
NAPHTHENES	38.7	36.8	40.7	38.9
AROMATICS	12.6	13.8	12.0	11.6
O.N., F-1 CLEAR			38.0	38.1
BROMINE NO.				0.60
VISCOSITY, CS, 100F.			0.834	
VISCOSITY, CS, 122F.			0.740	
IRP DIST., F.				
ST/ 5	94./ 206.	85./ 190.	99./ 198.	60./ 185.
10/30	240./ 295.	215./ 265.	222./ 285.	213./ 278.
50	356.	330.	324.	322.
70/90	365./ 395.	365./ 401.	356./ 392.	356./ 395.
95/99	406./ 428.	412./ 463.	408./ 467.	409./ 444.

<sup>1</sup>Inspections uncorrected for any C<sub>5</sub>-C<sub>6</sub> measured as gas.<sup>2</sup>Chemical Hydrogen Consumption

TABLE XVI

DOE CONTRACT EF-76-C-01-2315

HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

Whole Liq. Prod. N, ppm	1325	2800	645	505
RUN	81- 4	81- 4	81- 4	81- 4
FEED	40W 3394	WOW 3394	WOW 3394	WOW 3394
RUN HOURS	556.0- 580.0	444.0- 868.0	1252.0-1276.0	1636.0-1660.0
AVG.CAT.TEMP., F.	744.	742.	766.	767.
LHSV	0.58	0.61	0.59	0.59

....LIQUID PRODUCT INSPECTIONS....

400-650 F. PRODUCT

GRAVITY, API	33.7	36.8	37.5	37.4
ANILINE PT., F.	154.3	151.7	153.7	155.5
SULFUR, PPM.	3.7	35.0	14.0	6.0
NITROGEN, PPM.		388.00	453.00	356.00
MOLECULAR WEIGHT	198.	207.	202.	196.
POUR PT.,ASTM, F.	5.	5.		-5.
FREEZE PT.,ASTM, F.				9.
HIGH MASS, LV.PC.				
PARAFFINS	34.5	34.3	33.9	33.6
NAPHTHENES	45.5	45.2	46.5	46.5
AROMATICS	20.0	20.5	19.5	19.8
CLOUD PT.,ASTM, F.			4.	4.
ROMINE NO.			5.00	2.40
VISCOSITY, CS,100F.		3.300	3.030	
VISCOSITY, CS,122F.		2.480	2.390	
VISCOSITY, CS,210F.		1.250	1.180	
TRP DIST., F.				
ST/ 5	328./ 410.	341./ 412.	336./ 396.	322./ 402.
10/30	427./ 485.	429./ 487.	420./ 477.	425./ 482.
50	532.	536.	527.	530.
70/90	583./ 624.	585./ 632.	580./ 625.	581./ 626.
95/99	638./ 669.	649./ 731.	640./ 683.	643./ 763.
ASTM D-86 DIST., F				
ST/ 5	441./ 455.	446./ 464.	431./ 451.	
10/30	465./ 502.	467./ 494.	458./ 486.	
50	552.	521.	514.	
70/90	554./ 583.	553./ 588.	547./ 582.	
95/EP	610./ 616.	597./ 621.	597./ 620.	
LV.PC.OVHD	99.5	99.5	99.5	
COLOR, ASTM				2.0

TABLE XVII  
DOE CONTRACT EF-76-C-01-2315

HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

	1325	2800	645	505
Whole Liq. Prod. N, ppm	1325	2800	645	505
RUN	81- 4	81- 4	81- 4	81- 4
FEED	WOW 3394	WOW 3394	WOW 3394	WOW 3394
RUN HOURS	556.0- 580.0	844.0- 868.0	1252.0-1276.0	1636.0-1660.0
AVG.CAT.TEMP., F.	744.	742.	766.	767.
LHSV	0.58	0.61	0.48	0.59

....LIQUID PRODUCT INSPECTIONS....

650-800 F. PRODUCT

GRAVITY, API	30.7	29.9	31.5	31.6
ANILINE PT., F.	184.9	179.3	188.0	187.3
SULFUR, PPM.	20.0	55.0	13.0	11.0
NITROGEN, PPM.	1600.00	3000.00	960.00	702.00
MOLECULAR WEIGHT	314.	318.	310.	310.
POUR PT., ASTM, F.		70.	75.	70.
22 COMP., LV. PC. <sup>1</sup>				
PARAFFINS			(24.6)	(32.3)
NAPHTHENES			(59.7)	(40.6)
AROMATICS			(15.4)	(26.9)
SULFUR COMPOUNDS			(0.3)	(0.2)
BROMINE NO.			5.50	5.70
VISCOSITY, CS, 100F.		16.790	14.479	
VISCOSITY, CS, 122F.		11.170	9.970	
VISCOSITY, CS, 210F.		3.440	3.160	
TAP DIST., F.				
ST/ 5	534./ 637.	407./ 633.	463./ 635.	479./ 618.
10/30	653./ 687.	652./ 694.	651./ 689.	642./ 688.
50	723.	731.	724.	727.
70/90	757./ 792.	766./ 801.	758./ 793.	767.
95/99	807./ 842.	814./ 844.	806./ 836.	

<sup>1</sup>These 22-component group-type analyses do not appear consistent and may be in error.

TABLE XVIII  
DOE CONTRACT EF-76-C-01-2315

HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

Whole Liq. Prod. N, ppm	1325	2800	645	505
RUN	81- 4	81- 4	81- 4	81- 4
FEED	WOW 3394	WOW 3394	WOW 3394	WOW 3394
RUN HOURS	556.0- 580.0	844.0- 868.0	1252.0-1276.0	1636.0-1660.0
AVG.CAT.TEMP., F.	744.	742.	766.	767.
LHSV	0.58	0.61	0.58	0.59

....LIQUID PRODUCT INSPECTIONS....

800- EP F. PRODUCT

GRAVITY, API	26.9	25.5	28.4	28.6
ANILINE PT., F.	218.0	208.7	225.8	228.6
SULFUR, PPM.	20.0	65.0	<10.0	20.0
NITROGEN, PPM.	1600.00	4500.00	995.00	710.00
MOLECULAR WEIGHT	453.	472.	457.	452.
POUR PT.,ASTM, F.		110.	115.	115.
22 COMP., LV. PC. 1				
PARAFFINS			(32.7)	(20.6)
NAPHTHEMES			(40.1)	(57.6)
AROMATICS			(27.0)	(21.3)
SULFUR COMPOUNDS			(0.2)	(0.5)
BROMINE NO.			6.90	5.80
VISCOSITY, CS,122F.		74.000	41.480	55.169
VISCOSITY, CS,210F.		11.640	4.820	8.770
TGA DISTILLATION				
ST/ 5	159./ 749.	135./ 780.		
10/30	797./ 877.	816./ 884.		
50	918.	919.		
70/90	949./ 993.	950./ 997.		
95/EP	1017.	1043.		
LV.PC.OVHD	98.0	96.3		
TRP DIST., F.				
ST/ 5			518./ 786.	
10/30			801./ 841.	
50			876.	
70/90			923.	
95/99				
ASTM D-1160 DIST.,F				
ST/ 5				809./ 821.
10/30				825./ 833.
50				858.
70/90				900./ 976.
95/EP				1010./1073.
LV.PC.OVHD				99.0

<sup>1</sup>These 22-component group-type analyses do not appear consistent and may be in error.

TABLE XIX  
DOE CONTRACT EF-76-C-01-2315

HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

	1325	2800	645	505
Whole Liq. Prod. N, ppm	81- 4	81- 4	81- 4	81- 4
RUN	WOW 3394	WOW 3394	WOW 3394	WOW 3394
FEED	556.0- 580.0	844.0- 868.0	1252.0-1276.0	1636.0-1660.0
RUN HOURS	744.	742.	766.	767.
AVG.CAT.TEMP., F.	0.58	0.61	0.58	0.59
LHSV				

....LIQUID PRODUCT INSPECTIONS....

WHOLE LIQUID PRODUCT

GRAVITY, API	34.1	34.4	35.7	36.2
ANILINE PT., F.	179.0	173.5	177.8	178.1
SULFUR, PPM.	33.0	145.0	42.0	45.0
NITROGEN, PPM.	1325.00	2800.00	645.00	505.00
MOLECULAR WEIGHT	291.	289.	274.	
POUR PT.,ASTM, F.	55.	80.	80.	85.
BROMINE NO.	3.90	5.90	2.80	3.00
22 COMP., LV. PC.				
PARAFFINS	38.8	33.8	37.8	
NAPHTHENES	38.2	36.7	37.1	
AROMATIC	22.9	29.3	24.2	
SULFUR COMPOUNDS	0.1	0.2	0.9	
VISCOSITY, CS.100F.	7.390	7.429	5.730	5.570
VISCOSITY, CS.210F.	2.150	2.170	1.860	1.880
TRAP DIST., F.				
ST/ 5		134./ 299.	136./ 286.	100./ 285.
10/30		373./ 520.	347./ 491.	349./ 492.
50		638.	604.	604.
70/90		762./ 898.	729./ 890.	724./ 874.
95/99				
ASTM D-1160 DIST.,F				
ST/ 5	329./ 440.	328./ 422.	346./ 428.	302./ 397.
10/30	456./ 573.	451./ 569.	450./ 561.	435./ 550.
50	672.	683.	667.	666.
70/90	788./ 908.	791./ 932.	769./ 913.	765./ 903.
95/EP	1016.	968./1021.	1012.	1003.
LV.PC.OVHD (Excl. Trap)	44.0	95.0	93.0	93.0
Trap, %	2.5	4	6	6
In Flask, %	3.5	1	1	1
Hot Heptane Asphaltenes, ppm	56	108	22	15
Arsenic, Parts per Billion	7	10	3	3

TABLE XX

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL WOW 3394 WITH  
 ICR 106 AT 640 PPM WHOLE PRODUCT NITROGEN

Run	81-4		86-50	
	Run Hours	1660-1684		175-223
Avg. Cat. Temp., °F	775		760	
LHSV	0.60		0.61	
Total Pressure, psig	2199		2199	
H <sub>2</sub> Mean Pressure, psia	1836		1558	
Total Gas In, SCF/B	6970		6583	
Recycle Gas, SCF/B	4923		4769	
<u>No Loss Prod. Yields</u>	<u>Wt %</u>	<u>Vol %</u>	<u>Wt %</u>	<u>Vol %</u>
C <sub>1</sub>	0.35		0.45	
C <sub>2</sub>	0.78		0.59	
C <sub>3</sub>	0.81		0.54	
I-C <sub>4</sub>	0.18	0.30	0.12	0.19
N-C <sub>4</sub>	0.54	0.85	0.32	0.50
C <sub>5</sub> -180°F	5.56	7.29	1.62	2.30
180-300°F			5.25	6.54
300-550°F	30.22	34.32	32.64	36.96
550-650°F	17.52	19.20	17.18	18.81
650°F-End Point	42.55	45.08	39.60	41.95
Total C <sub>5</sub> +	95.97	105.89	96.28	106.55
Act./No Loss Recov.	100.61/103.24		99.14/102.91	
H <sub>2</sub> Cons. (Gross), SCF/B	2047		1814	
H <sub>2</sub> Cons. (Chemical), SCF/B	1986		1782	
<u>Whole Liquid Product Insp.</u>				
Gravity, °API	35.6		35.8	
Sulfur, ppm	18		43	
Nitrogen, ppm	640		630	
<u>TBP Distillation, °F</u> (Simulated by Chromatography)				
St/5	90/278		81/267	
10/30	343/489		331/476	
50	605		586	
70/90	724/867		702/852	
95	966/		911/	
Hot Heptane Asphaltenes, ppm			76	
Ramsbottom Carbon, %			0.10	

TABLE XXI

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL WOW 3394  
 WITH ICR 106 AT 640 PPM WHOLE PRODUCT NITROGEN

Run	81-4	86-50
Run Hours	1660-1684	175-223
Avg. Cat. Temp., °F	775	760
H <sub>2</sub> Mean Press., psia	1836	1558
<u>Liquid Product Inspections</u>		
<u>C<sub>5</sub>-180°F</u>		
Gravity, °API		84.6
Nitrogen, ppm		10
<u>180-300°F</u>		
	1	
Gravity, °API	58.8	57.6
Aniline Point, °F	129.4	124.4
Sulfur, ppm	5.9	60
Nitrogen, ppm	5.1	11.4
Octane, F-1 Clear		51.3
<u>Low Mass, LV %</u>		
Paraffins	52.1	50.4
Naphthenes	38.0	39.3
Aromatics	9.9	10.3
<u>TBF Distillation, °F</u> (Simulated by Chroma- tography)		
St/5	54/148	(48)/179
10/30	174/223	193/233
50	256	256
70/90	280/303	278/296
95/99	311/329	305/(468)

<sup>1</sup>A small amount of 180°F- is present.

TABLE XXII

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL WOW 3394  
 WITH ICR 106 AT 640 PPM WHOLE PRODUCT NITROGEN

Run	81-4	86-50
Run Hours	1660-1684	175-223
Avg. Cat. Temp., °F	775	760
H <sub>2</sub> Mean Pressure, psia	1836	1558
<u>Liquid Product Inspections</u>		
<u>300-550°F</u>		
Gravity, °API	40.9	40.4
Aniline Point, °F	140.7	136.8
Sulfur, ppm	1.5	6.2
Nitrogen, ppm	276	310
Molecular Weight	178	171
Cloud Point, °F		-46
Pour Point, °F	-45	-50
Freeze Point, °F	-32	-31
Smoke Point, mm	21	18
<u>High Mass, LV %</u>		
Paraffins	31.5	32.6
Naphthenes	46.2	42.4
Aromatics	22.2	25.0
<u>FIAM, LV %</u>		
Paraffins Plus Naphthenes	78	75
Aromatics	22	25
Viscosity, cSt, 122°F	1.455	1.451
<u>ASTM D 86 Distillation, °F</u>		
St/5	356/371	345/365
10/30	378/410	379/412
50	439	441
70/90	467/495	470/500
95/EP	505/528	510/521
% Overhead	99	99
<u>TBP Distillation of</u> <u>(Simulated by Chromatography)</u>		
St/5	263/321	270/321
10/30	343/409	342/410
50	455	454
70/90	496/534	495/535
95/99	547/564	548/-
Bromine Number		1.6
Color, ASTM D 156		-16
Copper Strip, ASTM D 130	1A	1B
Doctor Test	Negative	Negative
Flash Point, °F	146	
Oxygen, Wt %	0.007	

TABLE XXIII

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL WOW 3394 WITH  
 ICR 106 AT 640 PPM WHOLE LIQUID PRODUCT NITROGEN

Run	81-4	86-50
Run Hours	1660-1684	175-223
Avg. Cat. Temp., °F	775	760
H <sub>2</sub> Mean Press., psia	1836	1558
<u>Liquid Product Inspections</u>		
<u>550-650°F</u>		
Gravity, °API	34.9	34.7
Aniline Point, °F	169.8	166.7
Sulfur, ppm	1.3	4.9
Nitrogen, ppm	460	805
Molecular Weight	247	246
Cloud Point, °F		+28
Pour Point, °F	+25	+25
Freeze Point, °F	+27	+36
Group Type, LV %		
<u>High Mass</u>		
Paraffins	36.4	
Naphthenes	47.4	
Aromatics	16.2	
<u>22 Component</u>		
Paraffins	43.0	41.8
Naphthenes	36.0	31.6
Aromatics	21.1	26.6
Viscosity, cSt, 122°F	4.10	4.21
Viscosity, cSt, 210°F	1.77	1.735
<u>ASTM D 86 Distillation, °F</u>		
St/5	565/573	566/571
10/30	576/580	573/576
5	584	581
70/90	590/603	587/599
95/EP	609/614	604/616
% Overhead	99.5	99.5
<u>TBP Distillation, °F</u>		
St/5	373/549	517/533
10/30	561/587	564/587
50	608	609
70/90	628/650	629/649
95/99	657/667	656/664
Bromine Number		2.2
Copper Strip, ASTM D 130	1A	
Doctor Test	Negative	
Oxygen, Wt %	0.032	

TABLE XXIV

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL WOW 3394 WITH  
 ICR 106 AT 640 PPM WHOLE LIQUID PRODUCT NITROGEN

Run No.	81-4	86-50
Run Hours	1660-1684	175-223
Avg. Cat. Temp., °F	775	760
H <sub>2</sub> Mean Press., psia	1836	1558
<u>Liquid Product Inspections</u>		
<u>650°F+ Product</u>		
Gravity, °API	29.3	29.3
Aniline Point, °F	212.7	208.3
Sulfur, ppm	20	20
Nitrogen, ppm	890	1405
Molecular Weight	391	379
Pour Point, °F	+105	+110
Group Type, LV %		
<u>22 Component</u>		
Paraffins	31.5	27.8
Naphthenes	43.2	35.2
Aromatics	25.3	37.1
<u>Viscosity, cSt</u>		
122°F	22.14	19.98
210°F	5.709	5.33
<u>ASTM D 1160 Distillation, °F</u> (Corrected to 1 Atm.)		
<u>10 mm</u>		
St/5	706/714	
10/30	721/754	
50	799	
70/90	859/931	
95/EP	976/1011	
% Overhead	95.5	
% In Trap	0	
% In Flask	4.5	
<u>2 mm</u>		
St/5	684/716	693/706
10/30	729/772	718/754
50	807	795
70/90	868/950	849/940
95/EP	993/1077	975/1054
% Overhead	99.5	97.5
% In Flask	0.5	2.5
Hot Heptane Asphaltenes, ppm	345	600
Ramsbottom Carbon, %	0.15	0.26

TABLE XXV

DOE CONTRACT EF-86-C-01-2315

HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

	512		625		771	
Whole Liquid Product, N, ppm	512		625		771	
RUN	86- 51		86- 51		86- 51	
FEED	WOW 3394		WOW 3394		WOW 3394	
RUN HOURS	465.0- 489.0		537.0- 561.0		873.0- 897.0	
AVG.CAT.TEMP., F.	763.		765.		785.	
LHSV	0.59		0.60		0.60	
TOTAL PRESSURE, PSIG	2355.		2348.		2345.	
H2 MEAN PRESS., PSIA	1801.		1789.		1653.	
TOTAL GAS IN, SCF/B	9993.		9319.		9329.	
RECYCLE GAS, SCF/B	8163.		7436.		7388.	
NO LOSS PROD.YIELDS	WT.PC.	VOL.PC.	WT.PC.	VOL.PC.	WT.PC.	VOL.PC.
C1	0.50		0.45		0.63	
C2	0.68		0.61		0.79	
C3	0.61		0.54		0.85	
I-C4	0.13	0.21	0.11	0.18	0.20	0.33
N-C4	0.35	0.55	0.29	0.46	0.58	0.93
C5-180 F.	58.60	70.34	1.31	1.85	1.87	2.66
180-300 F.			5.75	7.22	6.44	8.08
300-550 F.			33.09	37.56	33.98	38.37
550-650 F.			17.65	19.29	18.55	20.20
650- EP F.			37.39	38.00	38.56	41.00
TOTAL C5+	96.00	108.35	96.38	106.95	95.40	106.09
ACT./NO LOSS RECOV.	101.73/102.91		100.86/103.01		100.33/103.08	
H2 CONS(GROSS), SCF/B	1830.		1883.		1941.	
H2 CONS(HCSON), SCF/B*	1786.		1849.		1894.	

....LIQUID PRODUCT INSPECTIONS....

C5-180 F. PRODUCT

GRAVITY, API 83.9 83.9

180-300 F. PRODUCT

GRAVITY, API 59.1 59.0  
 ANILINE PT., F. 129.6  
 O.N., F-1 CLEAR 48.7  
 NITROGEN, PPM. 2.70 10.00  
 LOW MASS. LV.PC.  
 PARAFFINS 55.2  
 NAPHTHENES 37.6  
 AROMATICS 7.3  
 TBP DIST., F.  
 ST/ 5 81./ 158. 80./ 175.  
 10/30 188./ 223. 192./ 223.  
 50 255. 254.  
 70/90 278./ 299. 275./ 295.  
 95/99 304./ 315. 302./ 320.

\*Chemical Hydrogen Consumption

TABLE XXVI

DOE CONTRACT EF-76-C-01-2315

HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

	512	625	771
Whole Liq. Prod. N, ppm	86- 51	86- 51	86- 51
RUN	WOW 3394	WOW 3394	WOW 3394
FEED	465.0- 489.0	537.0- 561.0	873.0- 897.0
RUN HOURS	763.	765.	785.
AVG.CAT.TEMP., F.	0.59	0.60	0.60
LHSV			

....LIQUID PRODUCT INSPECTIONS....

300-550 F. PRODUCT

GRAVITY, API	40.8	39.9
ANILINE PT., F.		140.2
NITROGEN, PPM.	400.00	585.00
HIGH MASS.LV.PC.		31.5
PARAFFINS		40.5
NAPHTHENES		28.0
AROMATICS		
POUR PT.,ASTM, F.	-40.	17.
SMOKE PT.,ASTM, MM	20.	-32.
FREEZE PT.,ASTM, F.	-36.	
TBP DIST., F.		
ST/ 5	276./ 320.	270./ 325.
10/30	341./ 403.	345./ 409.
50	453.	454.
70/90	493./ 533.	495./ 530.
95/99	546./ 625.	542./ 556.
ASTM D-86 DIST.,F		
ST/ 5		358./ 392.
10/30		394./ 421.
50		446.
70/90		472./ 498.
95/EP		508./ 536.
LV.PC.OVHD		99.0

550-650 F. PRODUCT

GRAVITY, API	34.4	33.8
ANILINE PT., F.		162.7
NITROGEN, PPM.	625.00	910.00
POUR PT.,ASTM, F.	25.	
MOLECULAR WEIGHT		242.
TBP DIST., F.		
ST/ 5		454./ 552.
10/30		562./ 588.
50		608.
70/90		629./ 650.
95/99		656./ 665.

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RFS

TABLE XXVII

DOE CONTRACT EF-76-C-01-2315HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

	86- 51	86- 51	86- 51
RUN FEED	WOW 3394	WOW 3394	WOW 3394
RUN HOURS	465.0- 489.0	537.0- 561.0	673.0- 897.0
AVG. CAT. TEMP., F.	763.	765.	785.
LHSV	0.59	0.60	0.60

## .....LIQUID PRODUCT INSPECTIONS.....

## 650- EP F. PRODUCT

GRAVITY, API	29.9	29.9	30.1
ANILINE PT., F.			202.6
NITROGEN, PPM.		845.00	1055.00
MOLECULAR WEIGHT		364.	356.
ASTM D-1160 DIST., F			
ST/ 5		665./ 691.	680./ 706.
10/30		706./ 754.	718./ 743.
50		794.	776.
70/90		845./ 940.	829./ 910.
95/EP		988./1055.	958./1038.
LV.PC.OVHD		97.0	99.0

Whole Liquid Product Properties

Gravity, °API	37.2	36.5	37.3
Aniline Point, °F		168.9	165.8
Sulfur, ppm		20	10
Nitrogen, ppm	512	625	771
Molecular Weight		274	271
Pour Point, °F		+75	+70
Ramsbottom Carbon, %		0.09	0.09
Bromine Number		2.5	3.0
Hot Heptane Asphaltenes, ppm		26	36
<u>Group Type, LV %</u>			
Paraffins		33.4	34.3
Naphthenes		35.9	32.7
Aromatics		30.7	32.9

Viscosity, cSt

100°F	4.312	3.802
210°F	1.7012	1.454

TBP Distillation (Simulated by Chromatography)

ST/5	56/244	66/245
10/30	310/459	306/456
50	575	565
70/90	687/839	668/814
95/99	865/980	868/1037

TABLE XXVIII

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106  
 AT 200 PPM WHOLE PRODUCT NITROGEN

Run	86-51			
Feed	WOW 3511			
Run Hours	3483-3507			
Average Catalyst Temperature, °F	772			
LHSV	0.31			
Total Pressure, psig	2348			
H <sub>2</sub> Mean Pressure, psia	1709			
Total Gas In, SCF/Bbl	9786			
Recycle Gas, SCF/Bbl	7784			
	Wt %		Vol %	
<u>No Loss Product Yields</u>				
C <sub>1</sub>	0.65			
C <sub>2</sub>	0.88			
C <sub>3</sub>	0.87			
iC <sub>4</sub>	0.19		0.30	
nC <sub>4</sub>	0.49		0.78	
C <sub>5</sub> -180°F	2.21		3.12	
180-350°F	12.89		15.78	
350-650°F	50.76		56.44	
650°F-EP	29.65		31.76	
Total C <sub>5</sub> +	95.52		107.10	
Act./No Loss Recov.	103.46		103.18	
H <sub>2</sub> Consumption (Gross), SCF/bbl	2001			
H <sub>2</sub> Consumption (Chemical), SCF/bbl	1948			
Boiling Range	Whole Liquid Product	180-350°F	350-650°F	650°F+
<u>Product Inspections</u>				
Gravity, °API	38.5	54.6	37.5	31.3
Aniline Point, °F	167.0	125.9	149.3	4.0
Sulfur, ppm	22	10	4.2	325
Nitrogen, ppm	205	8.3	155	
Molecular Weight	274			
Ramsbottom Carbon, %	0.05			
Hot Heptane Asphaltenes, ppm	46			
Arsenic, ppb	<5			
TBP Distillation, °F (Simulated by Chromatography)				
St/5	74/240	78/184	323/365	464/665
10/30	292/437	199/245	384/442	678/727
50	539	280	500	781
70/90	646/802	313/347	562/626	837/924
95/99	860/979	356/371	642/660	973/-
Octane No., F-1 Clear		46.6	-5	+105
Pour Point, °F				

TABLE XXIX

DOE CONTRACT EF-76-C-01-2315  
 DISTRIBUTION OF C<sub>5</sub>-180°F PRODUCT FROM  
 WHOLE SHALE OIL HYDROTREATING WITH  
 ICR 106 CATALYST  
 PILOT PLANT RUN 86-51

Time Onstream, Hours	537-561	873-897
Average Catalyst Temperature, °F	765	785
Composition, LV % of C <sub>5</sub> -180°F (By Chromatography)		
Propane	0.6	0.5
Isobutane	0.9	0.6
n-Butane	4.7	3.3
Isopentane	17.8	17.2
n-Pentane	27.9	25.5
2,2-Dimethylbutane	0.1	0.1
2,3-Dimethylbutane	1.2	1.1
2-Methylpentane	10.6	10.8
3-Methylpentane	6.0	5.9
n-Hexane	14.2	17.0
Isoheptanes	5.0	5.2
n-Heptane	1.5	1.4
<b>Total Paraffins</b>	<b>90.5</b>	<b>88.6</b>
Cyclopentane	1.5	1.4
Methylcyclopentane	3.4	4.4
Cyclohexane	2.4	3.1
Dimethylcyclopentanes, Ethylcyclopentane	0.7	1.4
Methylcyclohexane	-	0.1
C <sub>8</sub> Naphthenes	0.6	0.4
<b>Total Naphthenes</b>	<b>8.6</b>	<b>10.8</b>
Benzene	0.9	0.6
<b>Total Aromatics</b>	<b>0.9</b>	<b>0.6</b>
Octane Number, F-1 Clear (Estimate)	(70.4)	(68.9)
Isopentane/n-Pentane	0.6	0.7
Iso-C <sub>6</sub> /n-Hexane	1.3	1.1

TABLE XXX

DOE CONTRACT EF-76-C-01-2315  
INSPECTION OF BLENDS OF HYDROTREATED SHALE OIL

Blend Identification	SGQ 6127	WOW 3497
<u>Inspections of Blend</u>		
Gravity, °API	35.8	36.5
Aniline Point, °F	173.7	170.3
Sulfur, ppm	25	11
Nitrogen, ppm	850	616/575
Oxygen, ppm		50
Molecular Weight (Avg.)		270
Pour Point, °F		+75
Ramsbottom Carbon, %	0.09	0.11
Hot Heptane Asphaltenes, ppm		37
TBP Distillation, °F (Simulated by Chromatography)		
St/5	86/273	80/252
10/30	340/485	316/460
50	596	573
70/90	712/857	681/833
95/99	916/	890/
<u>ASTM D 86/D 1160</u>		
<u>Distillation, °F (at 2 mm)</u>		
St/5		213/313
10/30		354/522
50		615
70/90		710/822
95/EP		949/970
% Overhead		97
Carbon, Wt %		84.73
Hydrogen, Wt %		13.85
<u>Viscosity, cSt</u>		
at 122°F		3.229
at 210°F		1.795
Ash, ASTM D 486, Wt %		<0.003
Bromine Number		2.2
<u>Neutralization Number,</u>		
<u>mg KOH/g</u>		
Acid Number		<0.1
Base Number		1.0
pH		7.9



TABLE XXXII

DOE CONTRACT EF-76-C-01-2315  
 650°F+ HYDROTREATED SHALE OIL FEEDS FOR  
 CATALYTIC CRACKING AND HYDROCRACKING TESTS

Downstream Processing Test Type	Catalytic Cracking			Hydrocracking
	ASL 1864 Bottoms (CCL-5406)	SGQ 6115 (CCL-5407)	SGQ 6139 (CCL-5420)	SGQ 6152
Identification No.				
<u>Inspections</u>				
Gravity, °API	30.9	30.8	28.4	29.6
Aniline Point, °F	212.3	210.4	209.1	208.4
Sulfur, ppm	<10	<10	10	<10
Nitrogen, ppm	385	870 <sup>1</sup>	1300	885
Oxygen, ppm	150	200	60	150
Molecular Weight (Avg.)	365	383	384	359
Pour Point, °F	+105	+100	+105	+105
Ramsbottom Carbon, %	0.12	0.19	0.17	0.15
Hot Heptane Asphalt., ppm <sup>2</sup>	477	756	309	647
<u>Group Type, LV % (22 Component)</u>				
Paraffins			25.6	(20.6)
Naphthenes			46.4	(46.1)
Aromatics			28.0	(33.3)
Arsenic, Parts Per Billion			11	7
<u>TBP Distillation (Simulated by Chromatography), °F</u>				
St/5	631/661	535/666	601/673	576/663
10/30	674/721	682/734	691/748	676/722
50	774	787	806	776
70/90	830/917	841/933	865/959	831/915
95/-	975/-	1000/-	1024/-	964/-
<u>ASTM D 1160 Distillation, °F (2 mm)</u>				
St/5			691/717	550/716
10/30			727/760	729/756
50			811	802
70/90			860/954	842/916
95/EP			1015/1059	944/1073
% Overhead			97	99
Carbon, %		86.40	86.14	86.93
Hydrogen, %	13.22	13.67	12.08	13.07
<u>Viscosity, cs</u>				
At 122°F	16.37		22.03	18.85
At 210°F	4.79		5.70	4.97
At 300°F	2.25		2.59	

<sup>1</sup>870 ppm nitrogen is weighted average of components of blend.

<sup>2</sup>Because of the waxy nature of these samples, it is difficult to obtain reproducible results on the hot heptane asphaltene test.

TABLE XXXIII

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF NAPHTHA FRACTIONS FROM  
 HYDROTREATED SHALE OIL BLENDS

Naphtha ID No.	SGQ 6147	SGQ 6148	SGQ 6170	ASL 2068		SGQ 6149	SGQ 6137
				Cut 1	Cut 2		
Distilled from	← WOW 3497 →						SGQ 6127
Boiling Range, °F	St-180 <sup>1</sup>	180-300	180-350	300-350	350-400	300-400	St-400
Yield, LV % of Whole Liquid Blend	0.51	7.84	13.05	5.18	5.07	11.05	14.56
<u>Inspections</u>							
Gravity, °API	81.5	59.9	55.5	49.4	44.9	46.9	53.4
Aniline Point, °F		130.0	127.3	124.7	131.5	128.4	129.5
Sulfur, ppm	7.3	6.0	2.7			4.3	25
Nitrogen, ppm	1.8	15	56	148	258	204	148
Oxygen, ppm		(60)				(50)	
Freeze Point, °F				-84	-72		
<u>Octane No.</u>							
F-1 Clear		50.7	46.0				41.6
F-2 Clear		50.7					
<u>Group Type, LV %<sup>2</sup></u>							
Paraffins	86.8	57.4	50.8	44.7	45.1	44.4	49.9
Naphthenes	12.1	35.5	38.3	41.7	37.1	40.2	38.6
Aromatics	1.0	7.1	10.8	13.6	17.8	15.4	11.4
Olefins	0.1						
<u>ASTM D 86 Distillation, °F</u>							
St/5		174/189	188/220	300/305	358/360	314/324	170/180
10/30		209/227	231/259	306/314	362/365	325/387	226/271
50		242	279	318	368	346	301
70/90		259/278	299/320	324/328	372/380	355/374	328/357
95/EP		287/361	328/362	333/386	385/423	378/418	368/425
% Overhead (Incl. Trap)		99	99	99.5	99.5	99.5	99.0
% in Trap		-	-	-	-	-	0.5
<u>TBP Distillation, °F (Simulated by Chromatography)</u>							
St/5		54/144	78/157	197/270	310/342	215/285	56/158
10/30		161/219	193/247	283/314	349/368	300/335	197/266
50		255	282	337	385	361	312
70/90		280/306	312/344	353/368	400/419	386/410	353/393
95/99		315/477	353/465	372/563	426/443	421/478	406/483

<sup>1</sup>Uncorrected for C<sub>3</sub>-C<sub>4</sub>'s measured as gas.

<sup>2</sup>St-180°F group type measured by chromatography, other fractions by mass spectrometry (low mass).

TABLE XXXIV

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF NARROW BOILING MIDDLE DISTILLATE  
 FRACTIONS FROM HYDROTREATED SHALE OIL BLEND SQQ 6127

Fraction No.	1	2	3	4	5
Boiling Range, °F	400-450	450-500	500-550	550-600	600-650
Yield, LV % of Whole Shale Oil Blend	8.8	8.7	8.3	9.5	11.8
Gravity, °API	41.1	38.4	36.1	35.7	33.2
Aniline Point, °F	131.1	140.5	147.7	162.9	170.3
Nitrogen, ppm	485	610	790	735	990
Freeze Point, °F	-43	-11	+1	+25	+49
Pour Point, °F	-60	-20	-5	+20	+45
Cloud Point, °F	-54	-20	-4	+20	+45
<u>Group Type, LV %</u>					
Paraffins	28.5	32.6	32.1		
Naphthenes	45.7	40.8	44.1		
Aromatics	25.8	26.6	23.9		

TABLE XXXV  
DOE CONTRACT EF-76-C-01-2315  
PROPERTIES OF MIDDLE DISTILLATE FRACTIONS FROM  
HYDROTREATED SHALE OIL BLENDS

ID No.	SGQ 6198	SGQ 6150	SGQ 6151	SGQ 6173	SGQ 6138
Distilled From	WOW 3497	WOW 3497	WOW 3497	WOW 3497	SGQ 6127
Boiling Range, °F	350-650	400-550	550-650	400-650	400-650
Yield, LV % of Whole Liquid Blend	49.29	25.04	19.07	44.11	47.05
<u>Inspections</u>					
Gravity, °API	37.4	38.4	34.2	36.4	36.6
Aniline Point, °F	149.1	142.1	166.4	153.7	153.0
Sulfur, ppm	40				
Nitrogen, ppm	540	480	620	510	815
Oxygen, ppm	80	100	90		
Freeze Point, °F	+10	-21	+35	+18	+17
Pour Point, °F	0	-25	+30	+10	+20
Cloud Point, °F		-25	+30	+10	
Smoke Point, mm		17	15		
<u>Group Type, LV %</u>					
Paraffine	28.7	31.9			37.4
Naphthenes	44.1	42.7			39.3
Aromatics	27.2	25.4			23.3
<u>ASTM D 86 Distillation, °F</u>					
St/5	400/427			441/462	430/447
10/30	437/467			470/497	457/488
50	508			527	522
70/90	548/586			559/591	560/598
95/EP	597/618			603/618	610/629
% Overhead	99.5			99.5	99.5
<u>TBP Distillation, °F (Simulated by Chromatography)</u>					
St/5	327/373	326/396	441/542		309/399
10/30	392/461	415/453	556/591		425/489
50	519	488	616		542
70/90	580/631	518/552	639/662		596/649
95/99	645/663	563/584	669/678		665/685
<u>Viscosity, cs</u>					
At 100°F	2.818				
At 210°F	1.138				
Thermal Stability, Reflectance					
After 90 Min. at 300°F, %				(55)*	(52.5)*
Filter Residue After 5 Hr O <sub>2</sub> Blow, ppm					
				(26)*	(16)*
Color, As Is					
				1.0	1.0
Color After 6 Hr O <sub>2</sub> Blow					
				2.5	2.5

\*Overheating in batch distillation may adversely affect product stability.

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TABLE XXXVI

DOE CONTRACT EF-76-C-01-2315  
 COMPARISON OF STRAIGHT-RUN VGO WITH  
 650°F+ HYDROTREATED SHALE OIL

Inspections	Arabian VGO	Hydrotreated Shale Oil		Sumatran VGO
		1	2	
Gravity, °API	22.2	29.3	29.6	31.9
Aniline Point, °F	183	213	208	217
Sulfur	2.46 Wt %	20 ppm	<10 ppm	1300 ppm
Nitrogen, ppm	780	890	885	385
Oxygen, ppm	1100		150	1000
Pour Point, °F	100	105	105	110
<u>Group Type, LV %</u>				
Paraffins	18	27	21	44
Naphthenes	22	49	46	31
Aromatics	33	24	33	23
Sulfur Comp.	27			2
Hot Heptane Asphaltenes, ppm	120	345	647	132
<u>ASTM D 1160 Distillation, °F</u>				
St/5	707/738	684/716	550/716	550/640
10/30	752/795	729/772	729/756	659/756
50	858	807	802	809
70/90	917/978	868/950	842/916	859/940
95/EP	998/1038	993/1077	944/1073	983/1019

<sup>1</sup>Small Pilot Plant - Middle-of-Run Product

<sup>2</sup>Large Pilot Plant - Blend of Product Throughout Run

TABLE XXXVII

DOE CONTRACT EP-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL, WOV 3394 WITH ICR 106  
 AT 1825 PSIA H<sub>2</sub>, 767°F, RUN 81-4  
 COMPARISON OF PRODUCT PROPERTIES AT SEVERAL NITROGEN LEVELS  
 (1.3-640 PPM NITROGEN)

Liquor Product Nitrogen, ppm LHSV	640		505		30		1.3	
	Wt %	LV %						
Time Onstream, Hr	1660-1684	0.6	1636-1660	0.6	1348-1396	0.3	1528-1576	0.2
Distribution, % of Whole								
Liquid Product								
St-300°F	5.0	5.7					5.0	5.5
300-550°F	31.8	32.8					42.5	42.9
550-650°F	18.4	18.4					20.4	20.1
650°F+	44.8	43.1					32.1	31.5
Whole Liquid Product Properties								
Gravity, °API	35.6		36.2		37.7		39.8	
Aniline Point, °F	18		178		183		42	
Sulfur, ppm	640		31		(95)		1.3	
Nitrogen, ppm			505		30			
Pour Point, °F			+85		+80			
Average Molecular Weight			15		269		35	
Hot Heptane Asphaltene, ppm			0.06		0.06		0.02	
Ramsbottom Carbon, %			3		3			
Arsenic, Parts per Billion								
Group Type, LV % (Weighted Average of Components)								
Paraffins	34.8		32.7		35.2		36.2	
Naphthenes	42.6		46.5		48.2		52.7	
Aromatics	22.6		20.8		16.5		11.1	
Viscosity, cSt at 122°F			5.57		5.18			
Viscosity, cSt at 210°F			1.88		1.67			
TBP Distillation, °F								
(Simulated by Chromatography)								
St/5	90/278		100/285		134/283		179/286	
10/30	343/489		349/492		342/476		335/450	
50	605		604		580		545	
70/90	724/867		724/874		690/842		644/796	
95/99	926/-		938/-		906/-		955/970	

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TABLE XXXVIII

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL WOV 3394  
 WITH ICR 106 AT 1825 PSIA H<sub>2</sub>, 767°F  
 COMPARISON OF PRODUCT PROPERTIES AT 640 PPM AND 1.3 PPM PRODUCT NITROGEN  
 PILOT PLANT RUN 81-4 AT 1660-1684 AND 1528-1576 HR

	St-300		300-550		550-650	
	640	1.3	640	1.3	640	1.3
Boiling Range, °F	58.8	56.6	40.9	41.7	34.9	37.3
Whole Product Nitrogen, ppm	129.4	132.0	140.7	149.5	169.8	184.6
Product Inspections	5.9		1.5		1.3	1.3
Gravity, API	5.1	0.8	276	1.3	460	
Aniline Point, °F			178		247	
Sulfur, ppm			21	23	+27	
Nitrogen, ppm			-32	-37		
Molecular Weight			31.5	28.8	43.0	40.4
Smoke Point, mm			46.2	60.7	36.0	46.3
Freeze Point, °F			22.2	10.5	21.1	13.2
Group Type, LV %	52.1	43.9				
Paraffins	38.0	51.1				
Naphthenes	9.9	5.0				
Aromatics						
TBP Distillation, °F						
(Simulated by Chromatography)						
St/5	54/148	137/196	263/321	261/318	373/549	342/546
10/30	174/223	213/248	343/409	340/402	561/587	557/585
50	256	264	455	448	608	608
70/90	280/303	280/297	496/534	492/531	628/650	629/657
95/99	311/329	302/312	547/564	546/590	657/667	700/-

TABLE XXXIX

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL WOW 3394  
 WITH ICR 106 AT 1825 PSIA H<sub>2</sub>, 767°F  
 COMPARISON OF PROPERTIES OF 650°F+ AT 640 PPM, 30 PPM,  
 AND 1.3 PPM PRODUCT NITROGEN  
 PILOT PLANT RUN 81-4

Nitrogen in Whole Liquid Product, ppm Time Onstream, Hr	640 1660-1684	30# 1348-1396	1.3 1528-1576
<u>650°F+ Inspections</u>			
Gravity, °API	29.3	32.6	36.4
Aniline Point, °F	212.7	222.7	236
Sulfur, ppm	20	13	36 (?)
Nitrogen, ppm	890	50	1.4
Pour Point, °F	+105		+100
Average Molecular Weight	391	401	384
Hot Heptane Asphaltenes, ppm	345	151	129
Ramsbottom Carbon, %	0.15		0.06
<u>Group Type, LV %</u>			
Paraffins	31.5	36.5	42.1
Naphthenes	43.2	47.7	46.2
Aromatics	25.3	15.8	11.7
<u>Viscosity, cSt</u>			
122°F	22.14		12.84
210°F	5.71		4.18
<u>TBP Distillation, °F</u>			
St/5	462/672		563/678
10/30	689/743		695/740
50	796		782
70/90	851/941		829/902
95/99	(1010)/-		951/-

\*Inspections at 1348-1396 hr calculated using inspections measured for 650-800°F and 800°F+ fractions.

TABLE XL

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF WHOLE SHALE OIL, WOV 3394 WITH ICR 106 AT 1825 PSIA H<sub>2</sub>, 767°F  
 COMPARISON OF PRODUCT PROPERTIES AT 505 PPM AND 30 PPM PRODUCT NITROGEN  
 PILOT PLANT RUN 81-4 AT 1636-1660 AND 1346-1398 HR

	St-400		400-650		650-800		800+	
	505	30	505	30	505	30	505	30
Boiling Range, °F	129.0	131.2	155.5	160.5	187.3	202.9	228.6	240.5
Whole Product Nitrogen, ppm	52.2	51.1	37.4	38.4	31.6	33.0	28.6	32.1
Wt % of Liquid Product	25	18	6	11	11	7	20	20
Vol % of Liquid Product	67	15	356	21	702	52	710	(48)
Product Inspections	38.1	36.7						
Gravity, °API								
Aniline Point, °F								
Sulfur, ppm								
Nitrogen, ppm								
Octane No., F-1 Clear								
Pour Point, °F								
Cloud Point, °F								
Molecular Weight								
Group Type, LV %								
Paraffins	49.4	44.8	33.6	32.8	32.3	35.0	20.6	37.8
Naphthenes	38.9	45.8	46.5	49.0	40.6	45.4	57.6	50.2
Aromatics	11.6	9.3	19.8	18.2	27.2	19.7	21.8	12.0
TBP Distillation, °F								
(Simulated by Chromatography)								
St/5	60/185	88/194	322/402	216/349	479/613	468/643		394/793
10/30	213/278	224/283	425/482	389/464	642/688	659/700		806/843
50	322	322	530	525	727	732		875
70/90	356/395	358/398	581/626	585/644	767/(936)	766/800		917/(1023)
95/99	409/444	414/446	643/763	669/752		815/877		

TABLE XLI

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF DEWATERED WHOLE PARAHO SHALE OIL AND  
 SHALE OIL COKER DISTILLATE FROM PARAHO SHALE OIL

Type Chevron ID No.	Whole Shale Oil	Shale Oil Coker
	WOW 3511	Distillate WOW 3524
<u>Inspections</u>		
Gravity, °API	20.5	32.7
Aniline Point, °F		102.5
Sulfur, Wt %	0.63	0.61
Total Nitrogen, Wt %	2.15	1.75
Oxygen, Wt %	1.19	0.94
Arsenic, ppm	26.5	6.3
Pour Point, ASTM, °F	85	25
Carbon, Wt %	84.58	85.00
Hydrogen, Wt %	11.42	12.36
Hydrogen/Carbon	1.61	1.73
Chloride, ppm	0.7	1.7
Iron, ppm	68	0.4
Average Molecular Weight	315	228
Ramsbottom Carbon, %	1.85	0.17
<u>Group Type, LV % (FIAM)</u>		
Paraffins and Naphthenes		36
Aromatics		41
Olefins		23
<u>Viscosity, cSt</u>		
At 100°F		2.92
At 122°F	25.45	2.13
Acid Neutralization No., mg KOH/g	2.4	1.1
Base Neutralization No., mg KOH/g	36	31
pH	9.1	9.0
Maleic Anhydride No., mg/g	47.2	27
<u>ASTM Distillation, °F (Corrected to 1 Atm.)</u>		
	D 1160	D 86/D 1160
	(Run at 2 mm)	
St/5	404/506	186/295
10/30	544/670	344/467
50	772	542
70/90	866/1010	595/655
EP	/1088	/684
% Overhead (Excl. Trap)	93	95
% in Trap	2.5	5
% in Flask	4.5	0
<u>TBP Distillation (Simulated by Chromatography), °F</u>		
St/5	188/396	51/236
10/30	452/607	298/433
50	734	514
70/90	840/958	584/652
95/99	(1021)/-	675/723

TABLE XLIII

DOE CONTRACT NO. EF-76-C-01-2315  
 PROPERTIES OF PARAHO SHALE OIL COKER DISTILLATE  
 WOW 3524 AND NARROW BOILING FRACTIONS FROM WOW 3524

Fraction No.	1	2	3	4	5	3 and 4
Boiling Range, °F	ST-180	180-350	350-500	500-650	650°F+	350-650
Wt % of WOW 3524	2.6	13.2	29.5	43.5	11.8	73.0
LV % of WOW 3524	2.6	14.6	30.0	41.9	10.9	71.9
<u>Inspections</u>						
Gravity, °API	75.7	50.8	35.6	26.7	20.5	
Aniline Point, °F		90.9	92.9	104.0		
Sulfur, Wt %	0.22	0.78	0.63	0.56	0.64	
Nitrogen, Wt %	0.115	0.75	1.34	2.08	2.58	
Oxygen, Wt %	0.31	0.39	0.92	0.90	0.76	
Pour Point, °F			-35	+25		+10
Arsenic, ppm	0.02	1.5	2.6	3.2	7.2	
Maleic Anhydride No. (Diene Value), mg/g		12	29	40	57	
Bromine Number	85	68	47	35	36	
Group Type, LV % (FIAM)						
Paraffins + Naphthenes		41	37	8		410/430
Aromatics		40	28	43		439/473
Olefins		19	35	49		514
ASTM D 86 Distillation, °F						550/585
St/5						559/612
10/30						98.5
50						
70/90						
95/EP						
% Overhead						
TBP Distillation, °F						
(Simulated by Chromatography)						
St/5						324/366
10/30						383/443
50						505
70/90						570/636
95/99						653/674

TABLE XLIII

DOE CONTRACT EF-76-C-01-2315  
 WHOLE PARAHO SHALE OIL COKING YIELDS

Coking Run Run No. Quantity Charged Maximum Column Top Temp., °F	Shell Still		Laboratory Glass Coking Still					
	CPD 358-2		306		315		318	
	2 Drums (110 Gallons) 843 Lb 450-550	LV %	250 Grams	250 Grams	250 Grams	250 Grams	250 Grams	250 Grams
	Wt %	LV %	Wt %	LV %	Wt %	LV %	Wt %	LV %
Yields								
Coke	18.27	-	9.32	-	8.12	-	14.32	-
<u>Coker Distillate</u>								
Actual	70.94	76.63	86.60	90.53	87.72	91.18	78.92	84.52
C <sub>5</sub> + Liquid	71.84	77.92	87.22	91.43	88.33	91.99	79.81	85.79
<u>Gas</u>								
By Difference	10.79		4.08	-	4.16	-	6.76	-
By Gas Volume + Gas Analyses	6.16		3.06	-	2.77	-	6.06	-
Coker Distillate Gravity, °API	32.7		26.1		26.5		31.3	

TABLE XLIV

DOE CONTRACT EF-76-C-01-2315  
 COKING OF WHOLE PARAHO SHALE OIL  
 COKE QUALITY

Inspection

Sulfur, Wt %	0.65
Nitrogen, Wt %	4.08
Arsenic, ppm	190
Moisture, Wt %	1.08
Volatiles, Wt %	12.9
Ash, Wt %	0.38
Real Density, g/cc <sup>1</sup>	1.300

Metals, ppm

Aluminum	27
Boron	2.4
Barium	0.3
Calcium	71
Chromium	1.4
Copper	2.0
Iron	>100
Magnesium	22
Molybdenum	0.8
Sodium	6.0
Nickel	3.7
Silicon	98
Titanium	2.5

<sup>1</sup>Measured by kerosene displacement.

TABLE XLV

DOE CONTRACT EF-76-C-01-2315  
 COKING OF WHOLE PARAHO SHALE OIL  
 GAS COMPOSITION\*

Compound	Wt %	Mole %
Hydrogen	1.57	17.81
Methane	33.52	47.67
Ethane	17.91	13.59
Ethylene	2.21	1.79
Propane	12.70	6.57
Propylene	6.14	3.33
Isobutane	1.01	0.40
Normal Butane	5.61	2.20
Butenes	4.74	1.93
Isopentane	0.74	0.23
Normal Pentane	2.88	0.91
Pentenes +	10.97	3.57

\*Average gas composition from coking  
 of two drums of Paraho shale oil.

TABLE XLVI

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF SHALE OIL COKER DISTILLATE WITH  
 ICR 106 CATALYST

	535	60	1800
Liquid Product Nitrogen, ppm	535	60	1800
RUN	81- 10	81- 10	81- 10
FEED	WOW 3524	WOW 3524	WOW 3524
RUN HOURS	157.- 181.	325.- 349.	901.- 925.
AVG.CAT.TEMP., F.	718.	733.	716.
LHSV	0.99	1.00	0.99
TOTAL PRESSURE, PSIG	1694.	1693.	1701.
H2 MEAN PRESS., PSIA	1501.	1490.	1521.
TOTAL GAS IN, SCF/B	8949.	9016.	9216.
RECYCLE GAS, SCF/B	7708.	7711.	8017.
NO LOSS PROD. YIELDS	WT.PC. VOL.PC.	WT.PC. VOL.PC.	WT.PC. VOL.PC.
C1	0.18	0.23	0.22
C2	0.48	0.67	0.66
C3	0.37	0.56	0.42
I-C4	0.13 0.19	0.10 0.15	0.07 0.10
N-C4	0.34 0.50	0.27 0.39	0.20 0.29
C5-180 F.	1.95 2.54	1.79 2.32	1.79 2.31
180-350 F.	15.83 18.24	16.32 18.79	13.76 15.75
350- EP F.	79.02 82.62	78.43 82.20	80.89 84.14
TOTAL C5+	96.80 103.41	96.55 103.32	96.44 102.20
ACT./NO LOSS RECOV.	98.71/102.11	103.05/102.20	97.40/102.02
H2 CONS(GROSS), SCF/B	1241.	1305.	1199.
H2 CONS(HCSON), SCF/B	1197.	1247.	1150.
Liquid Product Nitrogen, ppm	1400	500	330
RUN	81- 10	81- 10	81- 10
FEED	WOW 3524	WOW 3524	WOW 3524
RUN HOURS	925.- 949.	1093.- 1117.	1117.- 1165.
AVG.CAT.TEMP., F.	718.	734.	733.
LHSV	0.99	1.02	1.00
TOTAL PRESSURE, PSIG	1702.	1703.	1704.
H2 MEAN PRESS., PSIA	1523.	1491.	1486.
TOTAL GAS IN, SCF/B	9268.	9016.	9460.
RECYCLE GAS, SCF/B	8071.	7760.	8167.
NO LOSS PROD. YIELDS	WT.PC. VOL.PC.	WT.PC. VOL.PC.	WT.PC. VOL.PC.
C1	0.24	0.32	0.30
C2	0.71	0.70	0.68
C3	0.53	0.52	0.54
I-C4	0.08 0.13	0.10 0.16	0.10 0.16
N-C4	0.25 0.37	0.29 0.43	0.34 0.51
C5-180 F.	2.14 2.7A	1.84 2.40	2.00 2.62
180-350 F.	12.82 14.68	13.96 16.00	14.04 16.10
350- EP F.	81.42 84.69	80.52 84.01	80.33 83.85
TOTAL C5+	96.39 102.17	96.33 102.42	96.38 102.58
ACT./NO LOSS RECOV.	99.33/102.02	97.53/102.10	97.33/102.17
H2 CONS(GROSS), SCF/B	1197.	1256.	1293.
H2 CONS(HCSON), SCF/B	1148.	1191.	1234.

TABLE XLVII

DOM CONTRACT EP-76-C-01-2315  
INSPECTIONS OF 350°F+ PRODUCT FROM HYDROTREATING OF  
SHALE OIL COKER DISTILLATE

## Pilot Plant Run 81-10

	73-85	151-181	253-265	313-325	325-349	913-925
Time On Stream, Hr	718	718	717	732	733	718
Average Catalyst Temp., °F	250	535	500	80	60	1800
Whole Product Nitrogen, ppm (Approx.)						
<u>350°F+ Product Inspections</u>						
Gravity, °API	40.1	40.2	39.7		40.6	39.3
Aniline Point, °F		160.2			161.2	156.3
Sulfur, ppm		1.5			0.2	
Total Nitrogen, ppm	305	630	575	93	83	2060
Pour Point, °F		+30			+25	+30
Freeze Point, °F		+30			+30	
Molecular Weight		219				
<u>Group Type, LV % (High Mass)</u>						
Paraffins		39.6			41.6	42.5
Naphthenes		42.3			44.4	40.6
Aromatics		18.1			14.0	16.9
<u>ASTM D 86 Distillation, °F</u>						
St/5		386/410			380/408	
10/30		418/454			418/452	
50		498			495	
70/90		544/604			545/600	
95/EP		628/659			625/659	
% Overhead		98.5			99	
<u>TBP Distillation, °F (Simulated by Chromatography)</u>						
St/5	326/356	322/352	309/359	318/349	315/346	
10/30	382/450	380/450	386/453	375/445	372/443	
50	508	507	513	502	501	
70/90	572/646	571/643	579/653	568/650	568/651	
95/99	676/743	674/736	683/761	691/-	693/-	
<u>Thermal Stability</u>						
Reflectance, % After 90 Min. at 300°F	93		94	97		87.5
Filter Residue After 6 Hr O <sub>2</sub> Blow	3		1	0		0
Color, As is (ASTM D 1500)	<0.5		<0.5	<0.5		<1.5
Color, After 6 Hr O <sub>2</sub> Blow	<1.0		1.0	1.0		<3.0
<u>Viscosity, cSt</u>						
At 100°F		2.604			2.927	2.740
At 122°F		2.133				2.183
At 210°F		1.101			1.095	1.220

TABLE XLVIII

DOE CONTRACT EF-76-C-01-2315  
 INSPECTIONS OF 350°F+ PRODUCT FROM HYDROTREATING  
 SHALE OIL COKER DISTILLATE

Pilot Plant Run 81-10

	925-949	985-997	1057-1069	1081-1093	1093-1117	1117-1165
Time On Stream, Hr						
Average Catalyst Temp., °F	718	718	735	734	734	733
Whole Product Nitrogen, ppm	1400	1300	350	450	500	330
<u>350°F+ Product Inspections</u>						
Gravity, °API	39.4	39.3	40.1	39.8	39.9	39.9
Aniline Point, °F	156.2	156.4	160.3	158.9	159.4	160.8
Sulfur, ppm	5.3					0.93
Total Nitrogen, ppm	1565	1445	390	530	565	365
Pour Point, °F	+25	+25	+25	+25		+20
<u>Group Type, LV % (High Mass)</u>						
Paraffins	42.3	42.1	42.9	42.9	42.6	42.8
Naphthenes	40.8	40.9	41.7	41.3	41.4	41.5
Aromatics	17.0	17.1	15.5	15.7	16.0	15.7
<u>ASTM D 86 Distillation, °F</u>						
St/5	386/409					390/408
10/30	418/455					417/450
50	499					496
70/90	546/602					544/605
95/EP	628/662					629/659
% Overhead	98.5					98.5
<u>TBP Distillation, °F (Simulated by Chromatography)</u>						
St/5	319/350				319/349	320/349
10/30	378/449				367/438	366/433
50	506				505	500
70/90	569/639				575/655	576/658
95/99	670/731				683/743	688/753
<u>Thermal Stability</u>						
Reflectance, % After 90 Min. at 300°F		88.5	87.5	92.0		
Filter Residue After 6 Hr O <sub>2</sub> Blow		1	4	1		
Color, As is (ASTM D 1500)		<1.5	1.0	<0.5		
Color, After 6 Hr O <sub>2</sub> Blow		2.5	<1.5	1.0		
<u>Viscosity, cSt</u>						
At 100°F		2.643	2.642	2.711		
At 122°F		2.126	2.189	2.113		
At 210°F		1.115	1.105	1.103		

TABLE XLIX

DOE CONTRACT NO. EF-76-C-01-2315  
 PROPERTIES OF FRACTIONS OF HYDROTREATED DIESEL PRODUCT FROM  
 SHALE OIL COKER DISTILLATE FROM RUN 81-10 AT 1117-1165 HOURS  
 AVERAGE CATALYST TEMPERATURE 733°F, WHOLE PRODUCT NITROGEN 330 PPM

	350-500	500-600	600+	350-600
Boiling Range, °F	350-400	400-450	450-500	500-600
Wt % of 350°F+ Product	47.0	31.0	22.0	78.0
LV % of 350°F+ Product	47.9	30.5	21.6	78.4
Inspections	43.1	37.3	36.8	149.6
Gravity, °API	141.9	161.9	194.7	
Aniline Point, °F	0.46	0.86	(15)	
Sulfur, ppm	335	350	355	-15
Nitrogen, ppm	-70	+10	+55	-20
Pour Point, °F	-58	+10		
Cloud Point, °F	-39	+22		
Freeze Point, °F	23			
Smoke Point, mm				
Group Type, LV %				
(High Mass)				
Paraffins	42.8	41.2	50.1	368/399
Naphthenes	41.5	41.1	31.2	406/431
Aromatics	15.7	17.6	18.7	461
(22 Component)				492/536
Paraffins				545/562
Naphthenes				99
Aromatics				
ASTM D 86 Distillation, °F				
St/5	390/408	515/521		316/343
10/30	417/450	523/527		356/413
50	496	532		460
70/90	544/605	538/550		515/571
95/EP	629/659	556/569		586/602
% Overhead	98.5	98.5		
TBP Distillation, °F				
(Simulated by Chromatography)				
St/5	320/349	473/504	573/604	
10/30	366/433	513/534	612/634	
50	500	558	653	
70/90	576/658	579/597	676/711	
95/99	688/753	603/612	730/792	

TABLE L

DOE CONTRACT NO. EF-76-C-01-2315  
 PROPERTIES OF FRACTIONS OF HYDROTREATED DIESEL PRODUCT FROM  
 SHALE OIL COKER DISTILLATE FROM RUN 81-10 AT 1165-1213 HOURS  
 AVERAGE CATALYST TEMPERATURE 734°F, WHOLE PRODUCT NITROGEN 380 PPM

	350+	350-550	550-600	600-650	650+	350-600	350-650
Boiling Range, °F							
Wt % of 350°F+ Product	100	63.8	15.5	12.6	8.1	79.3	91.9
LV % of 350°F+ Product	100	64.4	15.3	12.4	8.0	79.7	92.1
Inspections	39.8	41.6	37.3	36.2	37.1	40.6	40.3
Gravity, °API		144.6	171.5	183.3	209.4		
Aniline Point, °F	380						
Sulfur, ppm			+25	+50	+85		+5
Nitrogen, ppm			+26	+52			+6
Pour Point, °F		-40					+12
Cloud Point, °F		-38					
Freeze Point, °F		-20					
Smoke Point, mm		25				48.2	
Cetane Number		39.5					42.2
Group Type, LV %		41.0					41.0
(High Mass)		19.5					16.8
Paraffins							
Naphthenes							
Aromatics							
ASTM D 86 Distillation, °F							
St/5							395/407
10/30							417/448
50							488
70/90							528/575
95/EP							587/601
% Overhead							98.5
TBP Distillation, °F							
(Simulated by Chromatography)							
St/5		321/349	440/544	466/597	552/654	321/351	321/352
10/30		363/422	552/573	605/621	662/684	371/433	375/443
50		461	587	636	699	484	495
70/90		498/534	597/608	647/661	718/749	528/582	553/613
95/99		547/619	613/620	666/670	778/866	592/656	633/698

RFS RE 775651-1

TABLE LI

DOE CONTRACT EF-76-C-01-2315  
 INSPECTIONS OF 180-350°F PRODUCT FROM  
 HYDROTREATING OF SHALE OIL COKER DISTILLATE

Pilot Plant Run 81-10

	145-193	301-349	901-949	1093-1165
Time On Stream, Hr	718	733	718	734
Average Catalyst Temp., °F	535	70	1600	465
Whole Product Nitrogen, ppm (Approximate)				
<u>180-350°F Product Inspections</u>				
Gravity, °API	57.7	57.5	56.5	56.8
Aniline Point, °F	132.8	132.9	131.6	132.0
Sulfur, ppm	33	24	350	(420)
Nitrogen, ppm	99	26	515	200
<u>Octane Number</u>				
F-1 Clear	41.3	40.9	39.6	38.6
F-1 + 3 ml TEL/Gal.			64.0	62.1
F-2 Clear			40.7	40.0
F-2 + 3 ml TEL/Gal.			66.2	62.3
<u>Group Type, LV % (Low Mass)</u>				
Paraffins	60.7	59.2	58.4	58.0
Naphthenes	30.0	30.9	31.6	32.5
Aromatics	9.3	9.8	10.0	9.5
<u>ASTM D 86 Distillation, °F</u>				
St/5	226/240	233/245	234/245	232/244
10/30	243/254	247/259	248/261	248/258
50	267	271	272	270
70/90	281/297	282/297	285/301	283/299
95/EP	303/344	303/353	307/338	306/335
% Overhead	99	99	99	99
<u>TBP Distillation, °F (Simulated by Chromatography)</u>				
St/5	148/192	146/194	150/199	155/197
10/30	200/242	210/248	214/259	213/254
50	271	275	285	282
70/90	299/326	302/328	308/328	305/326
95/99	333/342	335/353	334/341	332/342

TABLE LII

DOE CONTRACT EF-76-C-01-2315  
 DISTRIBUTION OF C<sub>5</sub>-180°F PRODUCT FROM  
 HYDROTREATING OF SHALE OIL COKER  
 DISTILLATE WITH ICR 106  
 PILOT PLANT RUN 81-10

Time Onstream, Hours	1093-1165
Average Catalyst Temperature, °F	733
<u>Composition, LV % of C<sub>5</sub>-180°F</u>	
<u>(By Chromatography)</u>	
n-Butane	0.7
Isopentane	22.4
n-Pentane	14.4
2,2-Dimethylbutane	0.1
2,3-Dimethylbutane	1.8
2-Methylpentane	14.6
3-Methylpentane	8.7
n-Hexane	22.9
Isoheptanes	4.7
n-Heptane	0.3
Total Paraffins	90.6
Cyclopentane	0.9
Methylcyclopentane	4.0
Cyclohexane	2.5
Dimethylcyclopentanes,	0.3
Ethylcyclopentane	0.1
Methylcyclohexane	7.8
Total Naphthenes	1.5
Benzene	1.5
Total Aromatics	0.1
C <sub>4</sub> -C <sub>6</sub> Olefins	0.1
Total Olefins	67.5
Octane Number F-1 Clear (Observed)	1.6
Isopentane/n-Pentane	1.1
Iso-C <sub>6</sub> /n-Hexane	

2-8-78

RFS RE 775665-1

TABLE LIII

DOE CONTRACT NO. EF-76-C-01-2315  
 COMPOSITION OF GUARD BED AND CATALYST AFTER RUN 81-10,  
 HYDROTREATING OF SHALE OIL COKER DISTILLATE WOV 3524

Material	Guard Bed						Catalyst						
	1,2	3,4	5,6	7,8	9,10	11	12 <sup>1</sup>	13	14,15	16,17	18,19	20,21,22	23,24
Layer No.													
Wt of Layer, g	21	22	21	21	19	9	12	10	20	20	20	30	15
Approximate Vol, LV %	20	20	20	20	20	10	12	10	20	20	20	30	15
Composition, Wt %													
Arsenic	0.03	0.03	0.24	0.84	0.74	0.16	2.5	0.70	0.03	<0.01	<0.01	<0.01	<0.01
Iron	0.46	0.36	0.55	1.8	1.0	0.44	0.08	0.04	0.02	0.03	0.02	0.03	0.03
Zinc	0.02	0.01	0.02	0.57	0.47	0.01	ND	ND	ND	ND	ND	ND	ND
Carbon <sup>2</sup>	17.0	18.9			9.1			7.9	6.9			4.0	
Sulfur <sup>2</sup>	0.87	0.71			0.08			5.8	6.2			6.5	

<sup>1</sup>Layer 12 included a small amount of alumina (guard bed) that was removed manually before analysis.  
<sup>2</sup>ND = not detected.

RFS RE 773974-2

11-30-77

TABLE LIV

DOE CONTRACT EF-76-C-01-2315  
PROPERTIES OF DIESEL FRACTION FROM  
 HYDROTREATED WHOLE SHALE OIL

Chevron Identification No.	SGQ 6198
Yield, LV % from Whole Oil Hydrotreating	49.98
<u>Inspections</u>	
Gravity, °API	37.4
Aniline Point, °F	149.1
Sulfur, ppm	40
Nitrogen, ppm	540, 630
Oxygen, ppm	200
Pour Point, °F	0
Freeze Point, °F	+10
<u>Group Type, LV %</u>	
Paraffins	28.7
Naphthenes	44.1
Aromatics	27.2
<u>Simulated TBP Distillation (Chromatographic)</u>	
St/5	327/373
10/30	392/461
50	519
70/90	580/631
95/99	645/663
<u>ASTM D 86 Distillation</u>	
St/5	400/427
10/30	437/467
50	508
70/90	548/586
95/EP	597/618
Overhead, %	99.5
Viscosity, cSt at 100°F	2.818
Viscosity, cSt at 210°F	1.138
Thermal Stability, % Reflectance	56.5

TABLE LV

DOE CONTRACT EF-76-C-01-2315  
 DIESEL HYDROFINISHING--TYPICAL YIELDS  
 FROM RUN 66-185\*\*

Run	66-185		
Feed	SQQ 6198		
Run Hours	514-538		
Average Catalyst Temperature, °F	710		
LHSV	2.00		
Total Pressure, psig	1196		
H <sub>2</sub> Mean Pressure, psia	1096		
Total Gas In, SCF/Bbl	4990		
Recycle Gas, SCF/Bbl	4830		
<u>No Loss Product Yields</u>	Wt %	Vol %	
	C <sub>4</sub> <sup>-</sup>	0.17	44.57
	C <sub>5</sub> -500, °F	43.64	55.82
	500-EP, °F	56.42	100.39
	Total C <sub>5</sub> <sup>+</sup>	100.06	
Actual/No Loss Recovery	98.45/100.22		
H <sub>2</sub> Consumption (Chemical) SCF/Bbl	≤100*		

\*Hydrogen consumption was too low to be determined more accurately.

\*\*Yields were essentially invariable over the range of product nitrogen levels obtained.

TABLE LVI

DOE CONTRACT EF-76-C-01-2315  
 HYDROFINISHING A DIESEL FRACTION FROM HYDROTREATED SHALE OIL  
 INSPECTIONS OF FEED AND WHOLE LIQUID PRODUCT FROM RUN 66-185

	Feed	319-331	379-403	403-415	514-538
Run Hours	SGQ 6198	650	691	692	710
Average Catalyst Temp., °F	37.4	37.6	38.0	37.9	37.9
Gravity, °API	149.1	150.6	152.5	152	151.4
Aniline Point, °F	40	40	15	15	20
Sulfur, ppm	540	350	164	190	161
Nitrogen, ppm	0	-5	+5	-10	0
Pour Point, °F	35.5	35.4	35.4	35.5	35.1
Viscosity, SUS at 100°F	2.82	2.79	2.79	2.82	2.70
Thermal Stability, % Reflectance	56.5	80	79	(71.5)	85
Simulated TBP Distillation (Chromatographic), °F	327/373 392/461 519 580/631 645/663	317/363 383/447 509 569/630 646/669	280/363 383/446 507 570/639 649/668	309/372 393/461 519 579/631 645/662	321/367 387/452 514 577/639 655/676
St/5					
10/30					
50					
70/90					
95/99					

DCG RE 773960-3

10-28-77

TABLE LVII

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF DIESEL FRACTION FROM  
 HYDROTREATED SHALE OIL  
 FRACTION DISTILLED IN CONTINUOUS-DISTILLATION  
 PILOT PLANT

Chevron Identification No.

SGQ 6239

Insolubles

Gravity, °API  
 Aniline Point, °F  
 Sulfur, ppm  
 Nitrogen, ppm  
 Oxygen, ppm  
 Pour Point, °F

38.1  
 149.1  
 20  
 370, 415  
 70  
 -5

Group Type, LV %

Paraffins  
 Naphthenes  
 Aromatics

36.9  
 44.1  
 19.1

Simulated TBP Distillation (Chromatographic), °F

St/5  
 10/30  
 50  
 70/90  
 95/99

309/354  
 383/453  
 509  
 567/622  
 642/667

Viscosity, cSt at 100°F

2.592

Viscosity, cSt at 210°F

1.091

Thermal Stability, % Reflectance

55.5

TABLE LVIII

DOE CONTRACT EF-76-C-01-2315  
 PROPERTIES OF JET FRACTION FROM DIESEL HYDROFINISHING  
 IN RUN 66-185

	415-439	514-558	379-403	403-415
Run Hours				
Whole Liquid Product	184	161	164	190
Nitrogen, ppm				
Nominal Jet Cut Point, °F	500	500	450	475
<u>Inspections</u>				
Gravity, °API	41.0	41.0	43	42
Aniline Point, °F	136.7	134.8	134.4	133.8
Sulfur, ppm	8.1	5.4		
Nitrogen, ppm	139	127	70	100
Pour Point, °F	-40	-45		
Freeze Point, °F	-44	-40	-49	-53
Smoke Point, mm	19	18	23	21
Existent Gum, mg/100 ml	4	4	(8)	
Color, Saybolt			10	
<u>Group Type, LV %</u>				
Paraffins	34.2	33.7		
Naphthenes	43.4	43.7		
Aromatics	22.4	22.6		
<u>Simulated TBP Distillation (Chromatographic)</u>				
St/5	283/350	297/349	(157)/338	(180)/340
10/30	362/401	361/400	349/381	352/384
50	437	435	406	412
70/90	470/506	468/502	431/455	440/467
95/99	514/528	510/519	463/510	478/494

TABLE LIX

DOE CONTRACT EF-76-C-01-2315  
INSPECTIONS OF JET FRACTION FROM  
 HYDROTREATED SHALE OIL

Chevron Identification No.

SGQ 6245

Inspections

Gravity, °API	41.2
Aniline Point, °F	134.8
Sulfur, ppm	20
Nitrogen, ppm	290
Oxygen, ppm	33
Pour Point, °F	-45
Freeze Point, °F	-37
Smoke Point, mm	19
Color, Saybolt	14
Neutralization No., mg KOH/g	0.05

Group Type, LV %

Paraffins	33.6
Naphthenes	40.0
Aromatics	26.4

Simulated TBP Distillation (Chromatographic), °F

St/5	284/349
10/30	367/414
50	444
70/90	468/495
95/99	502/523

TABLE LX

DOE CONTRACT EF-76-C-01-2315  
 HYDROFINISHING A JET FRACTION FROM  
 HYDROTREATED SHALE OIL  
 INSPECTIONS OF WHOLE LIQUID PRODUCT FROM RUN 66-187

	246-258	258-282
Run Hours		
Average Catalyst Temperature, °F	729	729
Estimated H <sub>2</sub> Consumption, SCF/Bbl	150	150
<u>Inspections</u>		
Gravity, °API	41.5	42.0
Aniline Point, °F	139.9	139.4
Nitrogen, ppm	5.1	1.9
Pour Point, °F	-50	-40
Freeze Point, °F	-43	-42
Smoke Point, mm	21	21
Existent Gum, mg/100 ml	5	2
Color, Saybolt		26
<u>Group Type, LV %</u>		
Paraffins	34.7	34.8
Naphthenes	43.2	43.7
Aromatics	22.1	21.5
<u>Simulated TBP Distillation (Chromatographic)</u>		
St/5	272/341	251/338
10/30	353/399	351/395
50	435	430
70/90	468/506	462/497
95/99	528/(628)	508/550
<u>ASTM D 86 Distillation, °F</u>		
St/5		380/392
10/30		398/412
50		424
70/90		440/460
95/EP		468/496
% Overhead		99.0

11-10-77

DCG

RE 773964-4

TABLE LXI

DOE CONTRACT EF-76-C-01-2315

SINGLE STAGE EXTINCTION RECYCLE HYDROCRACKING  
OF 650°F+ HYDROTREATED SHALE OIL WITH ICR 106

RUN	81- 8	81- 8		
FEED	SGR 6152	SGR 6152		
RUN HOURS	215.0- 263.0	263.0- 287.0		
AVG.CAT.TEMP., F.	794.	796.		
LHSV	0.60	0.60		
PER PASS CONVERSION	41.35	36.30		
TOTAL PRESSURE, PSIG	2300.	2304.		
H2 MEAN PRESS., PSIA	2005.	2010.		
TOTAL GAS IN, SCF/B	5399.	5679.		
RECYCLE GAS, SCF/B	4973.	5122.		
NO LOSS PROD. YIELDS	WT.PC.	VOL.PC.	WT.PC.	VOL.PC.
C1	0.38		0.48	
C2	0.57		0.65	
C3	2.11		2.36	
I-C4	2.14	3.34	2.37	3.70
N-C4	2.38	3.58	2.80	4.20
C5-180 F.	10.39	13.91	10.49	14.09
180-300 F.	22.93	27.37	22.42	26.78
300-525 F.	61.24	67.93	60.53	67.14
TOTAL C5+	94.57	109.22	93.45	108.02
ACT./NO LOSS RECOV.	101.68/102.27		101.57/102.22	
H2 CONS(GROSS), SCF/B	1512.		1523.	
H2 CONS(HCSON), SCF/B*	1312.		1287.	

\*Chemical Hydrogen Consumption

TABLE LXII

DOE CONTRACT EF-76-C-01-2315  
 C<sub>5</sub>-180°F AND 180-300°F NAPHTHA PRODUCT PROPERTIES  
 FROM HYDROCRACKING OF 650°F+  
 HYDROTREATED SHALE OIL SGQ 6228 WITH ICR 106 CATALYST

Run 81-8

Time Onstream, Hr	215-263	263-287
<u>Product Inspections</u>		
<u>C<sub>5</sub>-180°F Product</u>		
Gravity, °API	84.3	84.9
Group Type, LV % (by Chromatography)		
Paraffins	88.5	90.8
Naphthenes	10.9	8.7
Aromatics	0.4	0.4
Olefins	0.1	0.1
<u>180-300°F Product</u>		
Gravity, °API	60.8	60.9
Aniline Point, °F	140.7	140.2
Group Type, LV % (Low Mass)		
Paraffins	60.7	61.6
Naphthenes	36.1	34.7
Aromatics	3.2	3.7
Octane Number		
F-1 Clear	48.3	48.4
F-2 Clear	49.5	
TBP Distillation, °F (Simulated by Chromatography)		
ST/5	149/191	140/189
10/30	197/224	195/220
50	253	250
70/90	275/299	270/295
95/99	307/-	300/313

TABLE LXIII

DOE CONTRACT EF-76-C-01-2315  
 WHOLE LIQUID PRODUCT AND 300°F+ JET PRODUCT  
 PROPERTIES FROM HYDROCRACKING OF 650°F+ HYDROTREATED  
 SHALE OIL, SGQ 6228, WITH ICR 106 CATALYST

Run 81-8

Time Onstream, Hr	215-263	263-287
<u>Inspections</u>		
<u>Whole Liquid Product</u>		
Nitrogen, ppm	0.69	0.65
<u>300-525°F Product</u>		
Gravity, °API	47.2	47.2
Aniline Point, °F	163.6	161.9
Group Type, LV % (High Mass)		
Paraffins	53.5	53.9
Naphthenes	35.8	33.6
Aromatics	10.7	12.5
Smoke Point, mm	27	30
Freeze Point, °F	-66	-68
<u>ASTM D 86 Distillation, °F</u>		
SE/5	344/360	343/360
10/30	367/388	365/386
50	415	414
70/90	445/475	444/476
95/EP	484/511	485/507
Overhead, LV %	98.5	98.5
<u>TBP Distillation, °F (Simulated by Chromatography)</u>		
SE/5	278/316	276/315
10/30	331/382	330/378
50	430	425
70/90	476/518	469/517
95/99	528/-	528/542

TABLE LXIV

DOE CONTRACT EF-76-C-01-2315  
 DISTRIBUTION OF C<sub>5</sub>-180°F PRODUCT FROM  
 HYDROCRACKING OF 650°F+ HYDROTREATED  
 SHALE OIL WITH ICR 106 CATALYST  
 PILOT PLANT RUN 81-8

Time Onstream, Hours Average Catalyst Temperature, °F	215-263 794	263-287 796
Composition, LV % of C <sub>5</sub> -180°F (By Chromatography)		
n-Butane	-	
Isopentane	(19.0)	31.3
	(?)	
n-Pentane	21.0	20.7
2,2-Dimethylbutane	0.4	0.3
2,3-Dimethylbutane	1.7	1.4
2-Methylpentane	15.4	12.0
3-Methylpentane	10.1	7.9
n-Hexane	17.9	14.1
Isoheptanes	3.0	2.9
n-Heptane	-	0.1
C <sub>8</sub> Isoparaffins	-	0.1
Total Paraffins	88.5	90.8
Cyclopentane	1.0	0.7
Methylcyclopentane	8.1	6.4
Cyclohexane	1.0	0.9
Dimethylcyclopentanes, Ethylcyclopentane	0.7	0.7
C <sub>8</sub> Naphthenes	0.1	-
Total Naphthenes	10.9	8.7
Benzene	0.4	0.4
Total Aromatics	0.4	0.4
C <sub>4</sub> -C <sub>6</sub> Olefins	0.1	0.1
Total Olefins	0.1	0.1
Octane Number F-1 Clear (Estimate)	(70.0)	(73.7)
Isopentane/n-Pentane	(0.9)	1.5
Iso-C <sub>5</sub> /n-Hexane	1.5	1.5

TABLE LXV

DOE CONTRACT NO. EF-76-C-01-2315  
INSPECTIONS OF BLEND OF HYDROTREATED SHALE OIL,  
 625°F+ BOTTOMS, AND 625-850°F FRACTION

Blend Identification	WOW 3535	SGQ 6214	SGQ 6228
Boiling Range	Whole Liquid Product <sup>1</sup>	625°F+ <sup>2</sup>	625-850°F <sup>3</sup>
<u>Inspections of Blend</u>			
Gravity, °API	37.2	30.8	30.9
Aniline Point, °F	170.6	206.6	194.0
Sulfur, ppm	25		84
Nitrogen, ppm	405	530	535
Oxygen, ppm	80		60
Molecular Weight (Avg.)	253		318
Pour Point, °F	+80	+105	+85
Ramsbottom Carbon, %	0.07		0.05
Hot Heptane Asphaltenes, ppm	38		336
<u>Group Type, LV % (22 Component)</u>			
Paraffins	38.3		41.1
Naphthenes	36.8		33.8
Aromatics	27.9		25.1
<u>TBP Distillation, °F (Simulated by Chromatography)</u>			
St/5	88/253		471/609
10/30	313/452		633/689
50	559		733
70/90	667/818		780/830
95/99	876/		847/881
<u>ASTM D 86/D 1160 Distillation, °F</u>			
St/5	195/290		621/660
10/30	335/499		671/696
50	586		727
70/90	688/844		764/807
95/EP	896/970		831/877
% Overhead	99		99
Carbon, Wt %	86.42		87.06/88.57
Hydrogen, Wt %	13.58		14.06/13.82
<u>Viscosity, cSt</u>			
at 100°F			15.46
at 122°F	2.938		10.29
at 210°F	1.508		3.394

<sup>1</sup>Product from 86-51 at 2071-2887 hours.

<sup>2</sup>37 LV % of whole liquid product based on TBP distillation of WOW 3535.

<sup>3</sup>74.7 Wt %, 74.8 LV % of SGQ 6214 (625°F+).

TABLE LXVI

DOE CONTRACT EP-76-C-01-2315  
 HYDROCRACKING OF 625-850°C HYDROTREATED SHALE OIL  
 WITH ICR 106 CATALYST

	81- 11	81- 11	81- 11	81- 11	81- 11	81- 11
RUN FEED	81- 11 S60 6225	81- 11 S60 6228				
RUN HOURS	728.- 764.	764.- 800.	764.- 800.	790.-	908.-	932.-
AVG. CAT. TEMP., F.	785.	784.	784.	790.	790.	790.
LHSV	1.00	1.00	1.00	1.00	1.00	1.00
PER PASS CONVERSION	53.51	52.67	52.67	51.97	51.97	51.35
TOTAL PRESSURE, PSIG	2350.	2355.	2355.	2355.	2355.	2360.
H2 MEAN PRESS., PSIA	2129.	2128.	2128.	2145.	2145.	2150.
TOTAL GAS IN, SCF/B	8660.	9076.	9076.	8548.	8548.	8779.
RECYCLE GAS, SCF/B	7907.	8321.	8321.	7836.	7836.	8016.
NO LOSS PROD. YIELDS						
C1	0.31	0.31	0.31	0.30	0.30	0.31
C2	0.47	0.48	0.48	0.44	0.44	0.45
C3	2.01	2.10	2.10	1.98	1.98	2.03
I-C4	2.39	2.68	2.68	2.44	2.44	2.52
N-C4	1.81	3.70	4.15	1.75	1.75	1.83
C5-180 F.	10.09	2.70	3.08	10.25	10.25	10.25
180-300 F.	22.65	13.38	13.43	22.11	22.11	22.11
300-535 F.	62.36	26.79	27.03	62.75	62.75	62.75
TOTAL C5+	95.11	68.77	66.73	95.12	95.12	95.12
ACT./NO LOSS RECOV.	100.45/102.15	98.54/102.20	98.54/102.20	99.01/102.07	99.01/102.07	98.30/102.27
H2 CONS(GROSS), SCF/B	1407.	1432.	1432.	1369.	1369.	1484.
H2 CONS(H2CON), SCF/B	1236.	1264.	1264.	1192.	1192.	1304.
WT.PC.	0.31	0.31	0.31	0.30	0.30	0.31
VOL.PC.	0.47	0.48	0.48	0.44	0.44	0.45
WT.PC.	2.01	2.10	2.10	1.98	1.98	2.03
VOL.PC.	2.39	2.68	2.68	2.44	2.44	2.52
WT.PC.	1.81	3.70	4.15	1.75	1.75	1.83
VOL.PC.	10.09	2.70	3.08	10.25	10.25	10.25
WT.PC.	22.65	13.38	13.43	22.11	22.11	22.11
VOL.PC.	62.36	26.79	27.03	62.75	62.75	62.75
WT.PC.	95.11	68.77	66.73	95.12	95.12	95.12
VOL.PC.	100.45/102.15	98.54/102.20	98.54/102.20	99.01/102.07	99.01/102.07	98.30/102.27
WT.PC.	1407.	1432.	1432.	1369.	1369.	1484.
VOL.PC.	1236.	1264.	1264.	1192.	1192.	1304.

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Chemical Hydrogen Consumption

RFS RE 77977-2

TABLE LXVII

DOE CONTRACT EF-76-C-01-2315  
 C<sub>5</sub>-180°F AND 180-300°F NAPHTHA PRODUCT PROPERTIES  
 FROM HYDROCRACKING OF 625-850°F  
 HYDROTREATED SHALE OIL SGQ 6228 WITH ICR 106 CATALYST

Run 81-11

Time On Stream, Hr	728-764	764-800	884-908
<u>Product Inspections</u>			
<u>C<sub>5</sub>-180°F Product</u>			
Gravity, °API	84.0	84.5	84.0
Group Type, LV % (by Chromatography)			
Paraffins	88.9	89.3	
Naphthenes	10.8	10.3	
Aromatics	0.2	0.3	
Olefins	0.1	0.1	
<u>Octane Number</u>			
F-1 Clear	73.8	74.7	75.0
F-2 Clear	71.6	71.8	71.6
<u>180-300°F Product</u>			
Gravity, °API	60.6	60.3	60.3
Aniline Point, °F	139.9	139.5	139.9
Group Type, LV % (Low Mass)			
Paraffins	52.3	51.7	52.9
Naphthenes	45.5	46.0	43.8
Aromatics	2.2	2.2	3.2
<u>Octane Number</u>			
F-1 Clear	50.2	51.3	50.9
F-2 Clear	51.2	51.2	
<u>ASTM D 86 Distillation, °F</u>			
St/5	213/218	213/221	
10/30	222/231	223/232	
50	240	240	
70/90	251/270	252/273	
95/EP	280/333	282/345	
% Overhead, LV %	99	99	
<u>TBP Distillation, °F (Simulated by Chromatography)</u>			
St/5	142/190	141/188	135/170
10/30	197/221	195/220	192/221
50	250	250	251
70/90	268/295	268/295	275/303
95/99	300/313	300/312	311/326

TABLE LXVIII

DOE CONTRACT EF-76-C-01-2315  
 WHOLE LIQUID PRODUCT AND 300°F+ JET PRODUCT  
 PROPERTIES FROM HYDROCRACKING OF 625-850°F HYDROTREATED  
 SHALE OIL, SQ 6228, WITH ICR 106 CATALYST

Run 81-11

Time On Stream, Hr	728-764	764-800	884-908	908-932
<u>Inspections</u>	0.47	0.57	0.18	0.89
<u>Whole Liquid Product</u>				
Nitrogen, ppm	47.6	45.5	47.2	47.1
300-535°F Product	167.2	166.9	167.3	165.9
Gravity, °API	54.4	54.1	53.8	54.2
Aniline Point, °F	41.1	41.2	39.5	38.3
Group Type, LV %	4.5	4.7	7.0	7.4
(High Mass)	34	37	31	34
Paraffins	-63	-76	-70	-63
Napthenes				
Aromatics				
Smoke Point, mm				
Freeze Point, °F				
Pour Point, °F				
ASTM D 86 Distillation, °F				
St/5	338/353	326/353	339/356	336/356
10/30	360/384	359/384	363/388	364/387
50	414	412	418	415
70/90	447/483	444/481	451/486	446/482
95/EP	494/502	492/506	500/515	497/513
Overhead, LV %	99.0	99.0	99.5	99.5
TBP Distillation, °F				
(Simulated by Chromatography)				
St/5	261/310	266/310	266/304	264/312
10/30	327/376	328/375	321/373	329/379
50	422	421	423	428
70/90	471/521	468/519	473/526	476/526
95/99	534/548	532/545	539/551	539/-

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1-11-78

TABLE LXIX

DOE CONTRACT EF-76-C-01-2315  
 DISTRIBUTION OF C<sub>5</sub>-180°F PRODUCT FROM  
 HYDROCRACKING OF 625-850°F HYDROTREATED  
 SHALE OIL WITH ICR 106 CATALYST  
 PILOT PLANT RUN 81-11

Time Onstream, Hours Average Catalyst Temperature, °F	728-764 785	764-800 784
<b>Composition, LV % of C<sub>5</sub>-180°F (By Chromatography)</b>		
n-Butane	2.8	2.6
Isopentane	18.5	18.9
n-Pentane	25.5	26.9
2,2-Dimethylbutane	0.4	0.4
2,3-Dimethylbutane	1.7	1.6
2-Methylpentane	14.0	13.6
3-Methylpentane	9.1	8.8
n-Hexane	14.4	13.7
Isoheptane	2.4	2.7
n-Heptane	0.1	0.1
C <sub>8</sub> Isoparaffins	0.01	0.02
<b>Total Paraffins</b>	<b>88.9</b>	<b>89.3</b>
Cyclopentane	1.0	1.0
Methylcyclopentane	8.0	7.8
Cyclohexane	1.1	1.0
Dimethylcyclopentanes, Ethylcyclopentane	0.7	0.5
C <sub>8</sub> Naphthenes	0.03	0.04
<b>Total Naphthenes</b>	<b>10.8</b>	<b>10.3</b>
Benzene	0.2	0.3
<b>Total Aromatics</b>	<b>0.2</b>	<b>0.3</b>
<b>Octane Number, F-1 Clear (Observed)</b>	<b>73.8</b>	<b>74.7</b>
Isopentane/n-Pentane	0.7	0.7
Iso-C <sub>6</sub> /n-Hexane	1.8	1.8

2-8-78

RFS

RE 775667-1

TABLE LXX

DOE CONTRACT EF-76-C-01-2315  
CRACKING CATALYST INSPECTIONS

Identification No.	CCL-4914
Catalyst Type	Davison CBZ-1
Source	Equilibrium Catalyst from an FCC Unit
Chevron Research Activity, LV % Conversion*	69.7
<u>Physical Properties</u>	
Surface Area, m/g	107
Apparent Bulk Density, g/cc	0.779
Pore Volume, cc/g	0.38
<u>Particle Size Distribution, Wt %</u>	
0-20 $\mu$	0.0
20-40 $\mu$	0.5
40-80 $\mu$	64.0
80+ $\mu$	35.5
Average Particle Size, $\mu$	72.6
<u>Metals, ppm</u>	
Nickel	266
Vanadium	318
Iron	3020
Copper	9
Chromium	75

\*Chevron Research activity is defined as the conversion (LV %) obtained in cracking a light East Texas gas oil (35.4°API; 500-700°F) in a fixed-fluidized test unit (FCTU) operated in a cyclic manner. Conversion is defined as 100- vol %, 430°F+ liquid. Test conditions are:

Reactor Temperature, °F	925
Feed Rate, ml/Min.	30
Feed Period, Min.	10
Catalyst Charge, g	500

TABLE LXXI

DOE CONTRACT EF-76-C-01-2315  
 CATALYTIC CRACKING OF HYDROTREATED SHALE OILS  
 OPERATING CONDITIONS AND PRODUCT YIELDS

Run Feed Catalyst	FCT 5-1038		FCT 5-1039	
	CCL-5406		CCL-5407	
	CCL-4914		CCL-4914	
<u>Operating Conditions</u>				
Temperature, °F	975		975	
Catalyst Charge, g	600		600	
WHSV	5.96		5.96	
Cat./Oil Weight Ratio	2.01		2.01	
	Wt %	LV %	Wt %	LV %
Conversion	78.84	80.98	73.45	75.38
<u>Product Yields</u>				
Coke	4.41		3.87	
H <sub>2</sub>	0.06		0.07	
C <sub>1</sub>	1.12		1.07	
Ethane	0.88		0.89	
Ethene	<u>0.78</u>		<u>0.71</u>	
Total C <sub>2</sub> 's	1.66		1.60	
Propane	3.17	5.43	2.46	4.23
Propene	<u>5.97</u>	<u>9.97</u>	<u>5.15</u>	<u>8.60</u>
Total C <sub>3</sub> 's	9.14	15.40	7.61	12.83
i-C <sub>4</sub>	7.15	11.06	5.73	8.88
n-C <sub>4</sub>	2.10	3.14	1.69	2.52
C <sub>4</sub> Olefins	<u>6.45</u>	<u>9.21</u>	<u>6.00</u>	<u>8.57</u>
Total C <sub>4</sub> 's	15.70	23.41	13.42	19.97
Light Gasoline (C <sub>5</sub> -250°F)	29.64	38.35	28.70	37.05
Heavy Gasoline (250-430°F)	<u>17.11</u>	<u>17.81</u>	<u>17.11</u>	<u>17.97</u>
Total Gasoline (C <sub>5</sub> -430°F)	46.75	56.16	45.81	55.02
Light Cycle Oil (430-625°F)	9.68	8.88	10.73	10.04
Heavy Cycle Oil (625°F+)	<u>11.48</u>	<u>10.14</u>	<u>15.82</u>	<u>14.58</u>
Total Cycle Oil (430°F+)	21.16	19.02	26.55	24.62
C <sub>3</sub> + Liquid Product	92.75	113.99	93.39	112.44

TABLE LXXII

DOE CONTRACT EF-76-C-01-2315  
 CATALYTIC CRACKING OF HYDROFINED ARABIAN LIGHT VACUUM  
 GAS OILS OPERATING CONDITIONS AND PRODUCT YIELDS

Run Feed Catalyst	PP 158-66D		PP 158-67C		PP 158-63B		PP 158-62D	
	CCL-4993		CCL-4914		CCL-4990		CCL-4990	
Operating Conditions	973		976		974		975	
Reactor Temperature, °F	5.01		5.08		5.04		5.03	
WHSV	3.80		7.81		4.37		7.12	
Cat./Oil Wt Ratio								
Conversion Product Yields	Wt %	LV %						
Coke	75.91	78.12	79.74	81.93	72.18	74.61	75.38	77.83
H <sub>2</sub>	4.67		6.41		6.47		8.27	
C <sub>1</sub>	0.18		0.18		0.24		0.24	
C <sub>2</sub>	1.25		1.19		1.38		1.27	
	2.22		1.98		2.56		2.18	
Propane	2.32	4.07	2.60	4.57	1.92	3.40	2.07	3.68
Propene	4.88	8.33	4.86	8.30	4.84	8.37	4.89	8.45
Total C <sub>3</sub> 's	7.20	12.40	7.46	12.86	6.76	11.77	6.96	12.13
1-C <sub>4</sub>	4.95	7.84	6.05	9.58	3.76	6.03	4.54	7.28
n-C <sub>4</sub>	1.76	2.68	1.98	3.03	1.44	2.22	1.59	2.45
C <sub>4</sub> Olefins	4.94	7.21	4.70	6.86	5.08	7.51	5.07	7.50
Total C <sub>4</sub> 's	11.65	17.73	12.74	19.47	10.28	15.76	11.20	17.23
Light Gasoline (C <sub>5</sub> -250°F)	28.49	36.91	30.14	39.13	24.96	32.75	25.71	33.78
Heavy Gasoline (250-430°F)	20.18	21.57	19.62	20.89	19.41	21.11	19.36	20.97
Total Gasoline C <sub>5</sub> -430°F)	48.66	58.48	49.76	60.02	44.37	53.86	45.06	54.74
Light Cycle Oil (430-625°F)	13.37	12.53	11.65	10.77	13.70	13.07	12.29	11.61
Heavy Cycle Oil (625°F+)	10.72	9.36	8.61	7.29	14.12	12.33	12.33	10.56
Total Cycle Oil (430°F+)	24.09	21.88	20.26	18.07	27.82	25.39	24.62	22.17
C <sub>3</sub> + Liquid Product	91.60	110.49	90.22	110.42	89.23	106.78	87.85	106.27

TABLE LXXIII

DOE CONTRACT EP-76-C-01-2315  
 CATALYTIC CRACKING OF HYDROTREATED SHALE OILS  
 PRODUCT QUALITY

Run Stock	FCT 5-1038				FCT 5-1039			
	Lt Gaso. ST-250°F	Hvy Gaso. 250-430°F	Lt Cycle Oil 430-625°F	Hvy Cycle Oil 625°F+	Lt Gaso. ST-250°F	Hvy Gaso. 250-430°F	Lt Cycle Oil 430-625°F	Hvy Cycle Oil 625°F+
Inspection								
Gravity, °API	65.2*	37.6	17.5	12.0	63.7*	39.0	20.3	18.3
Aniline Pt, °F	113.2	<30	<30		110.0	<30		
Breaine No.	62	13	20		86	22	28	
Sulfur, %		0.006	0.027	0.28		0.006	0.009	0.15
Nitrogen, ppm	2.6	39	130	726	9.7	141	388	1400
FIAM								
P + N	62	24			56	27	17	
O	31	8			38	13	4	
A	7	68			6	60	79	
Viscosity								
at 100°F			2.449	28.18	90.3		34.26	21.08
at 130°F			1.764	13.50	79.5		32.09	11.50
Four Pt, °F			0	+70			-20	+85
Octane No.								
F-1, Clear	89.0	95.3						
F-2, Clear	79.1	84.6						

\*Gravity of 140-250°F fraction does not include C<sub>5</sub>+ material recovered from the gas.

TABLE LXXIV

DOE CONTRACT EF-76-C-01-2315  
 CATALYTIC CRACKING OF HYDROTREATED SHALE OILS  
 GASOLINE OCTANE NUMBERS

Feed Feed Nitrogen, ppm	Hydrotreated Shale Oil		Hydrofined Arabian Light Gas Oil			
	385	870	320		860	
Conversion, LV %	80.98	75.38	78.12	81.93	74.61	77.84
<u>Octane Numbers</u>						
<u>Light Gasoline</u> <u>(C<sub>5</sub>-250°F)</u>						
F-1 Clear	89.0	90.3	91.2	-	90.2	-
F-2 Clear	79.1	79.5	80.8	-	78.8	79.4
<u>Heavy Gasoline</u> <u>(250-430°F)</u>						
F-1 Clear	95.3	93.3	94.4	94.4	93.0	93.6
F-2 Clear	84.6	82.8	83.3	84.2	82.8	83.5

TABLE LXXV

DOE CONTRACT NO. EF-76-C-01-2315  
 CATALYTIC CRACKING OF HYDROTREATED PARAHO SHALE OIL  
 OPERATING CONDITIONS AND PRODUCT YIELDS

Run PP 158-	511		512		513	
	SQG 6139 (CCL-5420) 650°F+ Hydro-treated Paraho Shale Oil					
	CCL-4914					
Feed Catalyst						
<u>Operating Conditions</u>						
Temperature, °F	930		930		975	
WHSV	7.85		7.86		7.71	
Cat/Oil Weight Ratio	6.71		4.39		7.10	
Severity (Cat./Oil + WHSV)	0.855		0.558		0.921	
	Wt %	LV %	Wt %	LV %	Wt %	LV %
Conversion	69.25	70.48	61.25	62.09	76.97	78.61
<u>Product Yields</u>						
Coke	3.61		3.01		4.87	
Hydrogen	0.14		0.19		0.20	
Methane	0.89		0.87		1.54	
Ethane, Ethene	1.29		1.37		2.20	
C <sub>2</sub> - Gas	2.32		2.43		3.94	
Propane	1.60	2.78	1.24	2.15	2.34	4.08
Propene	3.85	6.53	3.16	5.36	4.98	8.44
Total C <sub>3</sub> 's	5.45	9.31	4.40	7.51	7.32	12.52
Isobutane	3.88	6.05	2.77	4.35	4.48	7.03
n-Butane	1.08	1.63	0.79	1.20	1.42	2.15
Butenes	4.04	5.87	3.59	5.21	5.14	7.46
Total C <sub>4</sub> 's	9.00	13.59	7.15	10.76	11.04	16.64
Light Gasoline (C <sub>5</sub> -250°F)	27.30	35.01	23.49	30.02	29.77	38.19
Heavy Gasoline (250-430°F)	21.57	23.62	20.78	22.84	20.04	21.56
Total Gasoline (C <sub>5</sub> -430°F)	48.87	58.63	44.26	52.86	49.81	59.75
Light Cycle Oil (430-625°F)	12.77	12.43	13.62	13.41	10.83	10.30
Heavy Cycle Oil (625°F+)	17.98	17.09	25.13	24.50	12.20	11.09
Total Cycle Oil (430°F+)	30.75	29.52	38.75	37.91	23.03	21.39
C <sub>3</sub> + Liquid Product	94.07	111.05	94.56	109.04	91.20	110.30

TABLE LXXVI

DOE CONTRACT NO. EF-76-C-01-2315  
 CATALYTIC CRACKING OF HYDROTREATED PARAHO SHALE OIL  
 GASOLINE AND CYCLE OIL PROPERTIES

	PP 158-511B		PP 158-512B		PP 158-513	
	70.48		62.09		78.61	
	C5-250°F	250-430°F	C5-250°F	250-430°F	C5-250°F	250-430°F
<u>Run</u>						
<u>Conversion, LV %</u>						
<u>Inspections</u>						
<u>Gasolines</u>						
Gravity, °API	112.7	43.6	107.4	44.3	107.5	40.5
Aniline Point, °F	67.8	68.3	86.1	76.8	72.5	42.9
Bromine No., g/100 g		31		47	0.0040	25.9
Sulfur, Wt %	26	0.0010	24	<0.0010	16	0.0010
Nitrogen, ppm		153		186		127
<u>FIAM</u>						
Paraffins +	43	35	46	33	52	27
Naphthenes						
Olefins	51	21	49	28	42	19
Aromatics	6	44	5	39	6	54
<u>Octane Numbers</u>						
F-1 Clear	89.2	84.1	90.4	83.1	90.9	90.0
F-2 Clear	79.5	77.2	79.8	76.0	79.7	80.9
	430-625°F	625°F+	430-625°F	625°F+	430-625°F	625°F+
<u>Cycle Oils</u>						
Gravity, °API	24.1	20.5	26.0	24.5	20.6	13.9
Aniline Point, °F	62.1		74.9		38.7	
Bromine No., g/100 g	14		18		20.3	
Sulfur, Wt %	0.014	0.017	0.0070	0.0090	0.022	0.034
Nitrogen, ppm	380	1400	490	1200	450	1600
<u>FIAM</u>						
Paraffins +	25		21		2	
Naphthenes						
Olefins	13		22		27	
Aromatics	62		57		71	
	cst	SUS	cst	SUS	cst	SUS
Viscosity at 100°F	2.718	35.11	26.28	124.8	19.33	94.95
Viscosity at 130°F	1.747	-	8.720	54.54	10.68	61.42
Pour Point, °F	-25	+85	-25	+85	-30	+75
Ramsbottom Carbon		1.17	19.5	0.91	8.0	3.08
Diesel Index	15.0					

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TABLE LXXVII

DOE CONTRACT EF-76-C-01-2315  
 INSPECTIONS ON FEEDS USED IN SHALE OIL  
 HYDROTREATING FOR COLORADO SCHOOL OF MINES

Retort Location or Process	Laramie	TOSCO II
Chevron ID No.	SGQ 6131	SGQ 6132
<u>Inspections</u>		
Gravity, °API	28.4	28.0
Sulfur, Wt % (XRA)	0.916	0.987
Sulfur, Wt % (XRF)	0.75	0.82
Nitrogen, Wt %	1.57	1.57
Oxygen, Wt %	1.16	1.13
Carbon, Wt %	84.60	84.92
Hydrogen, Wt %	12.05	11.49
Arsenic, ppm	4.2	14.0
Iron, ppm	6.2	2.1
Low Level Asphaltenes, ppm	1300	908
Ramsbottom Carbon, Wt %	0.26	0.41
Molecular Weight	239	246
Hydrogen/Carbon	1.71	1.62
Bromine No.	44	71
Pour Point, °F	55	35
Viscosity, cSt at 100°F	5.521	3.908
Viscosity, cSt at 210°F	1.726	1.375
<u>ASTM D 86/D 1160</u>		
<u>Distillation</u>		
St/5	146/360	
10/30	392/516	
50	598	
70/90	685/800	
95/EP	841/924	
LV % Overhead	99	
<u>Simulated TBP Distillation</u>		
<u>(Chromatographic)</u>		
St/5	186/323	121/240
10/30	362/477	284/419
50	574	528
70/90	671/806	646/799
95/99	853/(946)	852/(963)

TABLE LXXVIII

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF TWO SHALE OIL DISTILLATES  
 INSPECTION OF THE MAJOR PRODUCTS

Chevron ID No., SQQ- Original Feed, Retort Process Conditions	6201	6204	6203	6202	6205	6206	6199	6200
	← Laramie →							
	← -0.5 LHSV, 1900 psig, 5000 SCF/Bbl OT H <sub>2</sub> , ICR 106 →							
<u>Product Inspections</u>								
Gravity, °API	31.1	34.3	36.3	38.4	30.4	33.8	36.7	39.2
Aniline Point, °F	128.5	144.7	157.3	167.6	123.6	136.7	151.1	163.6
Sulfur, Wt %	0.14	0.030	0.018	0.006	0.13	0.018	0.022	0.004
Nitrogen, Wt %	1.35	0.73	0.43	0.080	1.55	0.82	0.30	0.014
<u>Chromatographic Simulated TBP</u>								
St/5	265/347	244/344	215/330	226/331	270/338	228/323	192/300	216/303
10/30	385/482	380/476	364/461	363/456	370/468	350/444	333/428	335/424
50	577	570	554	544	567	542	515	504
70/90	677/810	670/809	652/795	638/779	672/811	647/793	619/775	600/751
95/99	856/967	856/975	842/991	827/921	859/985	844/969	835/945	815/929
Estimated H <sub>2</sub> Cons., SCF/Bbl	390	850	1060	1350	390	840	1190	1470
Estimated Liquid Yield, LV % of Feed	101.0	102.3	103.2	104.2	101.3	102.5	103.8	105.2
Quantity of Sample, g	1770	2140	1310	2900	1600	1640	2115	870

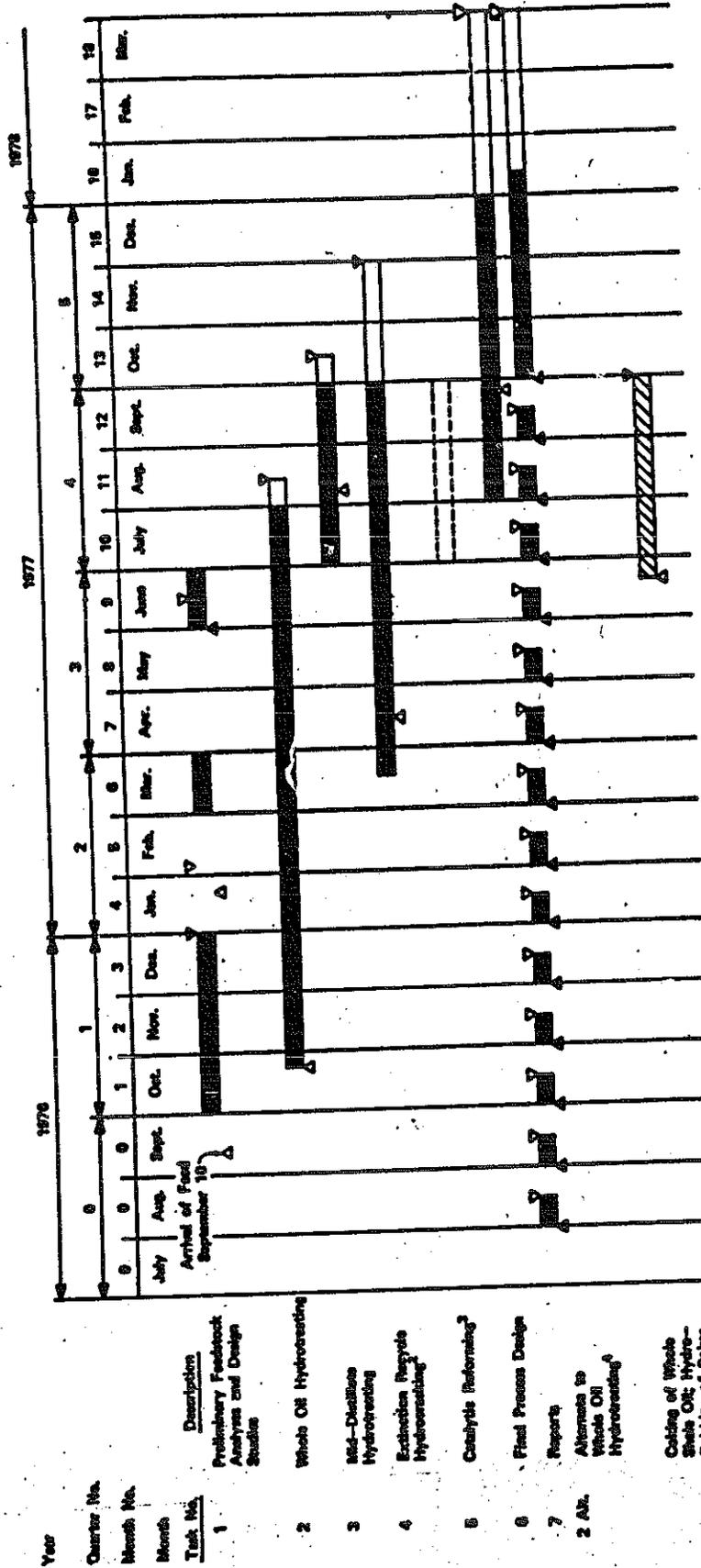
TABLE LXXIX

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF TWO SHALE OIL DISTILLATES  
 INSPECTIONS OF ADDITIONAL PRODUCT SAMPLES

Chevron ID No., SGQ- Original Feed, Retort Process Conditions	6210	6212	6211	6207	6213	6208	6209
	←-0.5 LHSV, 1900 psig, 5000 SCF/Bbl OT H <sub>2</sub>		Laramie		TOSCO II		
<u>Product Inspections</u>							
Gravity, °API	35.1	36.6	38.9	29.2	32.5	38.2	39.5
Aniline Point, °F	150.5	158.8	172.3	106.1	128.5	156.4	161.5
Sulfur, Wt %	0.046	0.010	0.002	0.46	0.052	0.024	0.002
Nitrogen, Wt %	0.77	0.36	0.008	1.65	1.02	0.20	0.035
<u>Chromatographic Simulated TBP</u>							
St/5	259/351	209/325	249/339	236/320	208/324	209/301	215/301
10/30	383/475	360/459	368/456	350/457	351/451	333/423	331/421
50	573	551	539	560	545	506	496
70/90	668/813	648/796	630/775	667/804	648/800	605/762	590/734
95/99	862/997	845/947	826/924	849/923	853/961	824/950	798/910

FIGURE 1

DOE CONTRACT EF-78-C-81-3276  
 REVISED SCHEDULE AS APPLIED TO  
 FIRST FEED (PARANO SHALE OIL)<sup>1</sup>



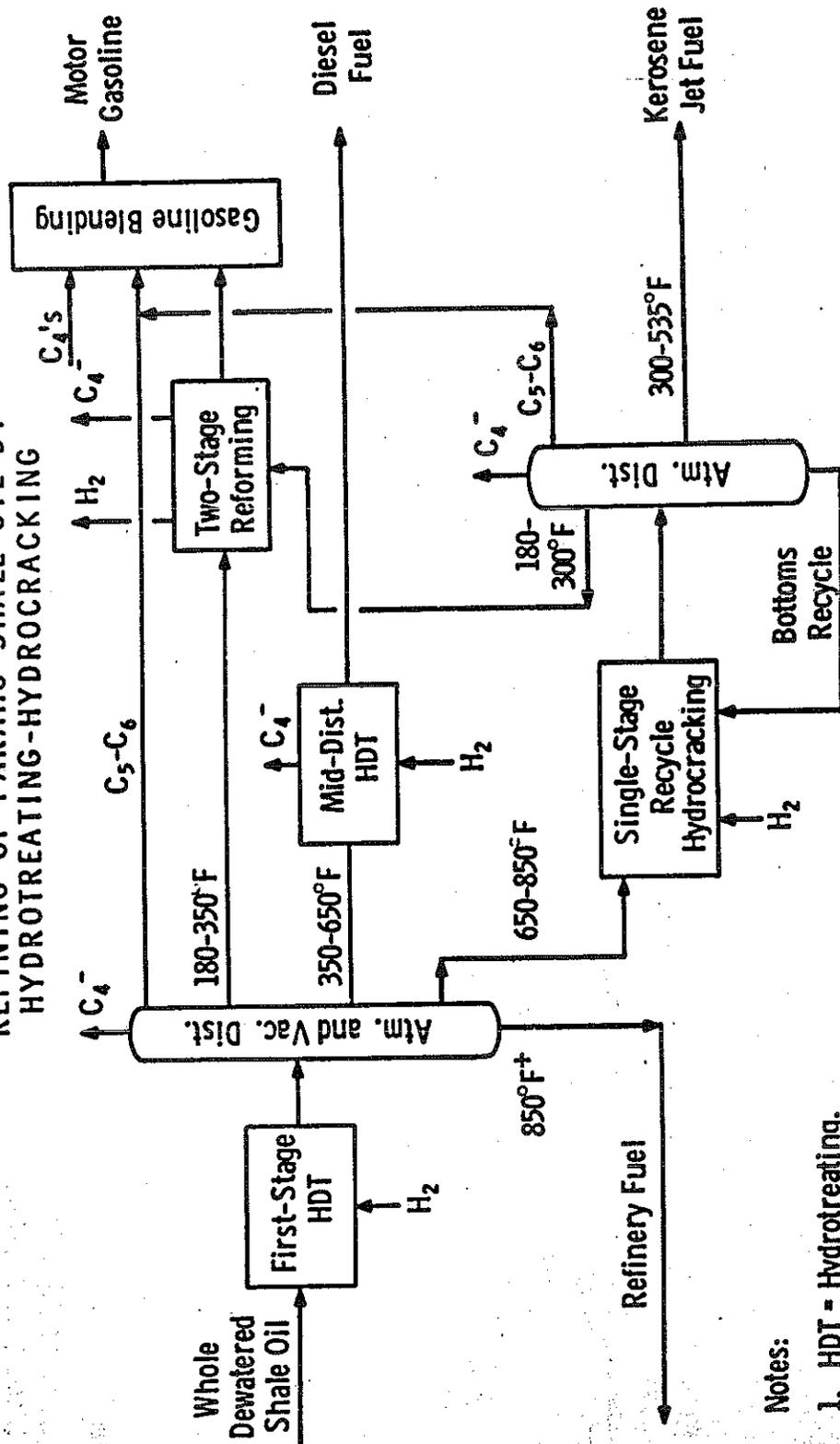
<sup>1</sup>In addition to the seven major tasks, an added task (Task 8) "Distillate Shale Oil Hydrocracking" was performed in August and September, 1977, as a result of a contract modification A004.  
<sup>2</sup>Also includes Task 4A - Alternate Process (FCC).  
<sup>3</sup>It is included only if deemed appropriate by the mutual agreement of the DOE Technical Representative and Chevron Research, subject of a contract modification.  
<sup>4</sup>Discussed in last paragraph of Task 2, work statement.

FIGURE 2

SCHMATIC FLOW DIAGRAM

CASE 1

REFINING OF PARAHO SHALE OIL BY  
HYDROTREATING-HYDROCRACKING



Notes:

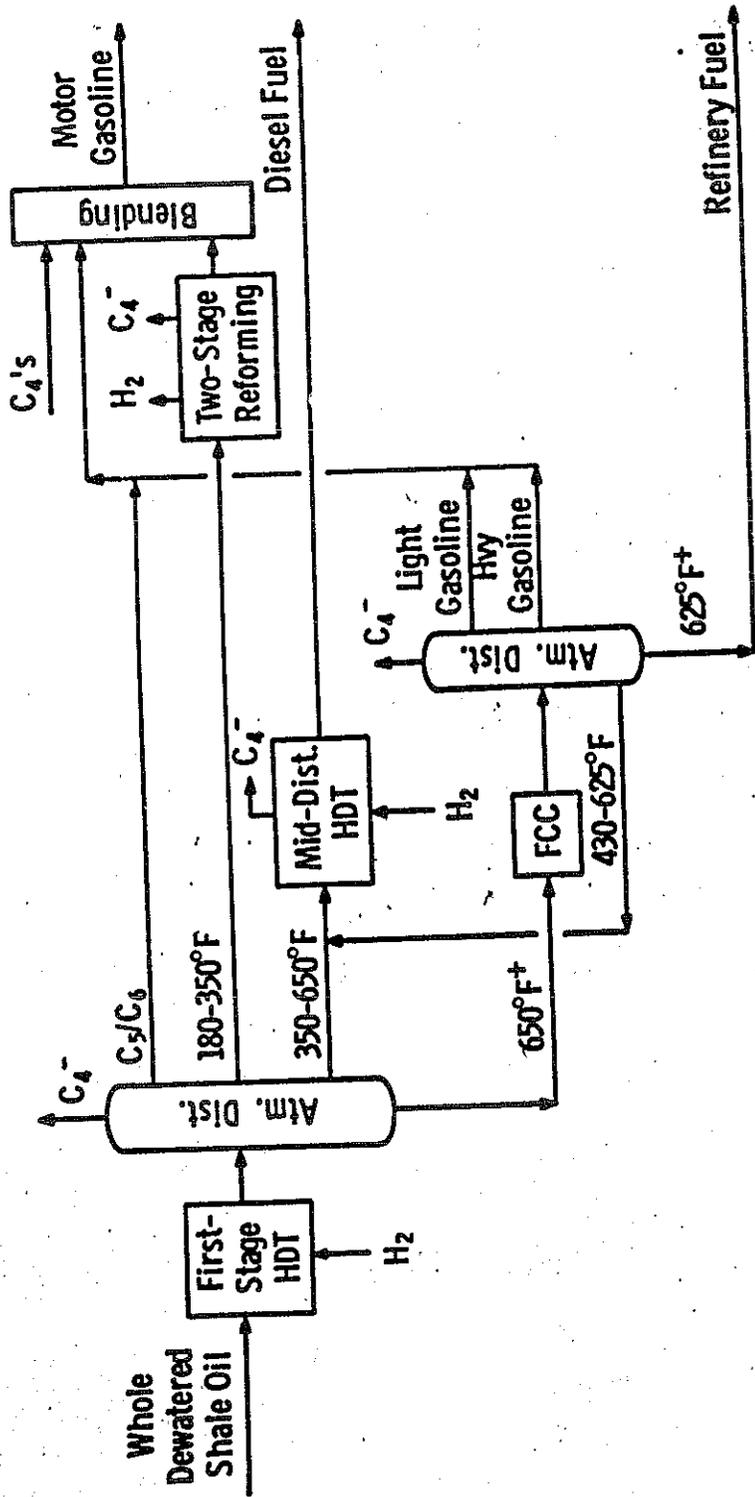
1. HDT - Hydrotreating.
2. C<sub>4</sub><sup>-</sup> is butanes and lighter. Used for H<sub>2</sub> plant feed, gasoline blending, and refinery fuel.
3. Foul gas and water treated to recover NH<sub>3</sub> and sulfur.
4. Some naphtha from first-stage HDT also used for H<sub>2</sub> plant feed.

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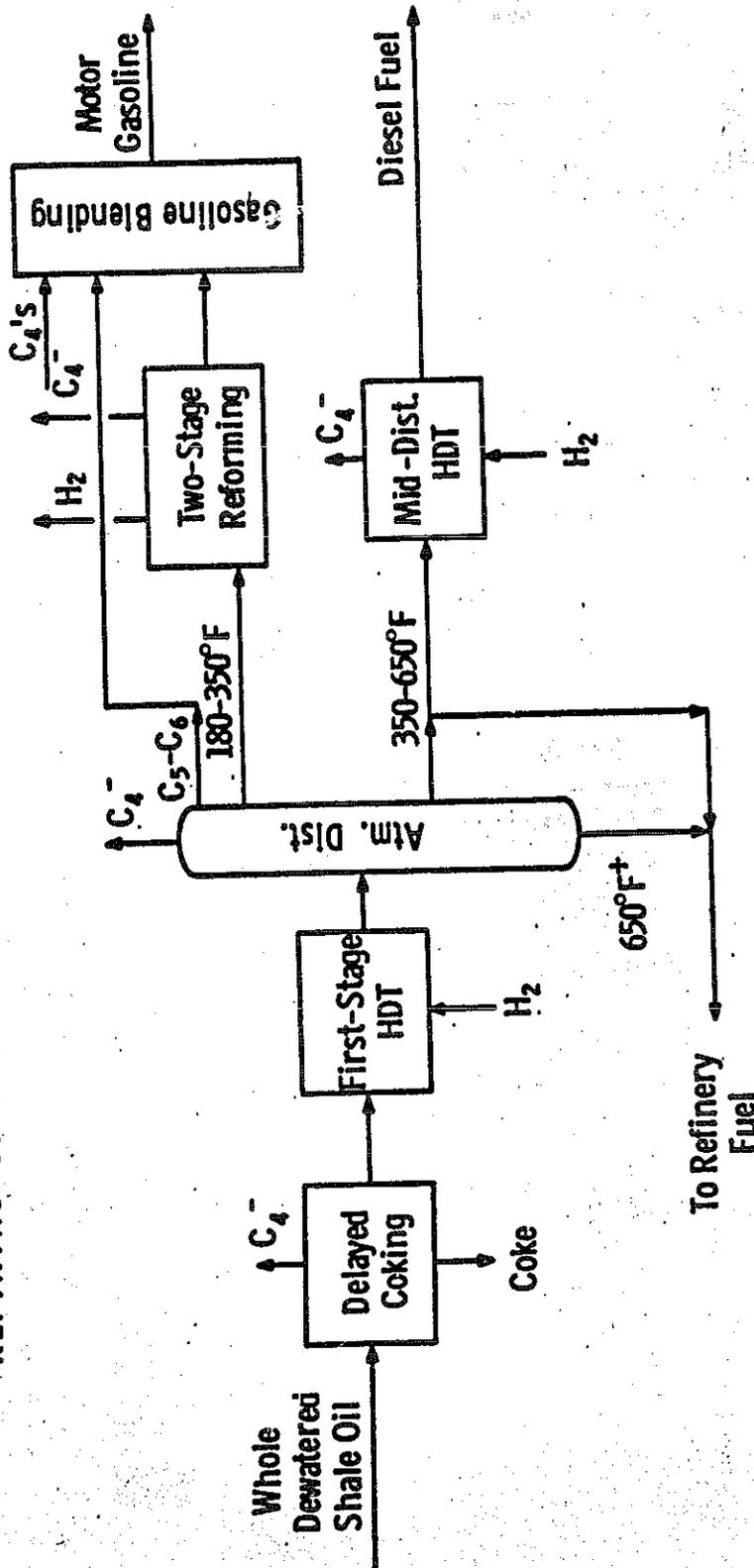
FIGURE 3  
 SCHEMATIC FLOW DIAGRAM  
 CASE 2  
 REFINING OF PARAHO SHALE OIL BY HYDROTREATING/FCC



Notes:

1. C<sub>4</sub><sup>-</sup> is butanes and lighter. Used for H<sub>2</sub> plant feed, gasoline blending, and refinery fuel.
2. Foul gas and water treated to recover NH<sub>3</sub> and sulfur.
3. HDT = Hydrotreating; FCC = Fluid Catalytic Cracking.

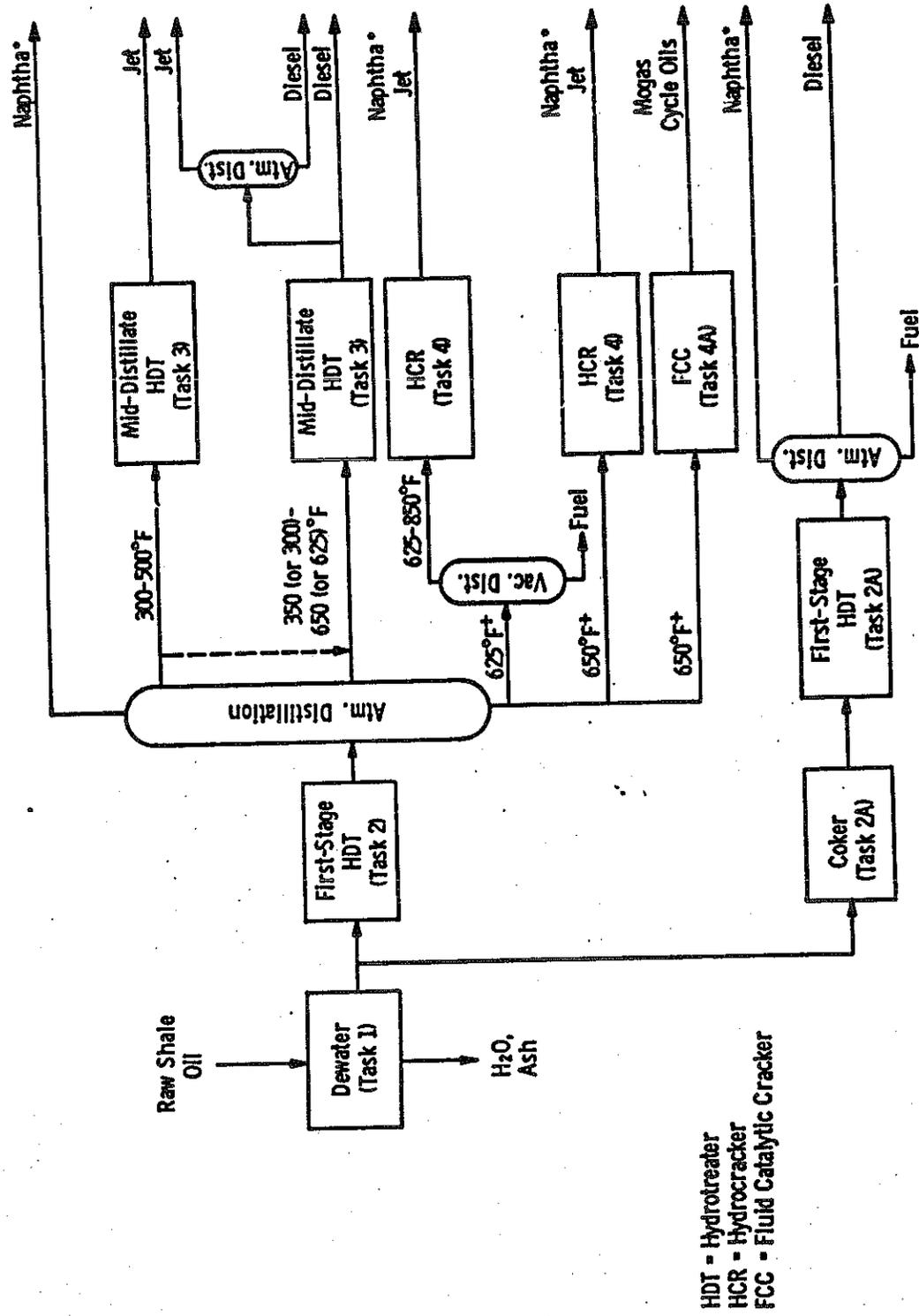
FIGURE 4  
 SCHEMATIC FLOW DIAGRAM  
 CASE 3  
 REFINING OF PARAHO SHALE OIL BY COKING-HYDROTREATING



Notes:

1. C<sub>4</sub><sup>-</sup> is butanes and lighter. Used for hydrogen plant feed, gasoline blending, refinery fuel, and liquified petroleum gas (LPG).
2. Foul gas and water treated to recover NH<sub>3</sub> and sulfur.
3. HDT - Hydrotreating.

FIGURE 5  
DOE CONTRACT EF-76-C-01-2315  
PILOT PLANT TESTS FOR PROCESSING PARAHO SHALE OIL



\*Naphtha to be Converted to High Octane Gasoline by Catalytic Reforming (Performance Predicted by Correlations)

FIGURE 6

DOE CONTRACT EF-78-C-01-2315  
HYDROTREATING OF WHOLE SHALE OIL  
COMPARISON OF WHOLE LIQUID PRODUCT NITROGEN FOR  
ICR 109, ICR 104, AND ICR 106 CATALYSTS

TEST CONDITIONS: 745°F, 0.6 LHSV, 2200 PSIG TOTAL  
(~1850 PSIA H<sub>2</sub>)  
5000 SCF/BBL RECYCLE GAS RATE  
(EXCEPT AS NOTED)

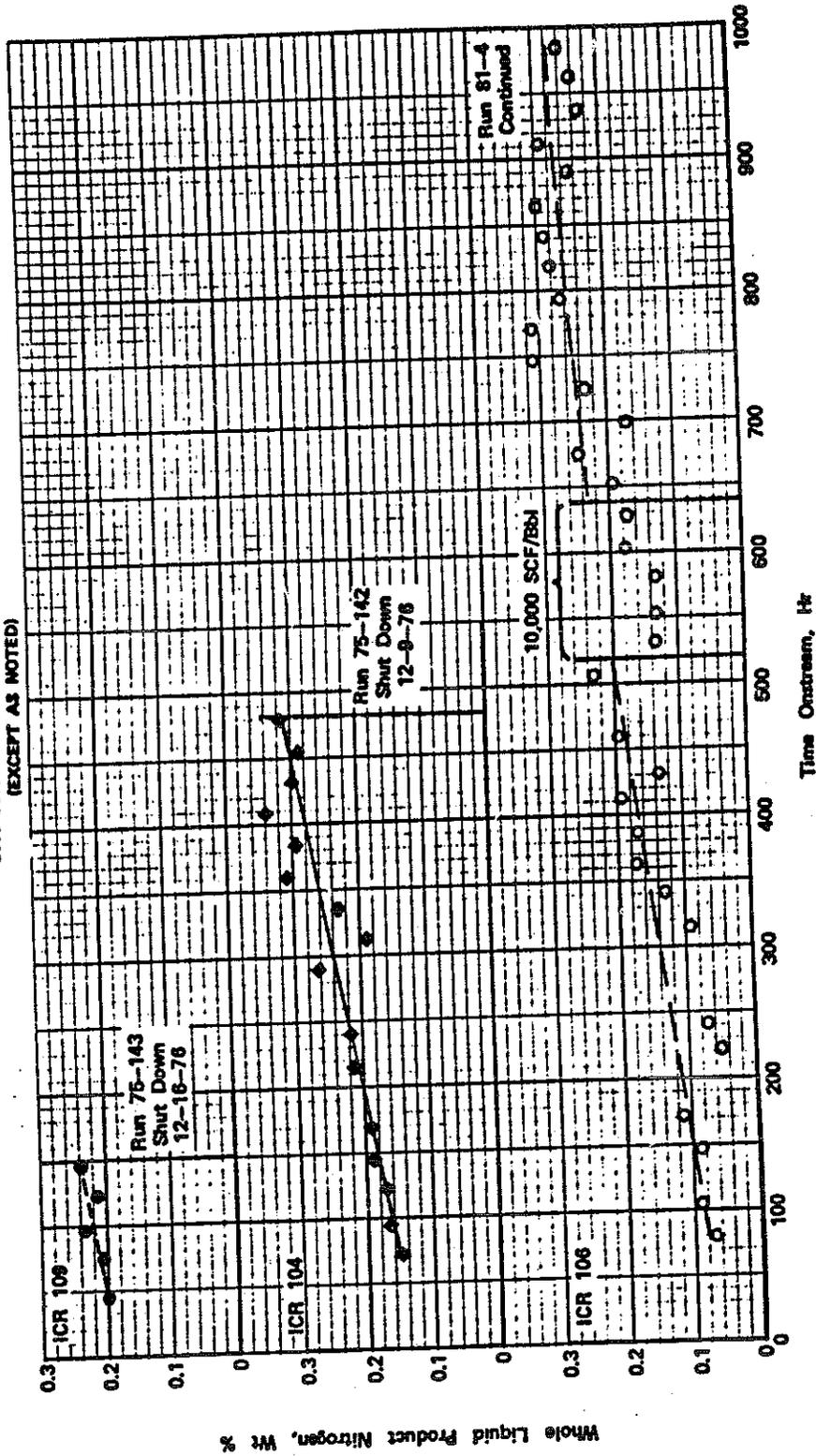


FIGURE 7

DOE CONTRACT EF-78-C-01-2315  
TWO-STAGE HYDROTREATING OF  
WHOLE SHALE OIL WITH  
ICR 106 CATALYST

TEST CONDITIONS: 745°F, EACH STAGE AT 1.2 LHSV (0.6 OVERALL LHSV)  
2200 PSIG TOTAL (~1850 PSIA H<sub>2</sub> FIRST STAGE,  
~2000 PSIA H<sub>2</sub> SECOND STAGE)  
5000 SCF/BBL RECYCLE GAS RATE

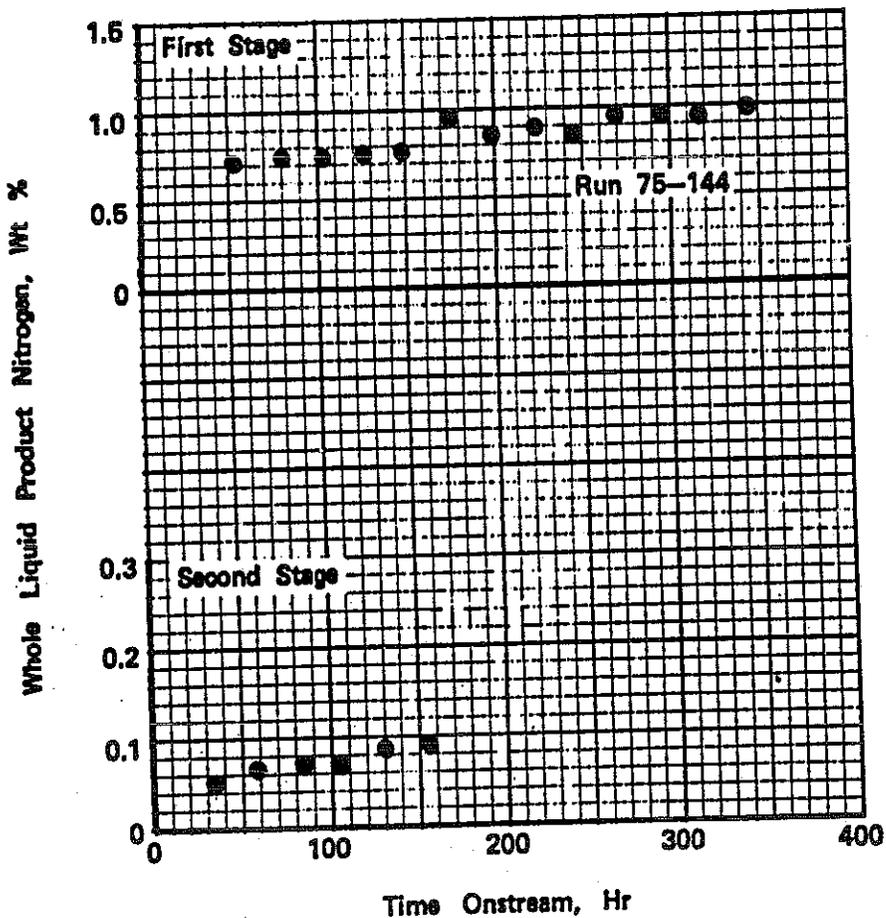


FIGURE 8  
 PILOT PLANT RUN 81-4 - PART I  
 DOE CONTRACT EF-76-C-01-2315  
 WHOLE SHALE OIL HYDROTREATING WITH ICR 106 CATALYST  
 2200 PSIG (~1850 PSIA H<sub>2</sub>) - 0.6 LHSV  
 745°F AVERAGE CATALYST TEMPERATURE

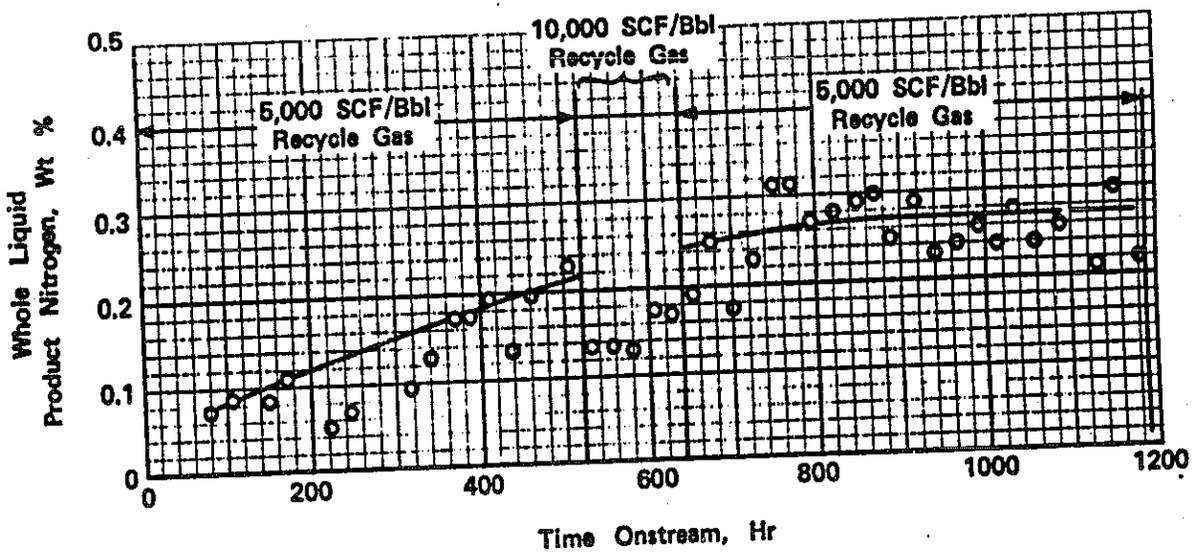
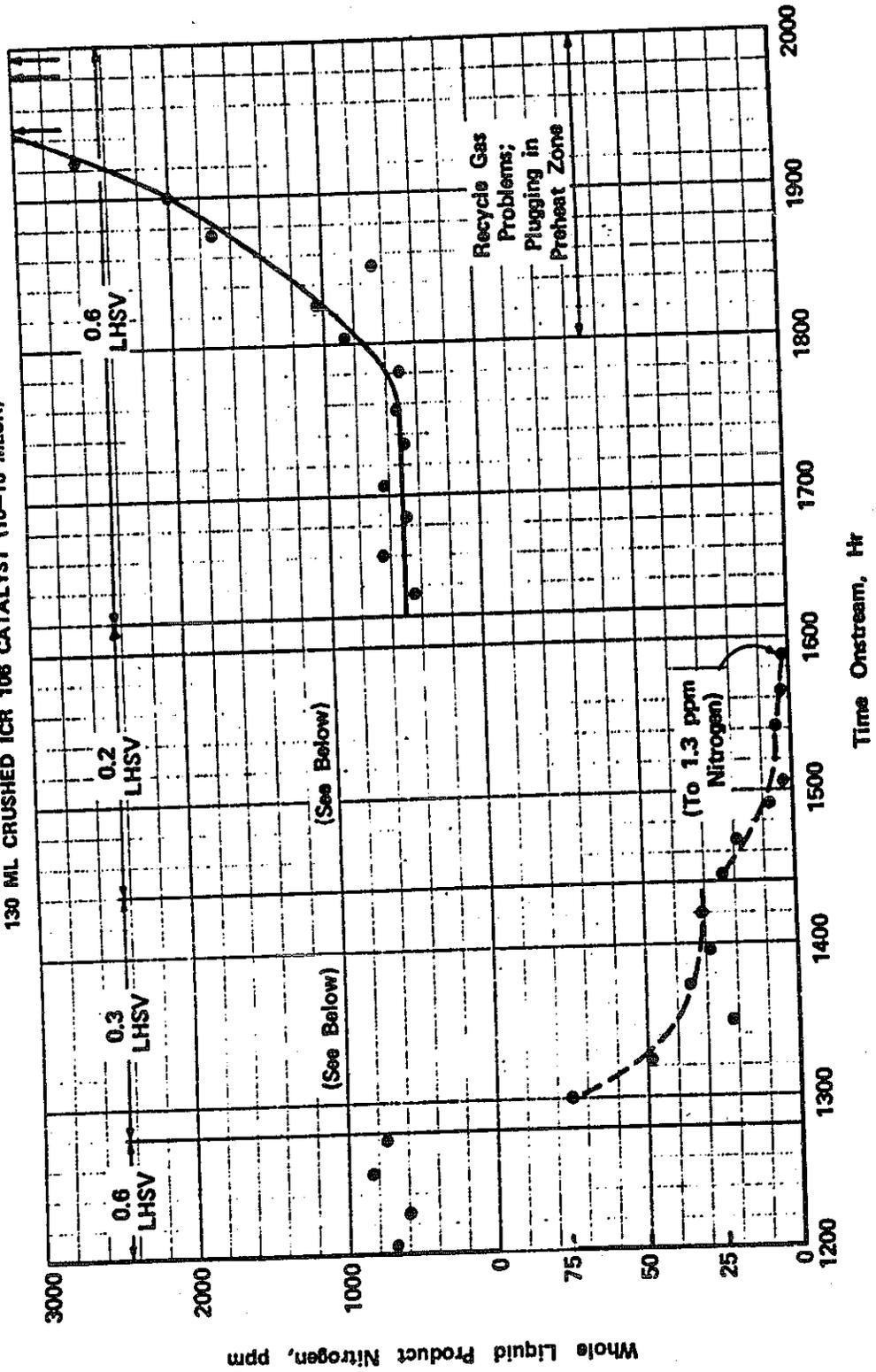


FIGURE 9

PILOT PLANT RUN 81-4 - PART II  
DOE CONTRACT EF-76-C-01-2315  
WHOLE SHALE OIL HYDROTREATING WITH ICR 106 CATALYST

767°F AVERAGE CATALYST TEMPERATURE  
2200 PSIG (~1850 PSIA H<sub>2</sub> + ~1600 PSIA H<sub>2</sub>)  
5000 SCF/BBL RECYCLE GAS RATE  
130 ML CRUSHED ICR 106 CATALYST (10-16 MESH)



**FIGURE 10**  
**PILOT PLANT RUN 81-4**  
**DOE CONTRACT EF-76-C-01-2315**  
**WHOLE SHALE OIL AND WHOLE LIQUID PRODUCTS WITH ICR 106 CATALYST**  
**767°F AVERAGE CATALYST TEMPERATURE**

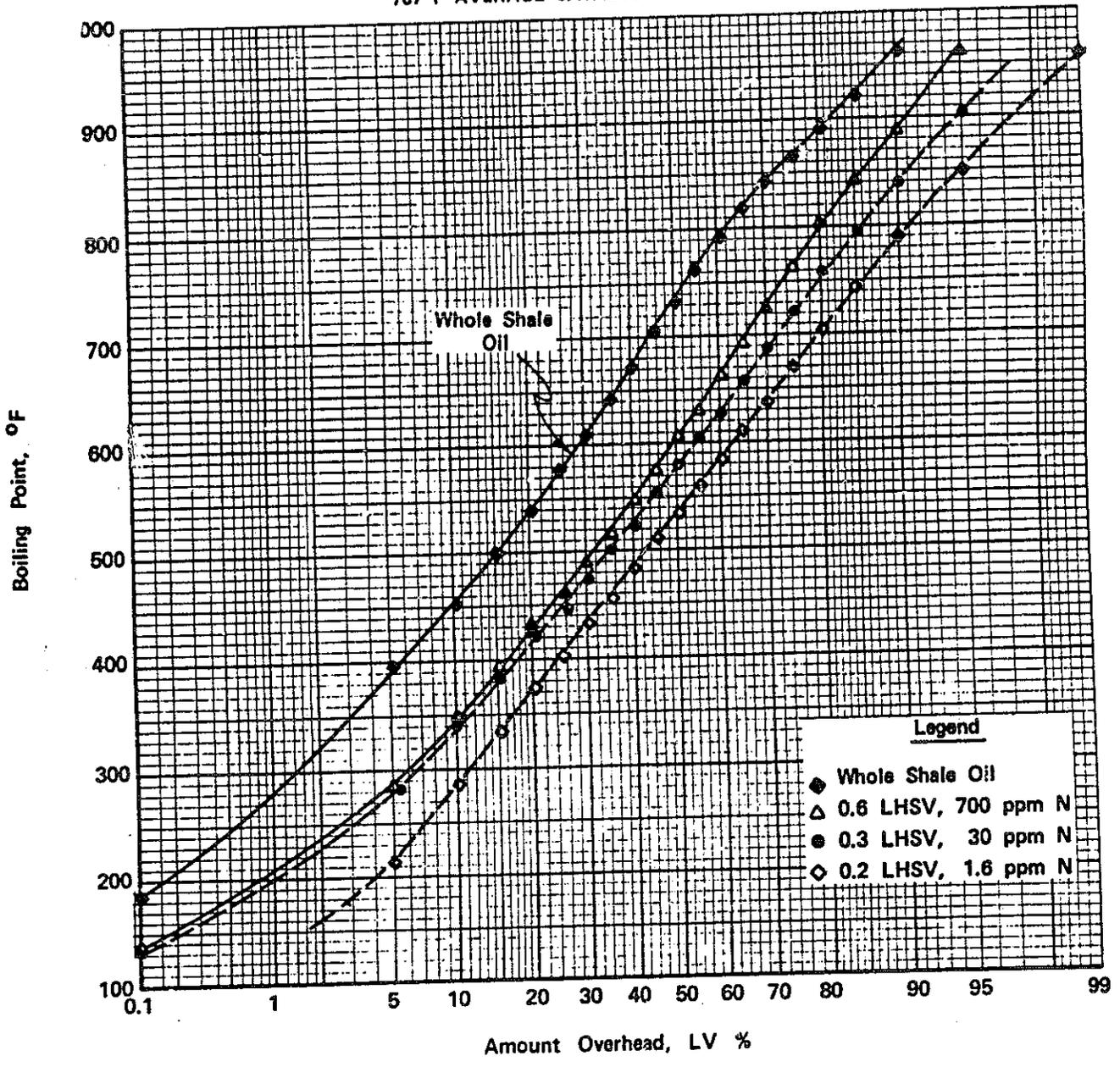
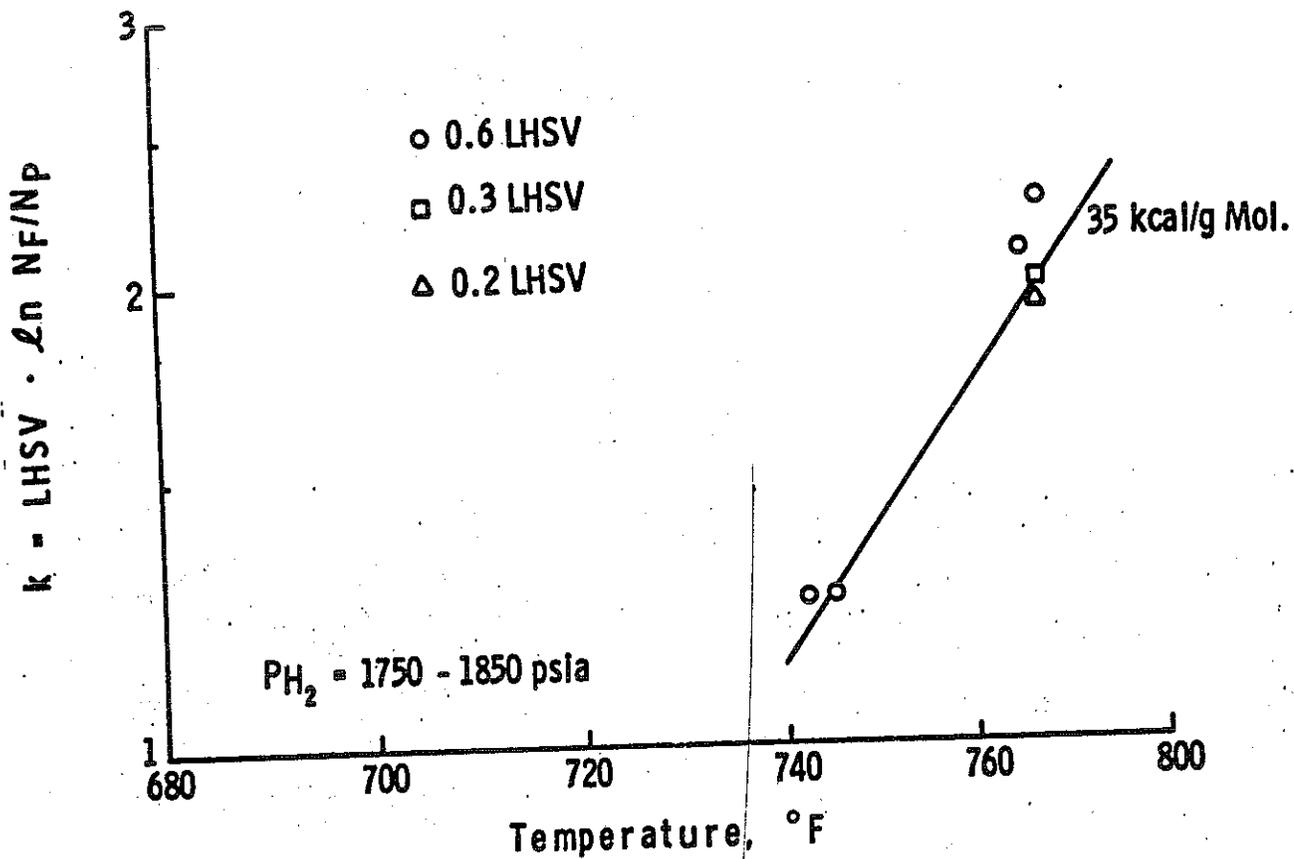
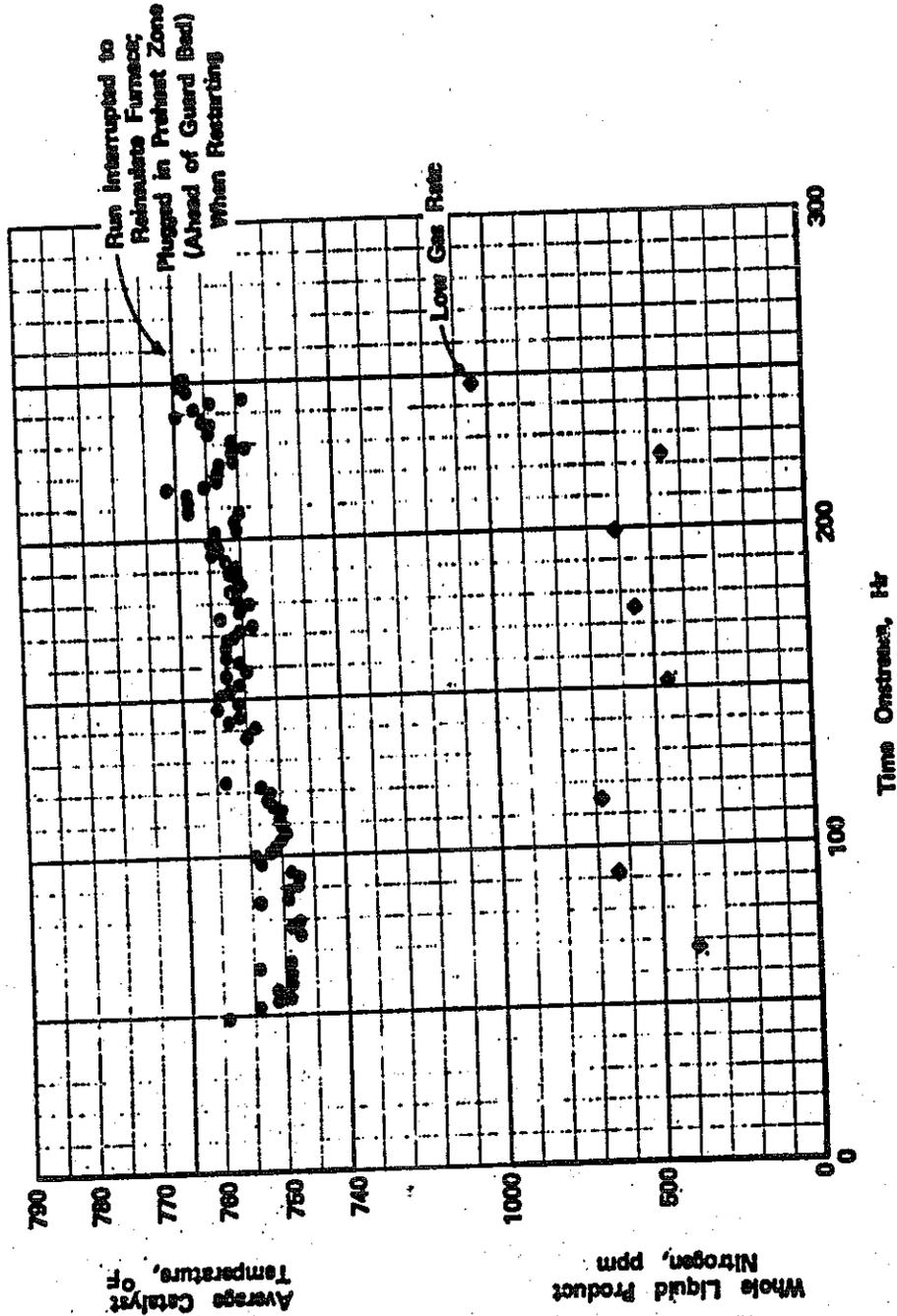


FIGURE 11  
 DOE CONTRACT EF-76-C-01-2315  
 DENITROGENATION KINETICS OF  
 WHOLE SHALE OIL ON ICR 106



**FIGURE 12**  
**PILOT PLANT RUN 86-50**  
**DOE CONTRACT EF-76-C-01-2315**  
**WHOLE SHALE OIL HYDROTREATING WITH ICR 106 CATALYST**  
 2200 PSIG (~1850 PSIA  $H_2$  + ~1600 PSIA  $H_2$ ) - 0.6 LHSV  
 5000 SCF/BBL RECYCLE GAS RATE



**FIGURE 13**  
**PILOT PLANT RUN 86-51**  
**DOE CONTRACT EF-76-C-01-2315**  
**WHOLE SHALE OIL HYDROTREATING WITH ICR 106**  
**CATALYST TEMPERATURE PROFILE AFTER 500 HR - AVERAGE CATALYST**  
**TEMPERATURE ~762°F - ~500 PPM WHOLE PRODUCT NITROGEN**

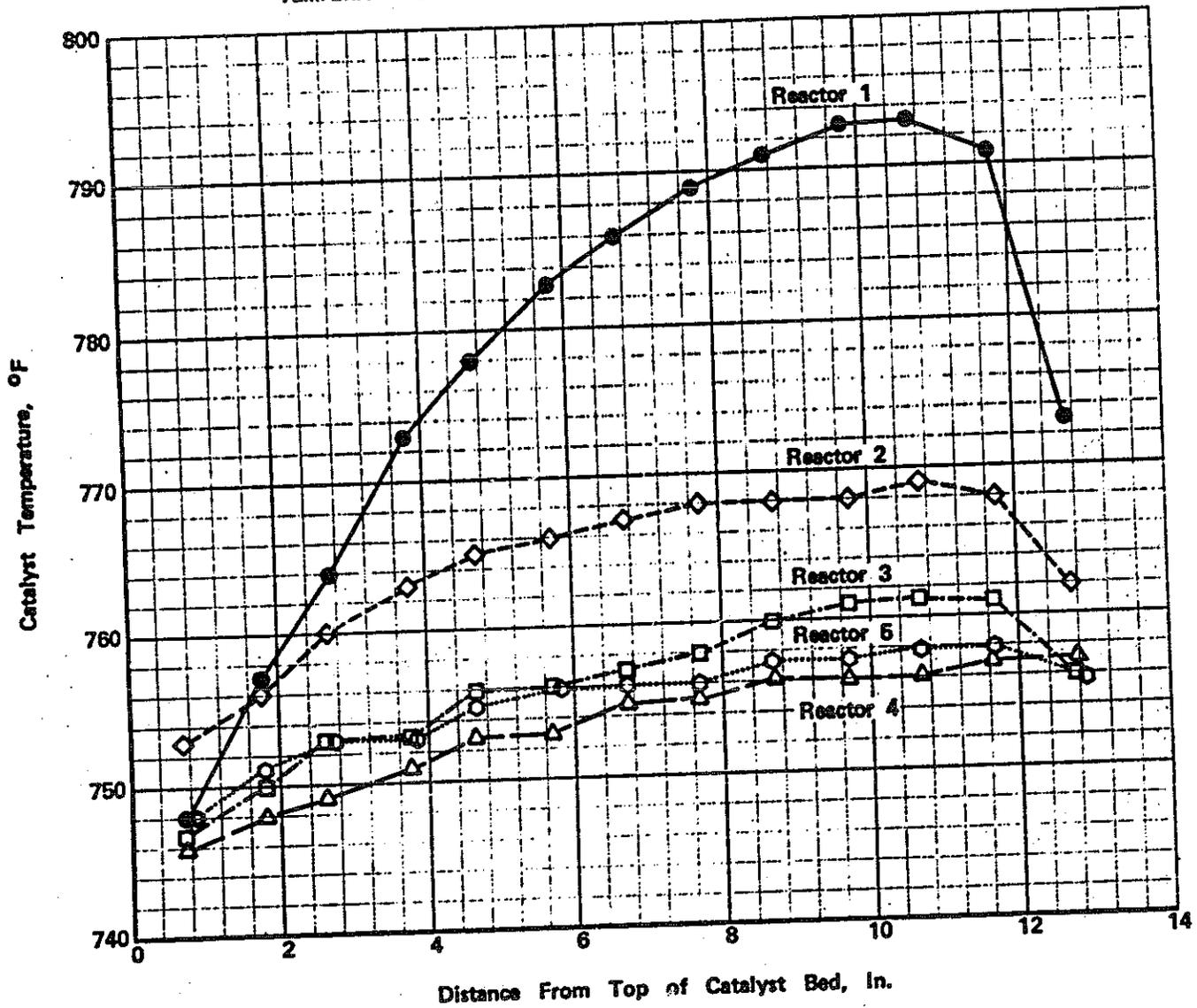


FIGURE 14  
 PILOT PLANT RUN 86-51  
 DOE CONTRACT EF-76-C-01-2315  
 WHOLE SHALE OIL HYDROTREATING WITH ICR 106  
 CATALYST TEMPERATURE PROFILE AFTER 1366 HR - AVERAGE CATALYST  
 TEMPERATURE  $\sim 787^{\circ}\text{F}$  -  $\sim 1800$  PPM WHOLE PRODUCT NITROGEN

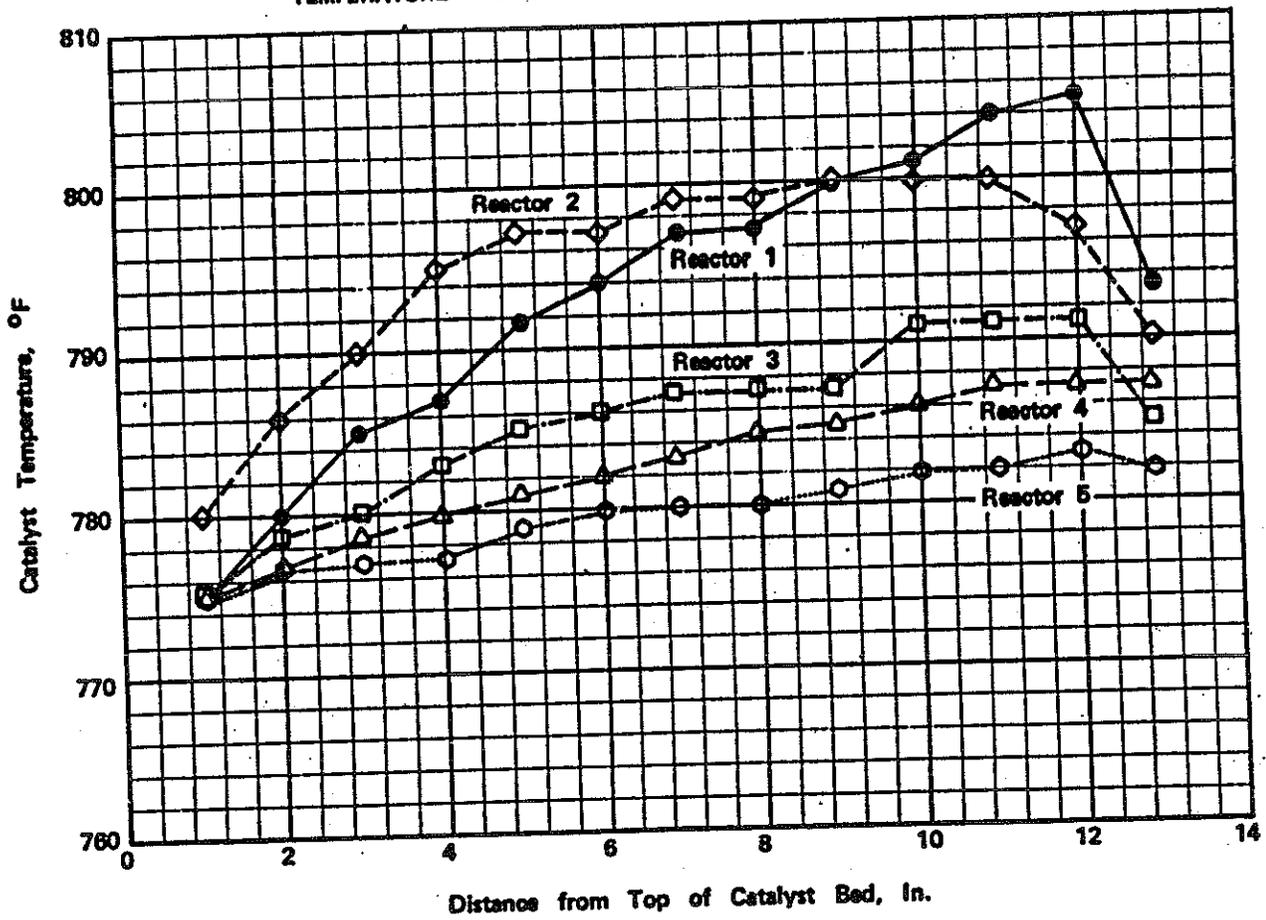
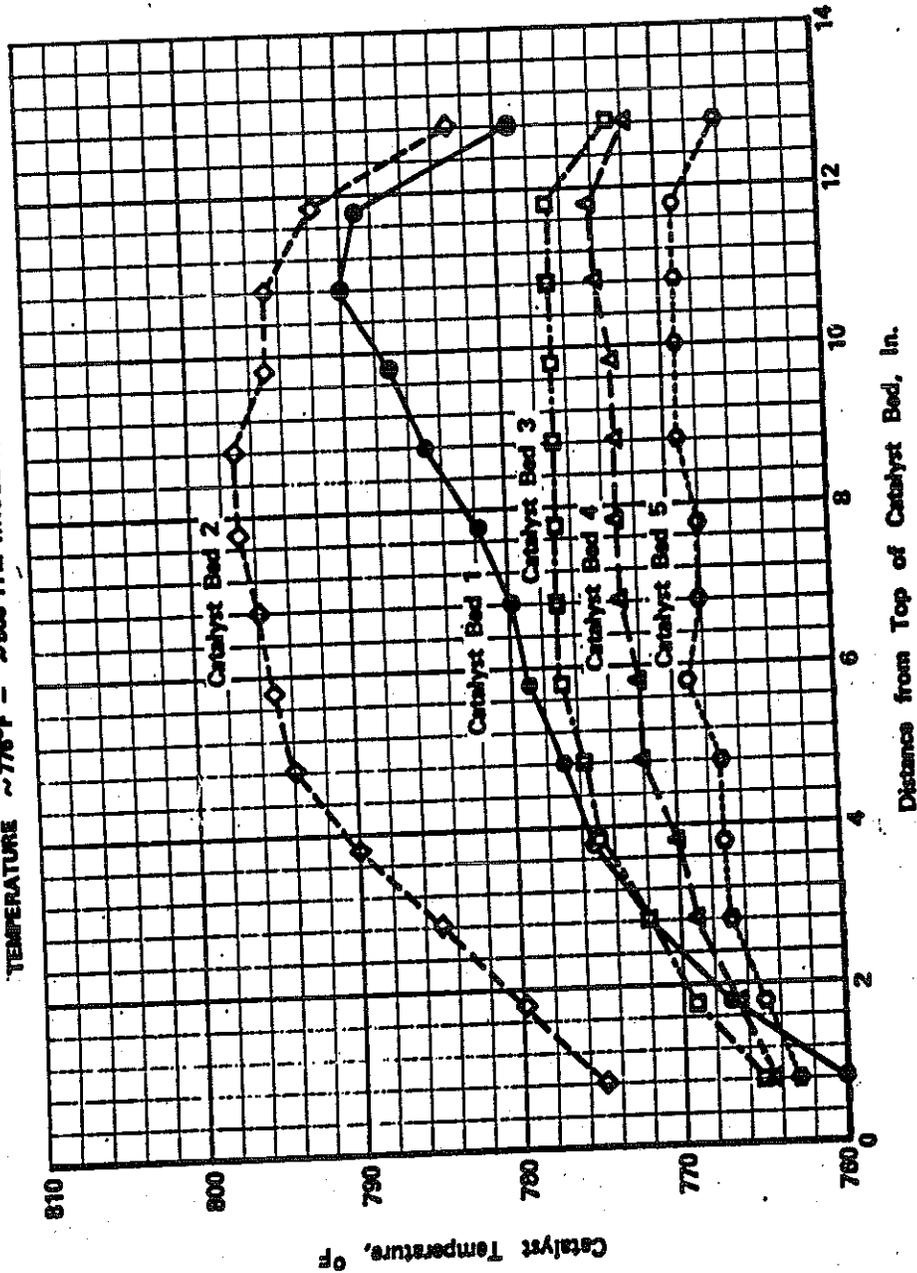


FIGURE 15

PILOT PLANT RUN 88-51  
DOE CONTRACT EF-76-C-01-2315  
WHOLE SHALE OIL HYDROTREATING WITH ICR 106

CATALYST TEMPERATURE PROFILE AT 2127 HR - AVERAGE CATALYST  
TEMPERATURE ~776°F - ~500 PPM WHOLE PRODUCT NITROGEN



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FIGURE 16

PILOT PLANT RUN 88-51  
DOE CONTRACT EF-76-C-01-2315  
WHOLE SHALE OIL HYDROTREATING WITH ICR 106  
CATALYST TEMPERATURE PROFILE AT 3039 HR  
AVERAGE CATALYST TEMPERATURE ~777°F  
~500 PPM NITROGEN IN WHOLE LIQUID PRODUCT

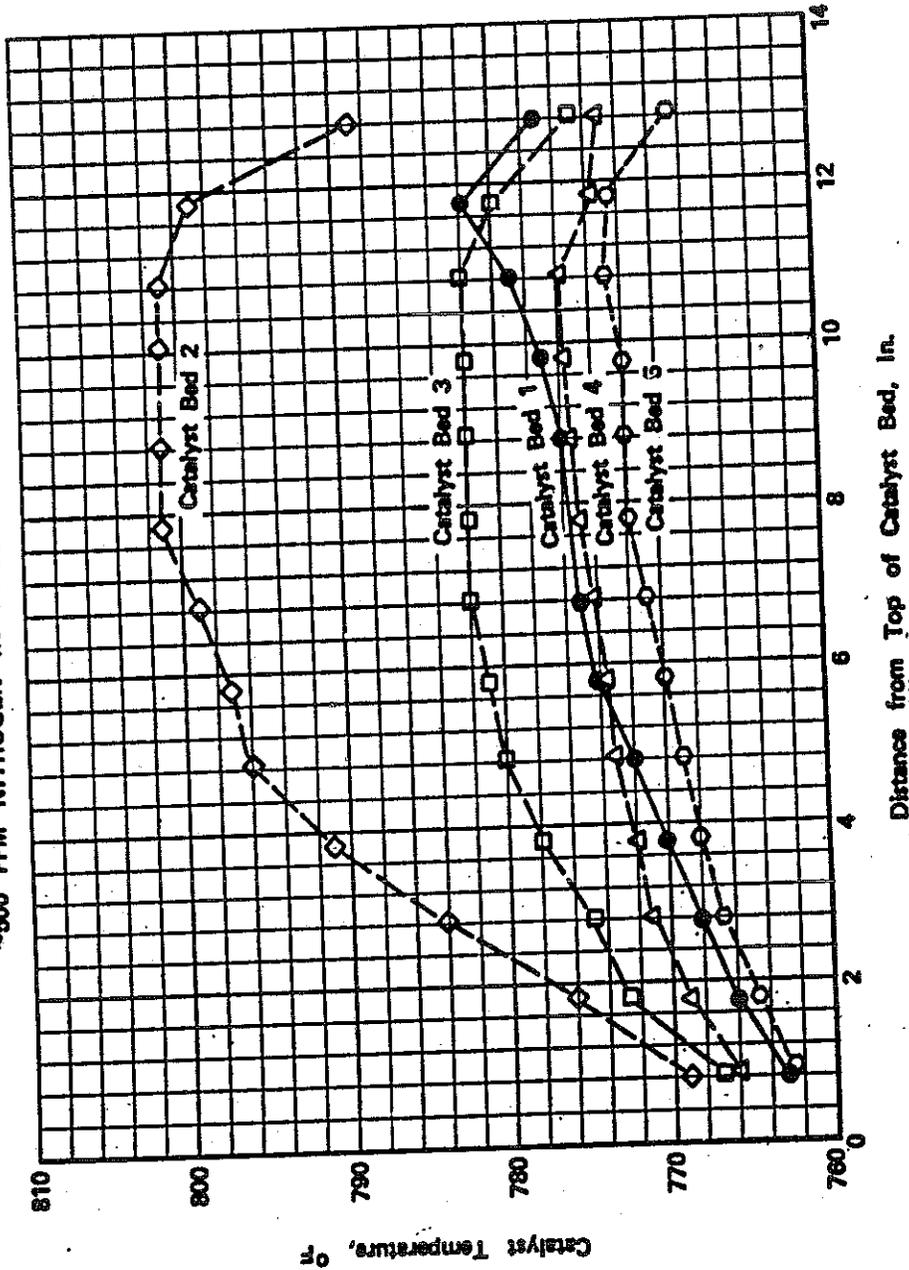


FIGURE 17

PILOT PLANT RUN 86-51  
DOE CONTRACT EF-76-C-01-2315  
WHOLE SHALE OIL HYDROTREATING WITH ICR 106

CATALYST TEMPERATURE PROFILE AT 3423 HR  
AVERAGE CATALYST TEMPERATURE ~781°F  
~800 PPM NITROGEN IN WHOLE LIQUID PRODUCT

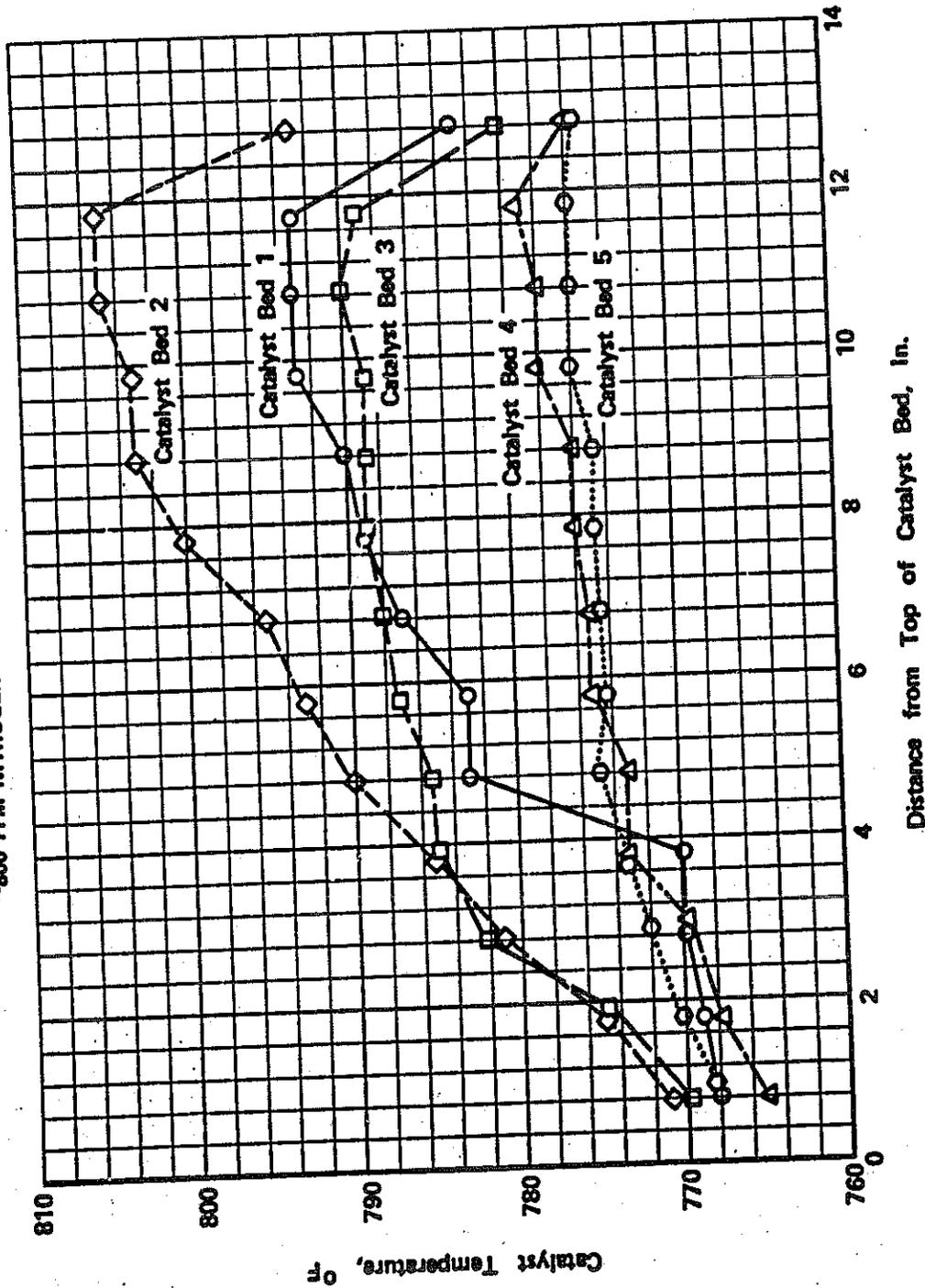


FIGURE 18

DOE CONTRACT EF-76-C-01-2315  
HYDROTREATING OF SHALE OIL WITH ICR 106  
0.6 LHSV, 8000 SCF/Bbl Recycle Gas Rate

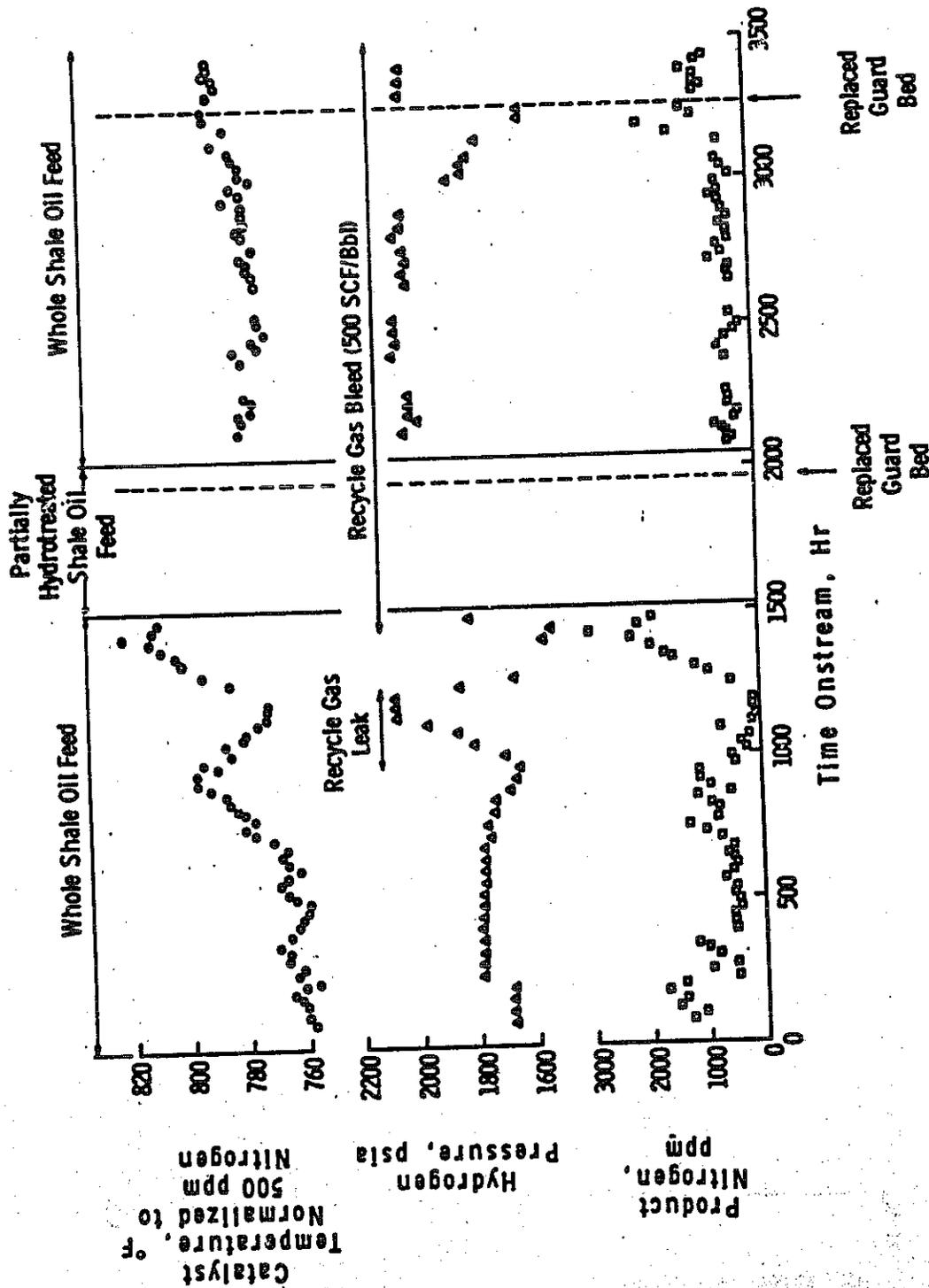


FIGURE 19

DOE CONTRACT EF-76-C-01-2315  
 COMPOSITION OF GUARD BED FROM PILOT PLANT RUN 86-51  
 AFTER 1950 HR ONSTREAM

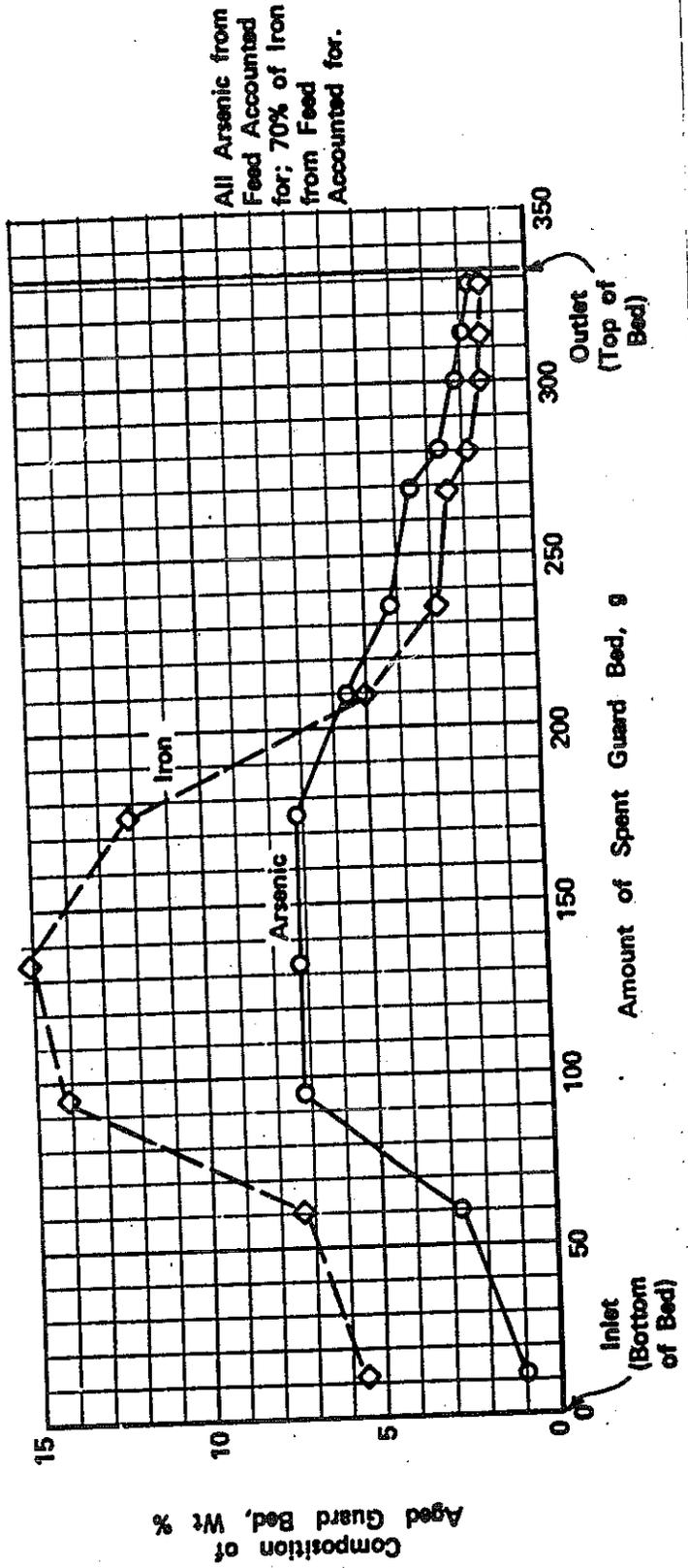
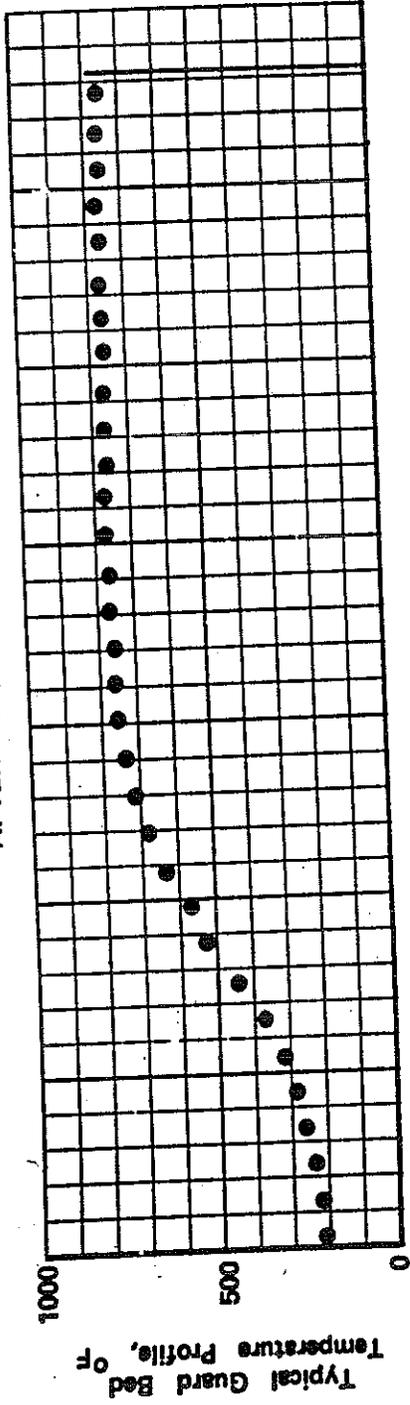


FIGURE 20

DOE CONTRACT EF-76-C-01-2315  
 COMPOSITION OF GUARD BED FROM PILOT PLANT RUN 86-51  
 AFTER 1950 HR ONSTREAM

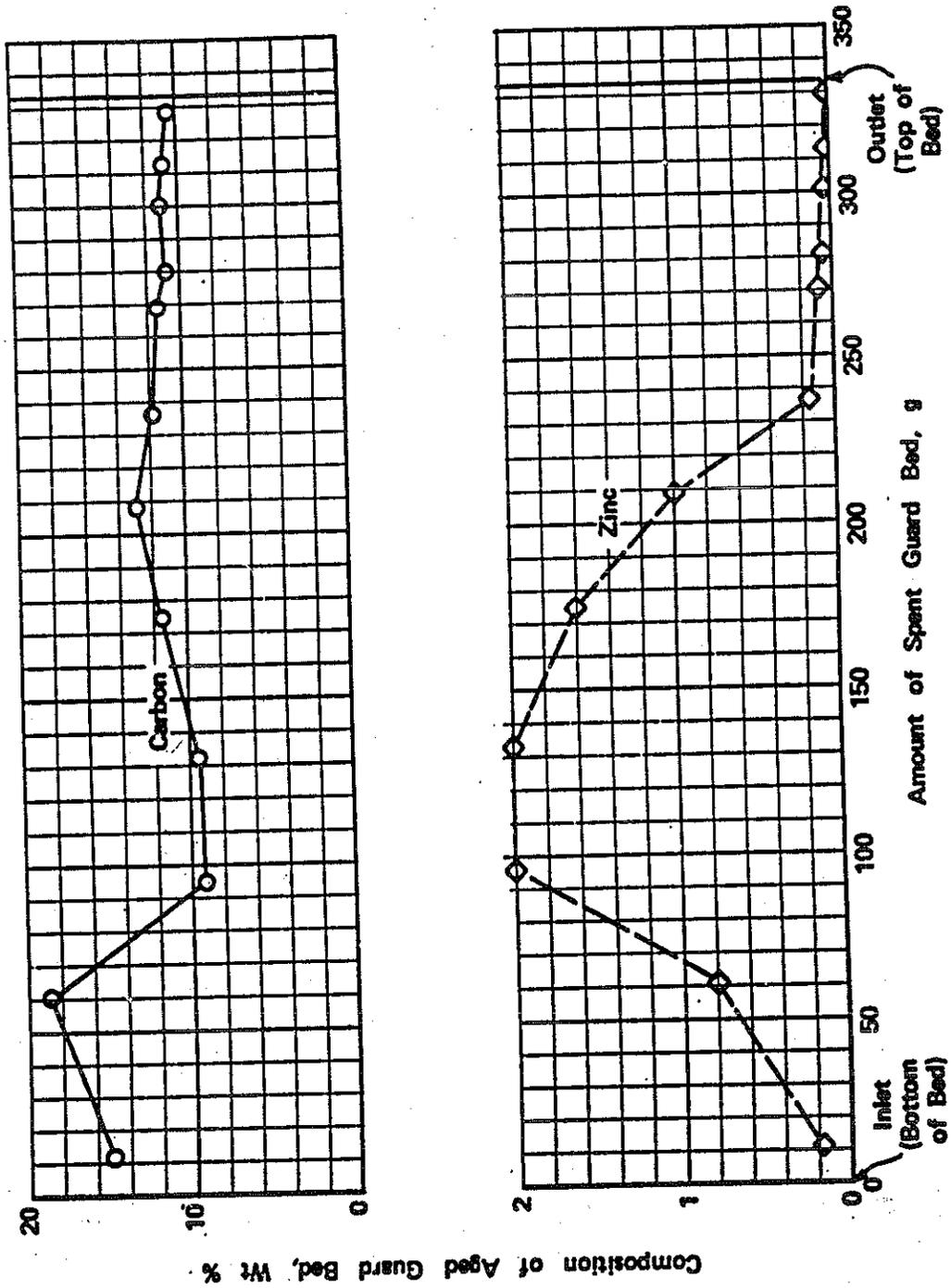


FIGURE 21  
 DOE CONTRACT EF-76-C-01-2315  
 COMPOSITION OF AGED GUARD BED FROM PILOT PLANT RUN 86-51  
 AFTER 1950 HR ONSTREAM

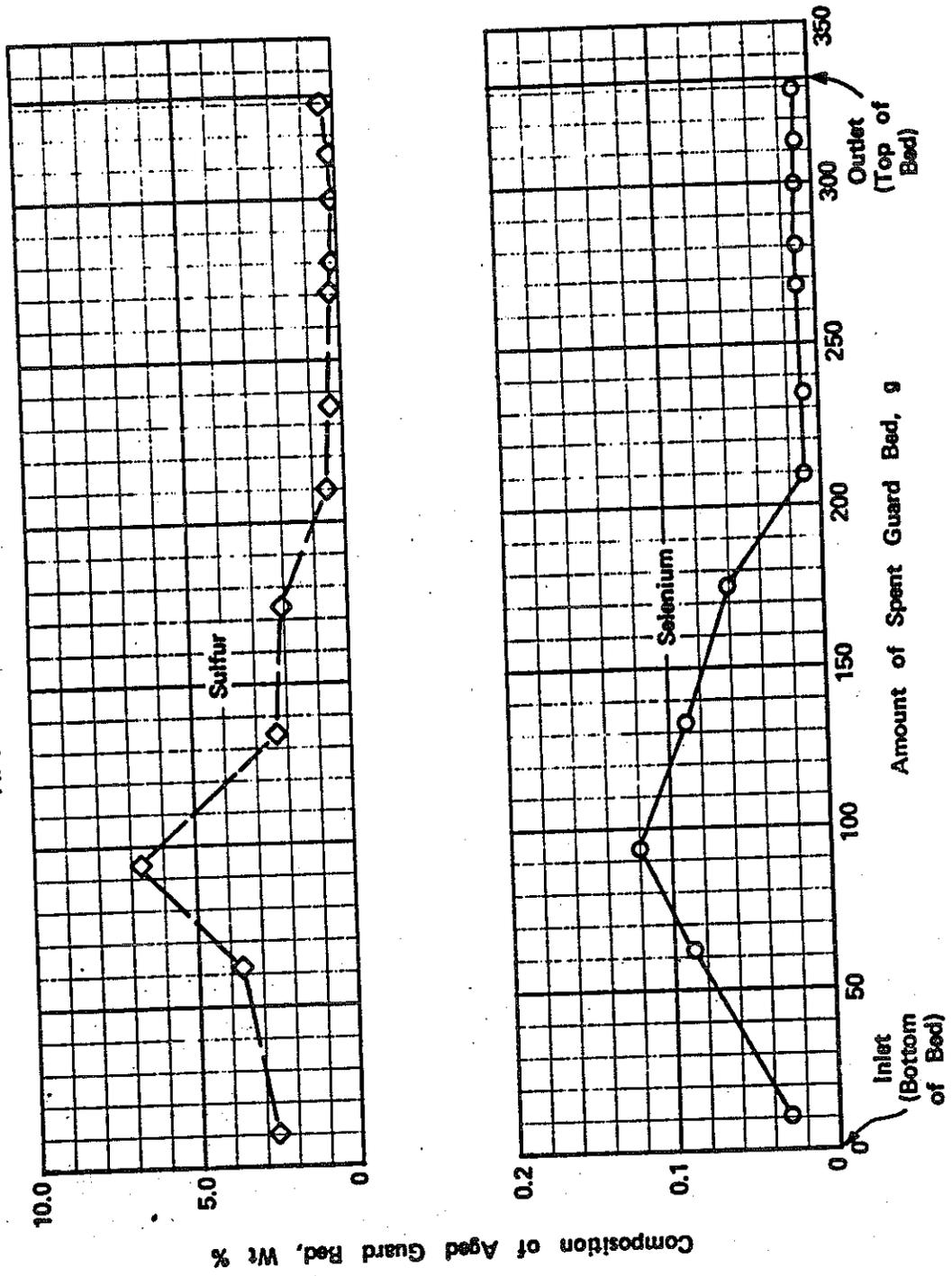


FIGURE 22

DOE CONTRACT EF-76-C-01-2315  
HYDROGEN CONSUMPTION VERSUS PRODUCT NITROGEN  
HYDROTREATING OF WHOLE SHALE OIL WITH ICR 106

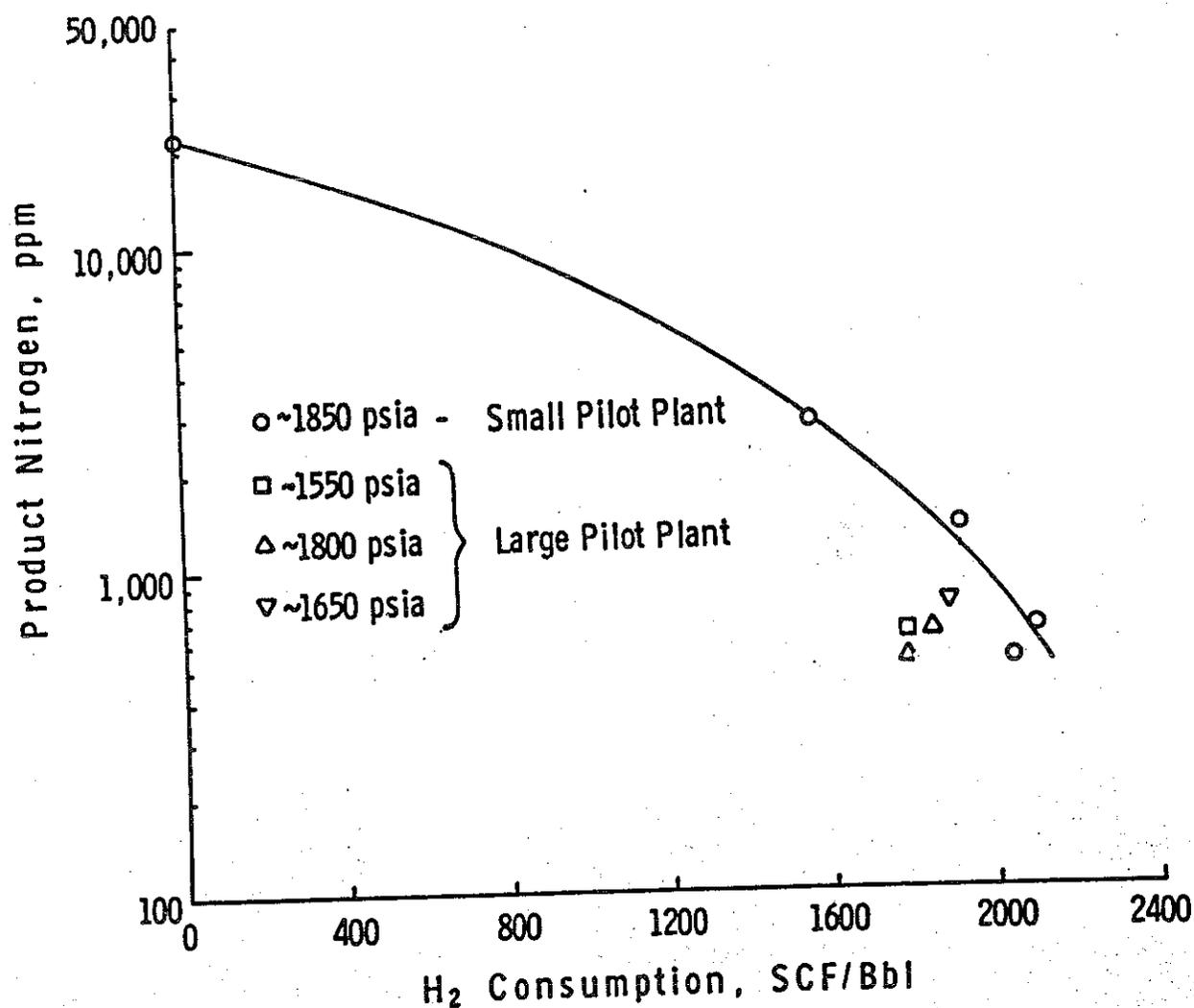


FIGURE 23  
DOE CONTRACT EF-78-C-01-2318  
COKING OF WHOLE PARAHO SHALE OIL  
EFFECT OF COKER DISTILLATE END POINT  
ON COKE YIELD

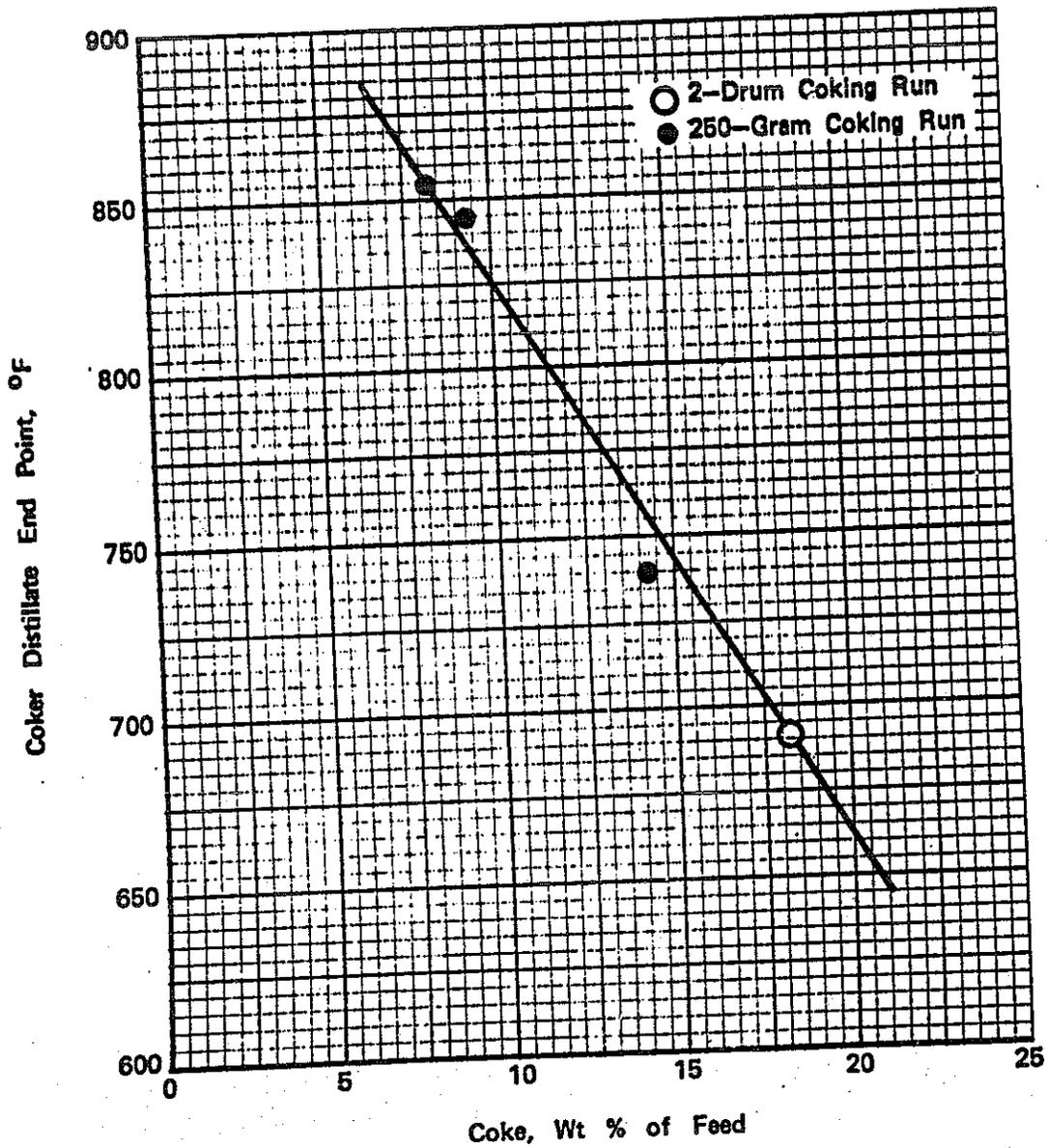


FIGURE 2A

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF SHALE OIL COKER DISTILLATE WITH ICR 106

1700 PSIG TOTAL, ~1500 PSIA H<sub>2</sub> PRESSURE  
 1.0 LHSV, 8000 SCF/BBL RECYCLE GAS RATE  
 PILOT PLANT RUN 81-10

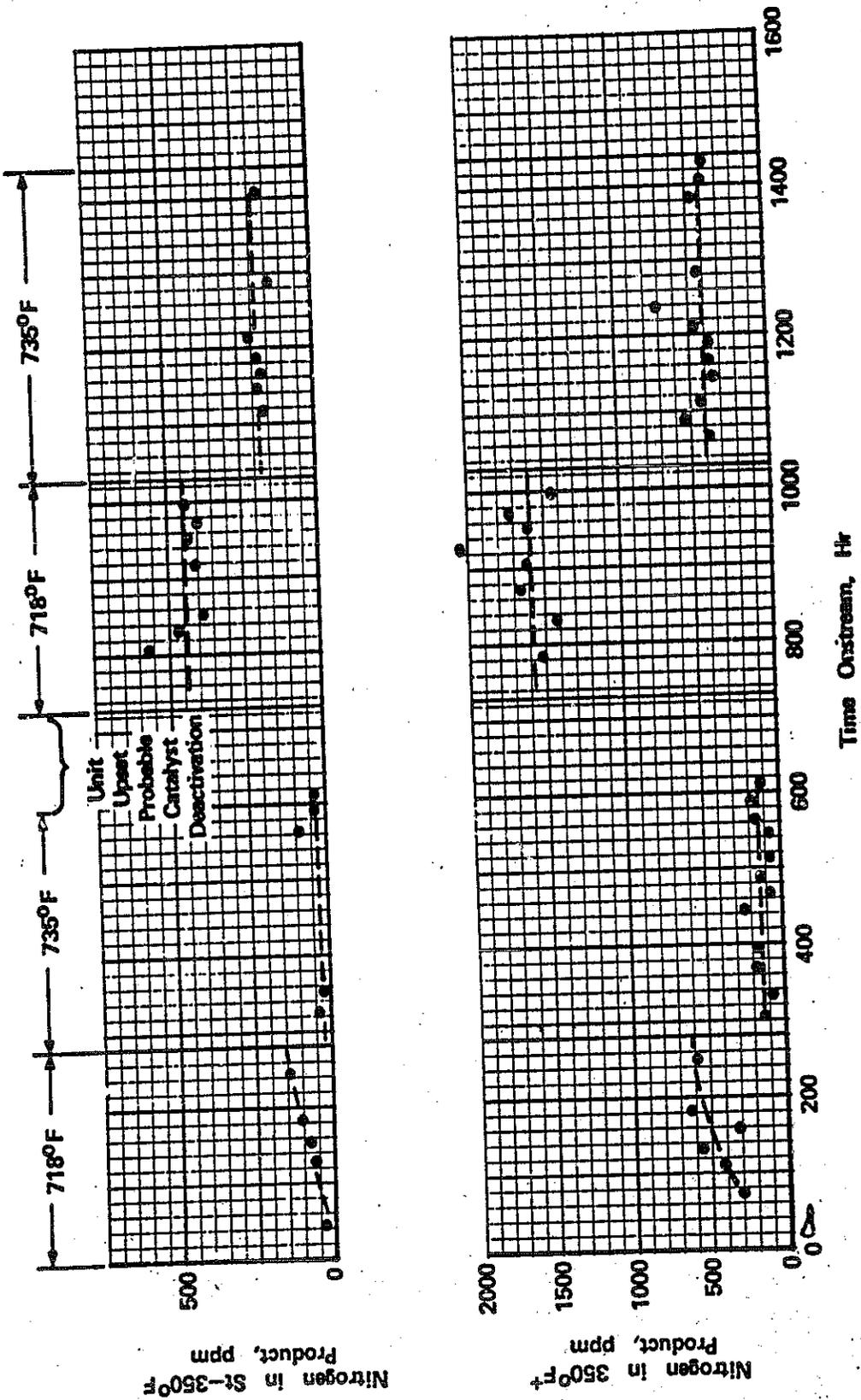


FIGURE 25

DOE CONTRACT EF-76-C-01-2315  
HYDROTREATING OF SHALE OIL COKER DISTILLATE WITH ICR 106

1700 PSIG TOTAL, ~1500 PSIA H<sub>2</sub> PRESSURE  
1.0 LHSV, 8000 SCF/BBL RECYCLE GAS RATE  
PILOT PLANT RUN 81-10

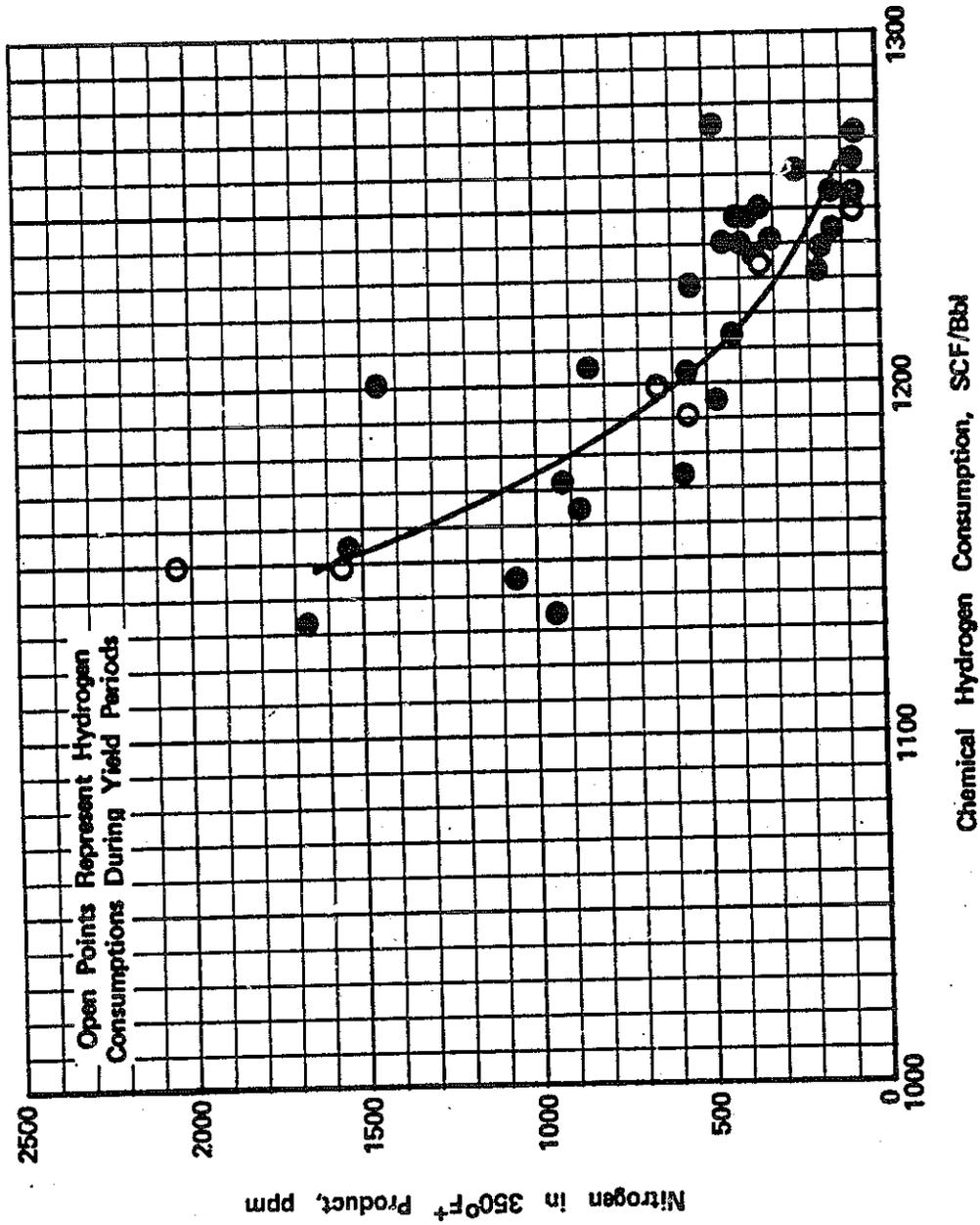
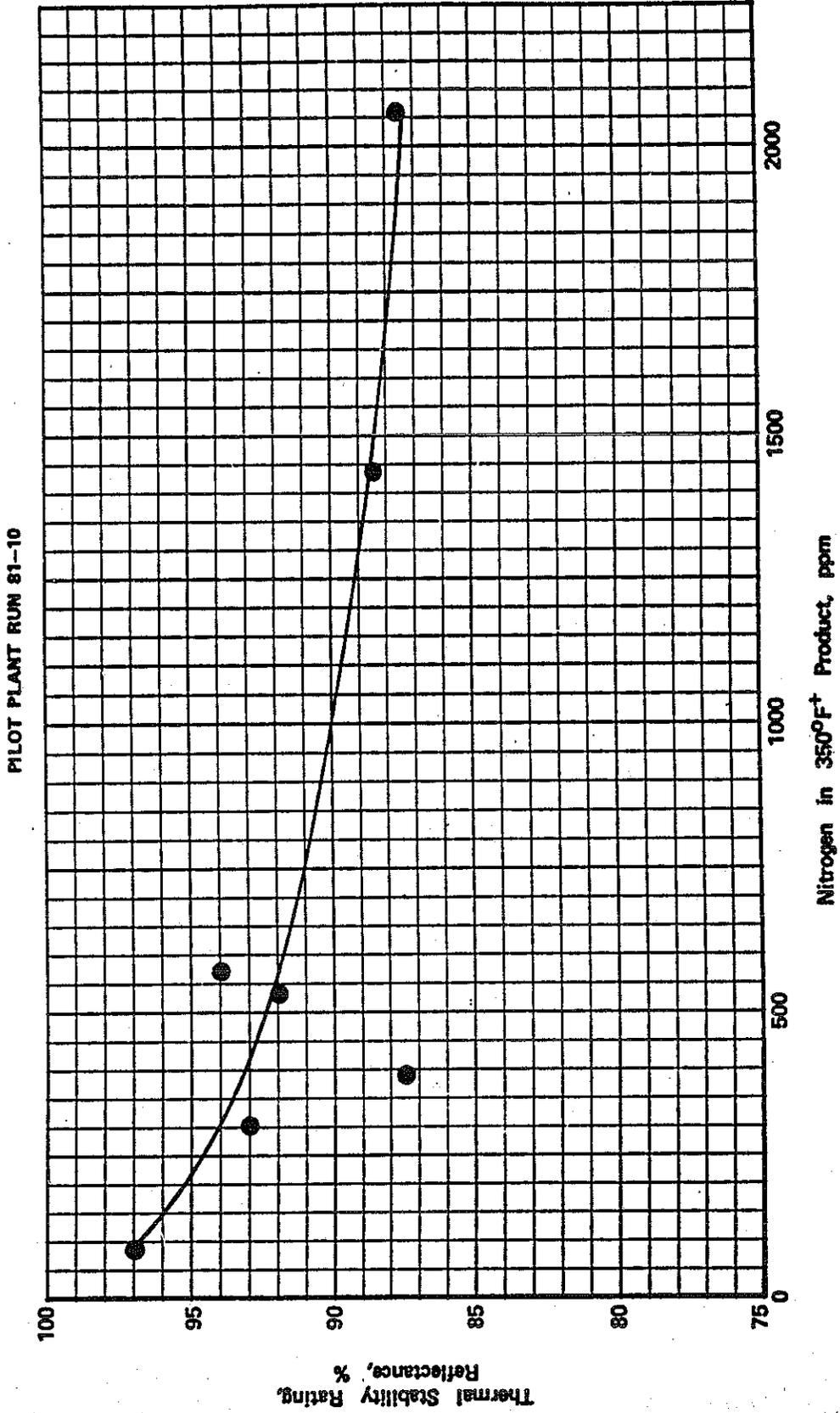


FIGURE 28

DOE CONTRACT EF-76-C-01-2315  
EFFECT OF NITROGEN ON THERMAL STABILITY OF 350°F<sup>+</sup> PRODUCT  
HYDROTREATING OF SHALE OIL COKER DISTILLATE



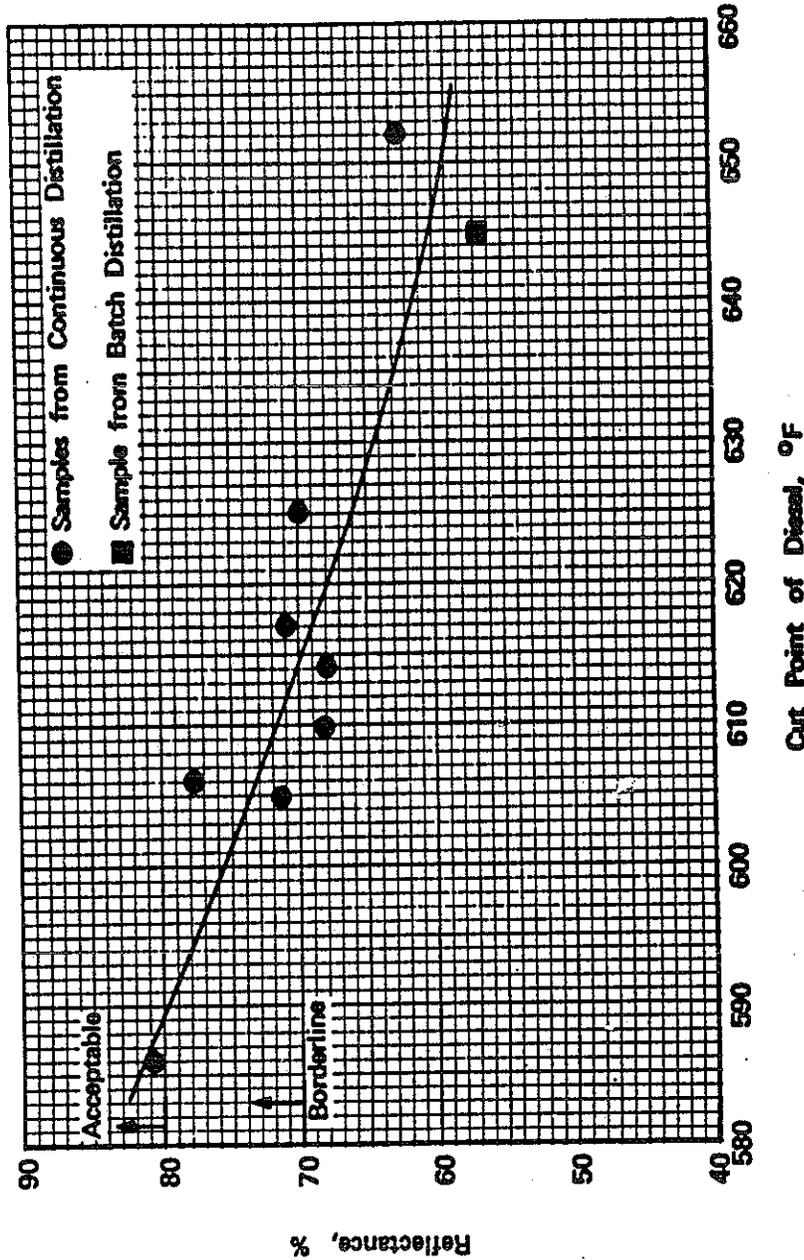
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FIGURE 27

DOE CONTRACT EF-76-C-01-2315  
PROPERTIES OF DIESEL CUT FROM HYDROTREATED SHALE OIL  
THERMAL STABILITY OF DIESEL\* AS A FUNCTION OF DIESEL CUT POINT

DIESEL CUT POINT IS DEFINED AS THE 95% POINT TEMPERATURE AS  
DETERMINED BY CHROMATOGRAPHIC SIMULATED TBP



\*350-400 ppm Nitrogen in Diesel Fraction

FIGURE 28

DOE CONTRACT EF-76-C-01-2315  
PROPERTIES OF DIESEL CUT FROM HYDROTREATED SHALE OIL  
EFFECT OF DIESEL CUT POINT ON  
FREEZE AND POUR POINT

CUT POINT IS DEFINED AS THE 96% POINT TEMPERATURE AS  
DETERMINED BY CHROMATOGRAPHIC SIMULATED TBP

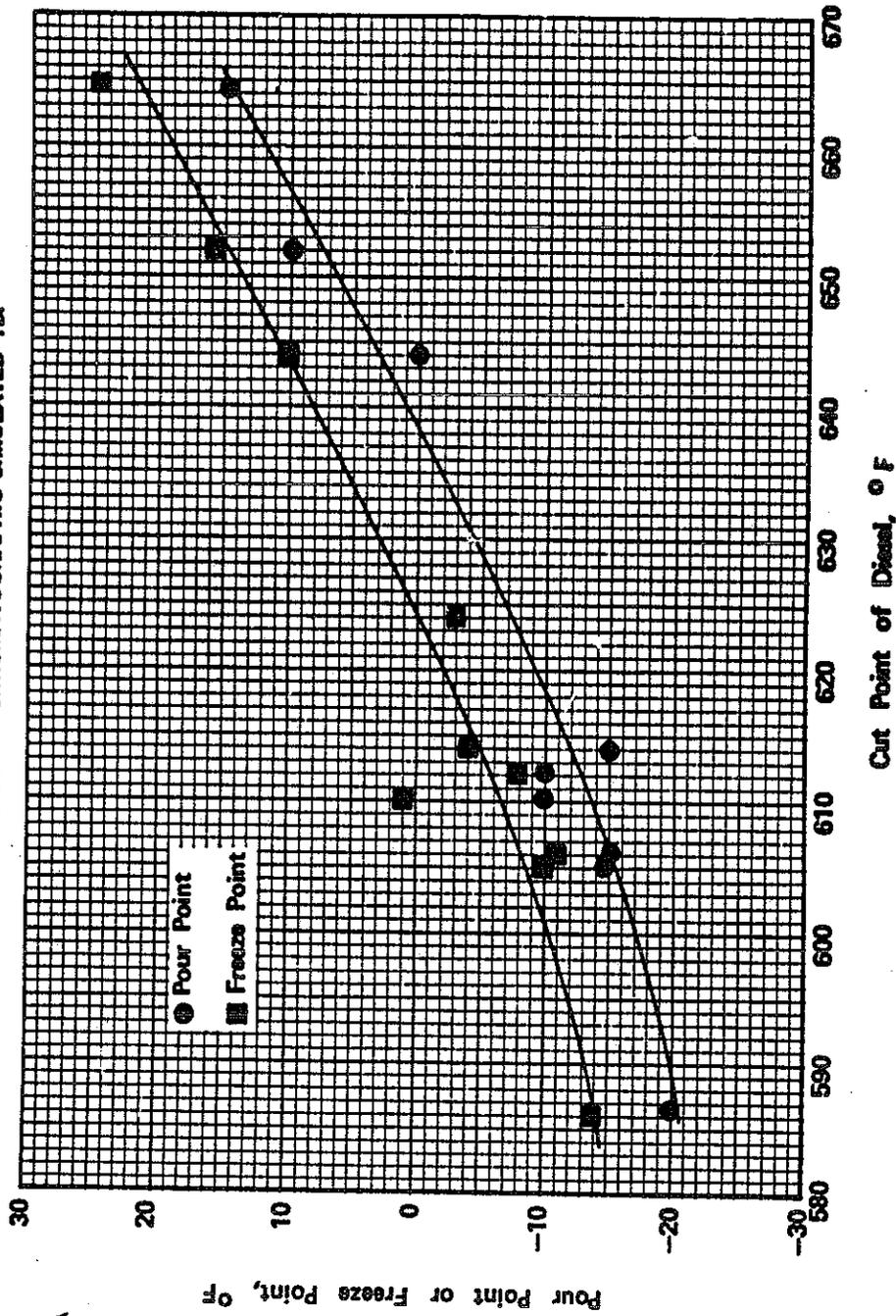


FIGURE 29

DOE CONTRACT EF-76-C-01-2315  
HYDROFINISHING OF DIESEL FRACTION FROM  
HYDROTREATED SHALE OIL

SGO 6198, 2.0 LHSV, 1200 PSIG,  
5000 SCF/BBL RECYCLE H<sub>2</sub>, ICR 106  
RUN 66-185

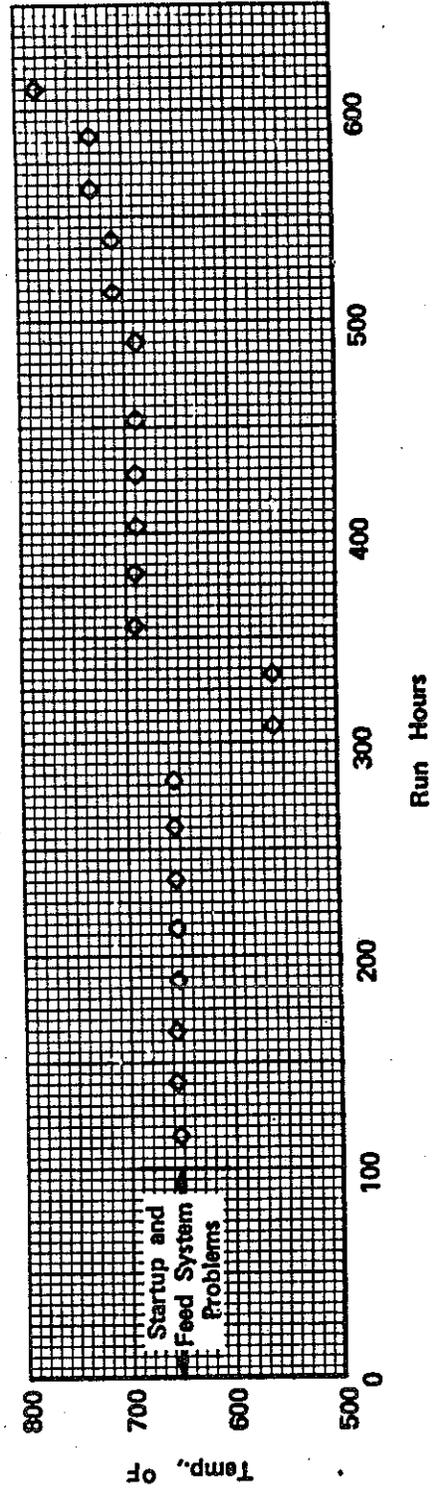
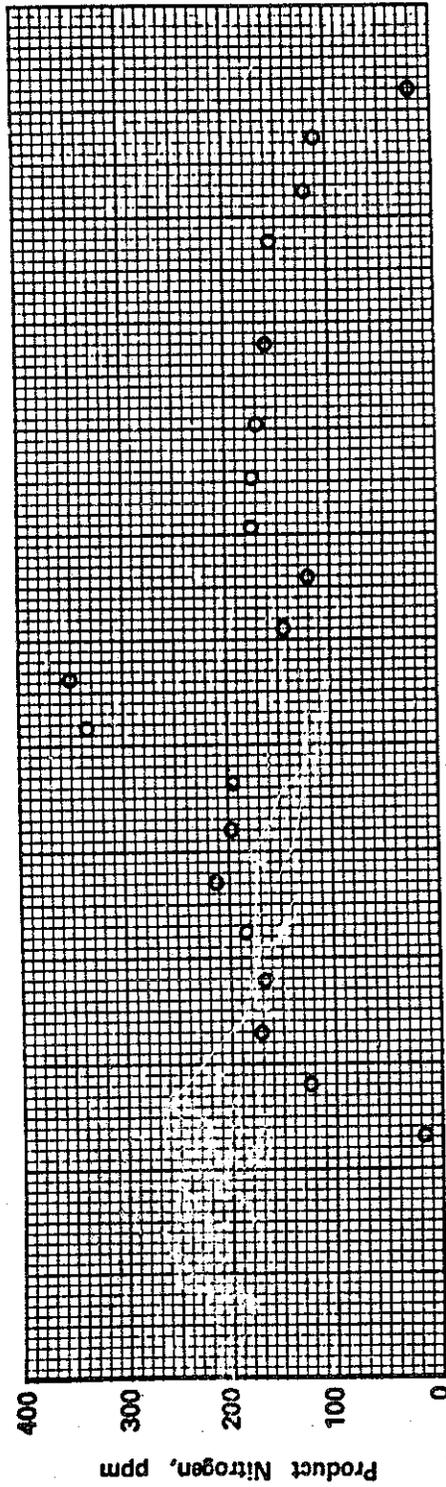


FIGURE 30

DOE CONTRACT EF-76-C-01-2315  
HYDROFINISHING OF DIESEL AND JET FRACTIONS  
FROM HYDROTREATED SHALE OIL

2.0 LHSV, 1200 PSIG, 5000 SCF/BBL  
ONCE-THROUGH H<sub>2</sub> ICR 108  
RUN 66-187

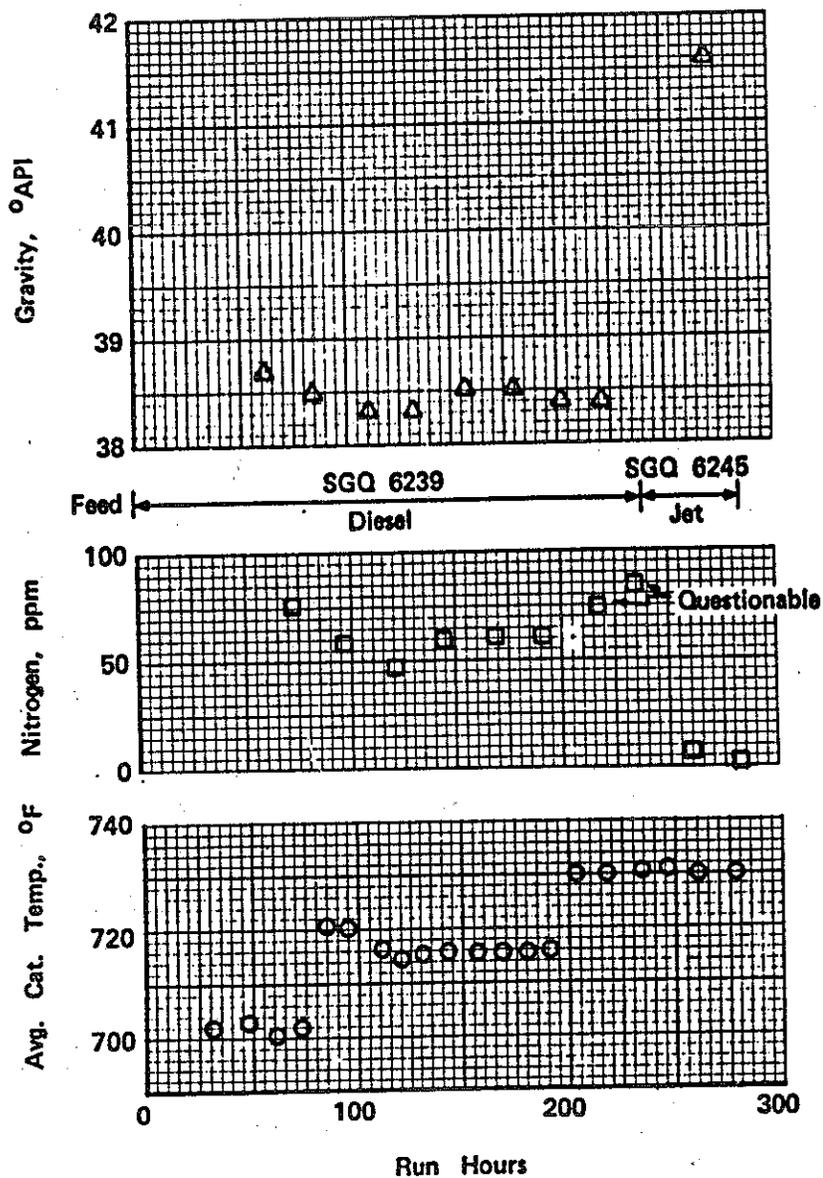
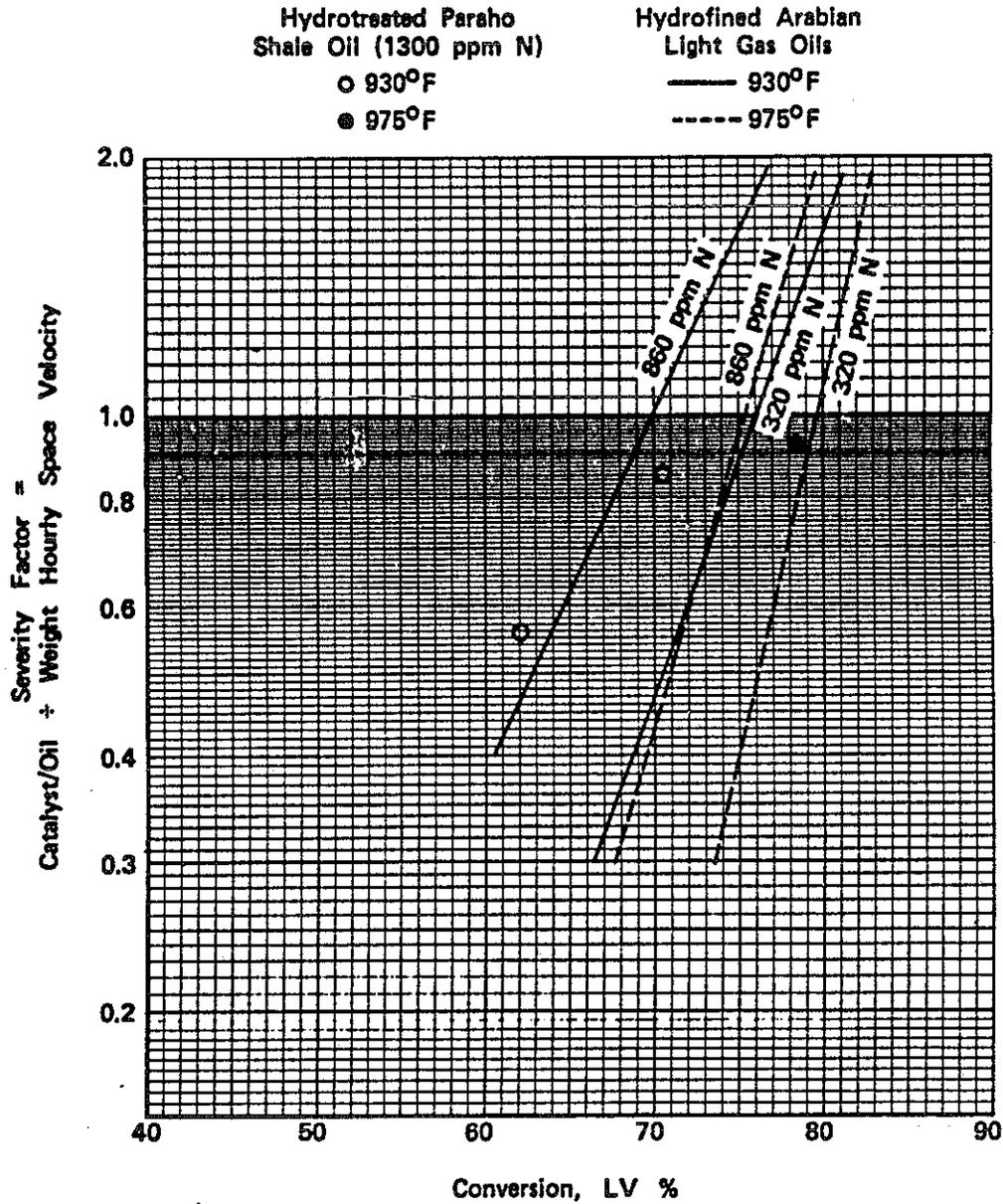


FIGURE 31  
 DOE CONTRACT EF-78-C-01-2315  
 CATALYTIC CRACKING OF  
 HYDROTREATED PARAHO SHALE OIL  
 SEVERITY VERSUS CONVERSION



**FIGURE 32**  
**DOE CONTRACT EF-78-C-01-2315**  
**CATALYTIC CRACKING OF**  
**HYDROTREATED PARAHO SHALE OIL**  
**COKE YIELD**

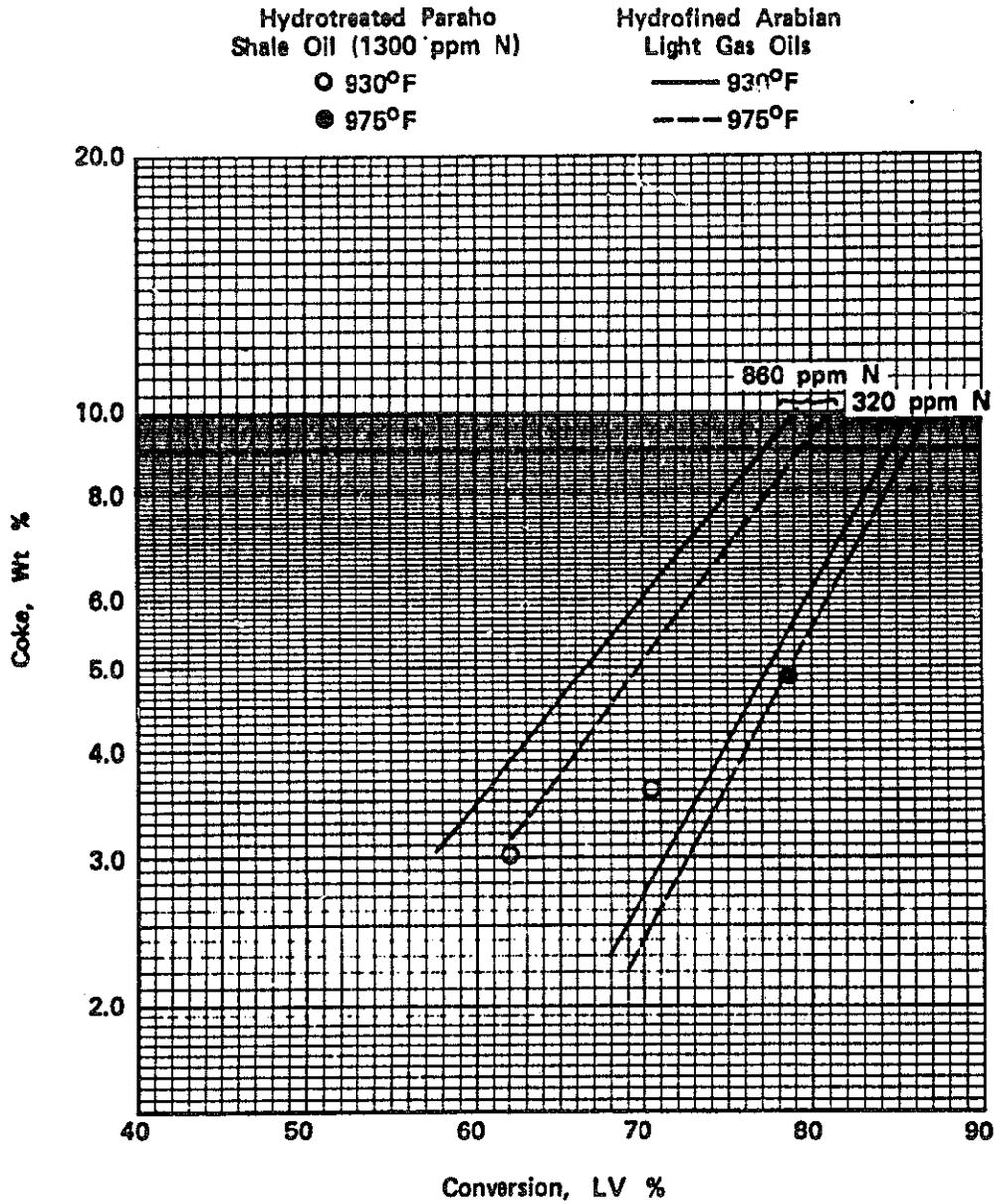
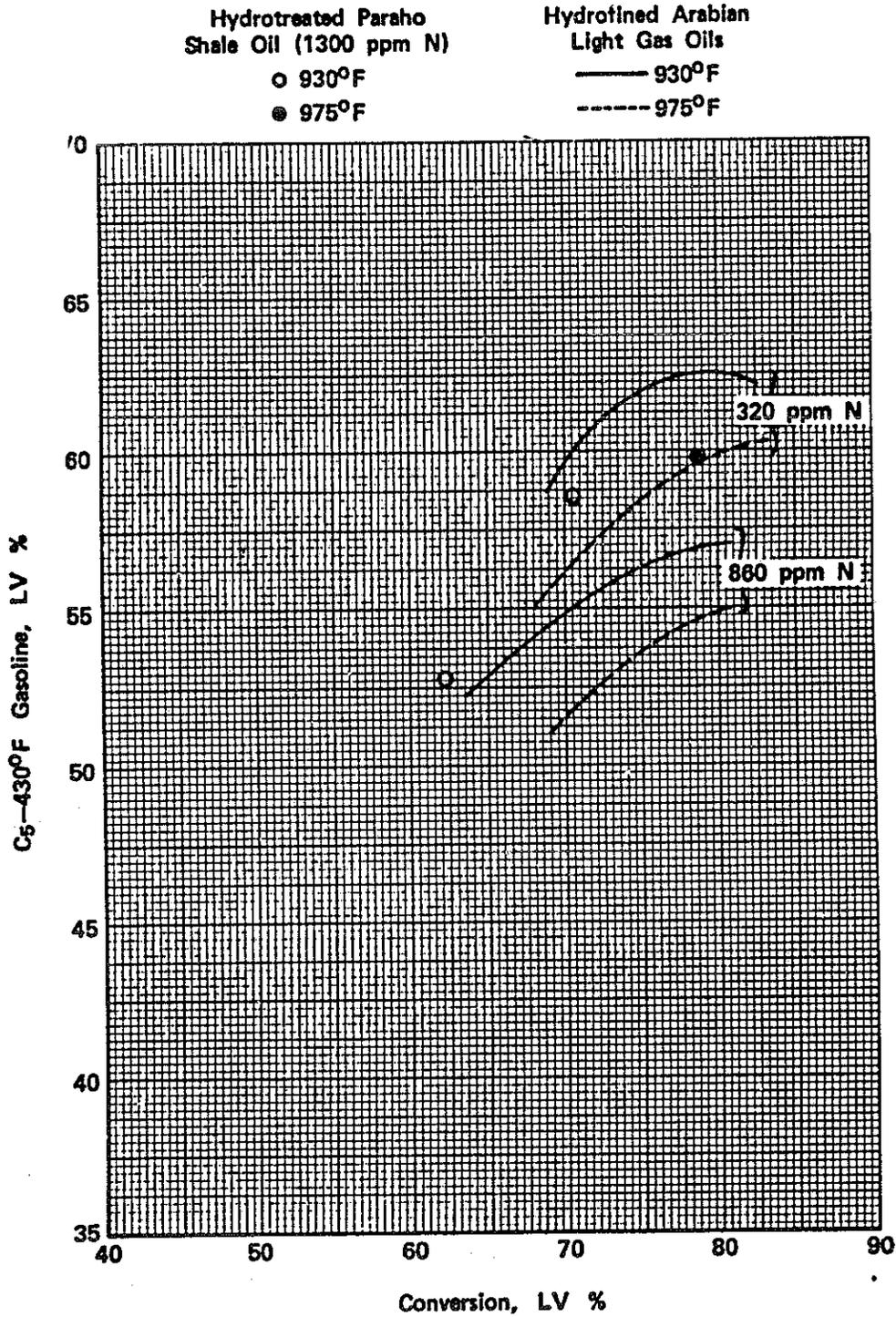


FIGURE 33

DOE CONTRACT EF-76-C-01-2315  
CATALYTIC CRACKING OF  
HYDROTREATED PARAHO SHALE OIL  
C<sub>5</sub>-430°F GASOLINE YIELD



**FIGURE 34**  
**DOE CONTRACT EF-76-C-01-2315**  
**CATALYTIC CRACKING OF**  
**HYDROTREATED PARAHO SHALE OIL**  
**C<sub>3</sub> YIELDS**

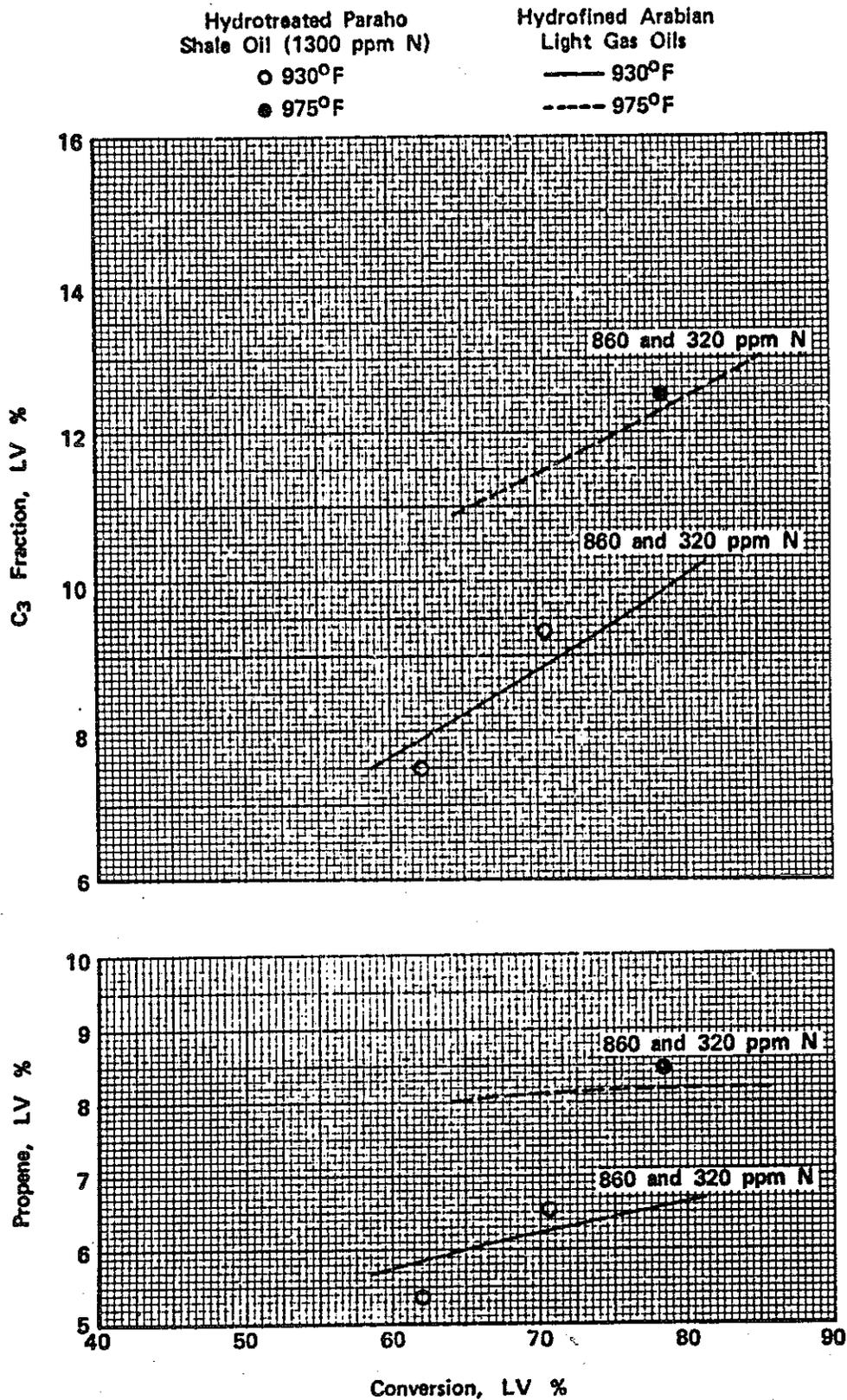


FIGURE 35

DOE CONTRACT EF-76-C-01-2315  
 CATALYTIC CRACKING OF  
 HYDROTREATED PARAHO SHALE OIL  
 C<sub>4</sub> YIELDS

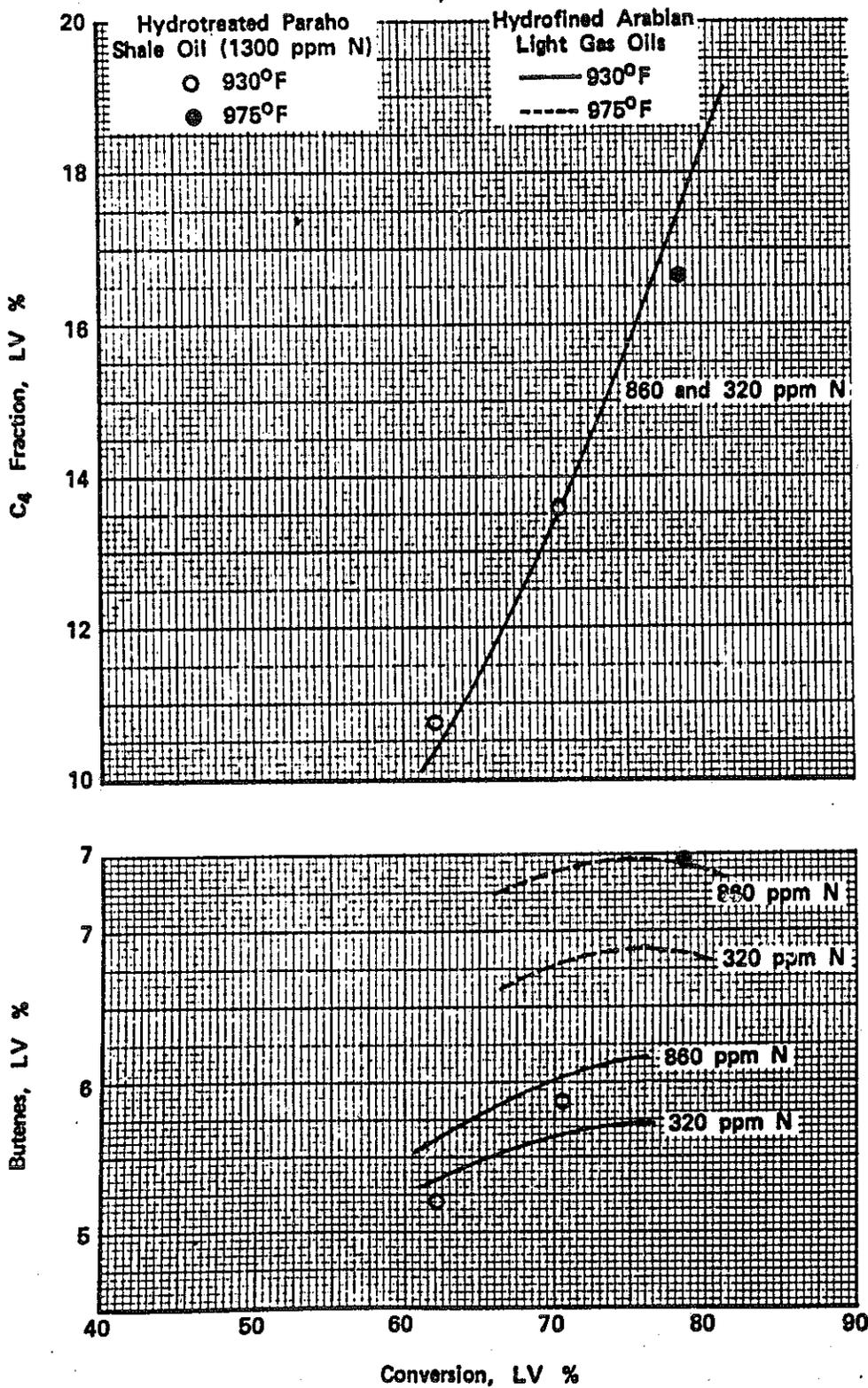


FIGURE 38

DOE CONTRACT EF-76-C-01-2315  
 CATALYTIC CRACKING OF  
 HYDROTREATED PARAHO SHALE OIL  
 F-1 CLEAR OCTANE NUMBER

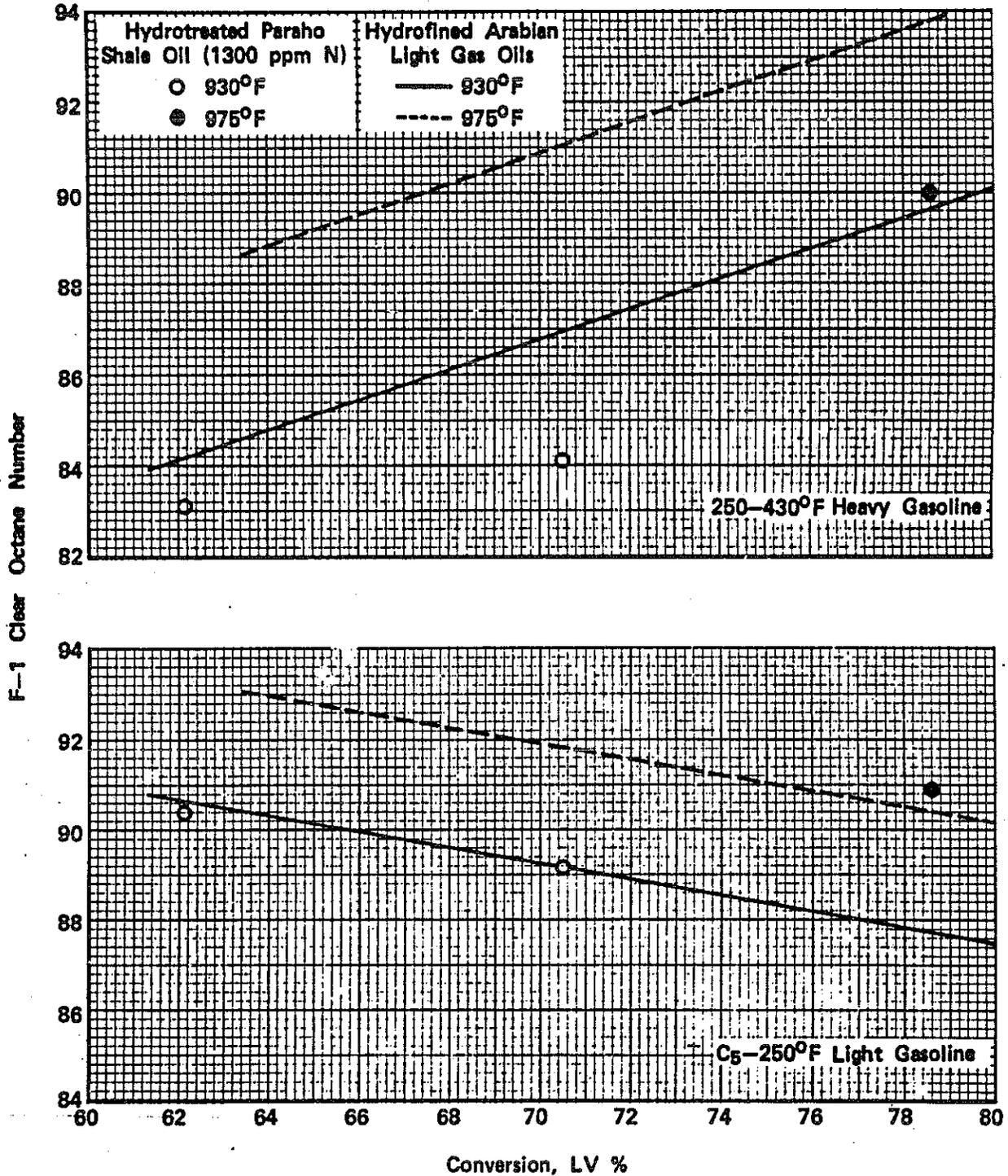


FIGURE 37

DOE CONTRACT EF-76-C-01-2315  
CATALYTIC CRACKING OF  
HYDROTREATED PARAHO SHALE OIL  
F-2 CLEAR OCTANE NUMBER

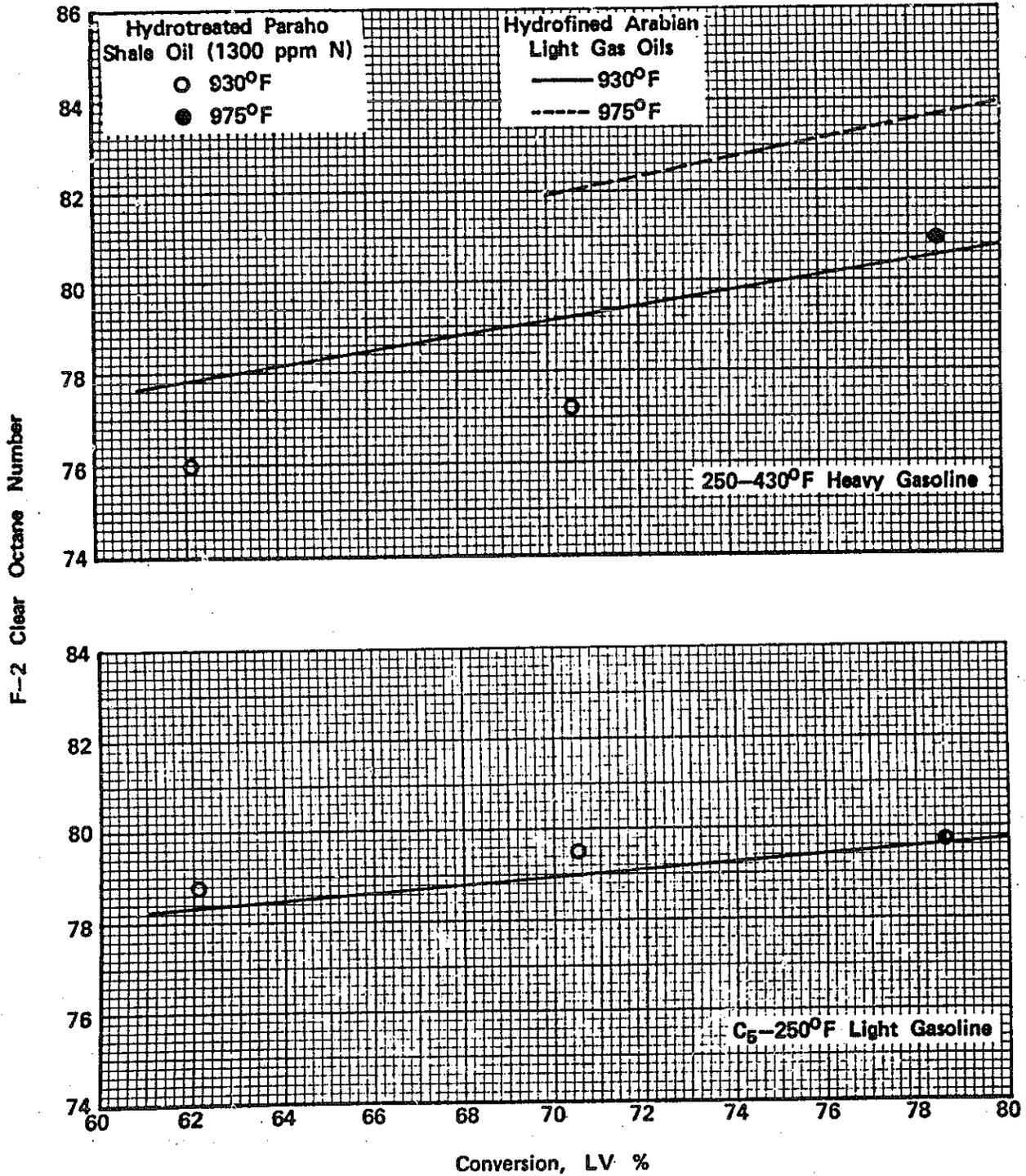


FIGURE 38

DOE CONTRACT EF-76-C-01-2315  
 HYDROTREATING OF TWO SHALE OIL DISTILLATES

0.5 LHSV, 1900 PSIG, 5000 SCF/BBL, ONCE-THROUGH H<sub>2</sub>  
 ICR 108

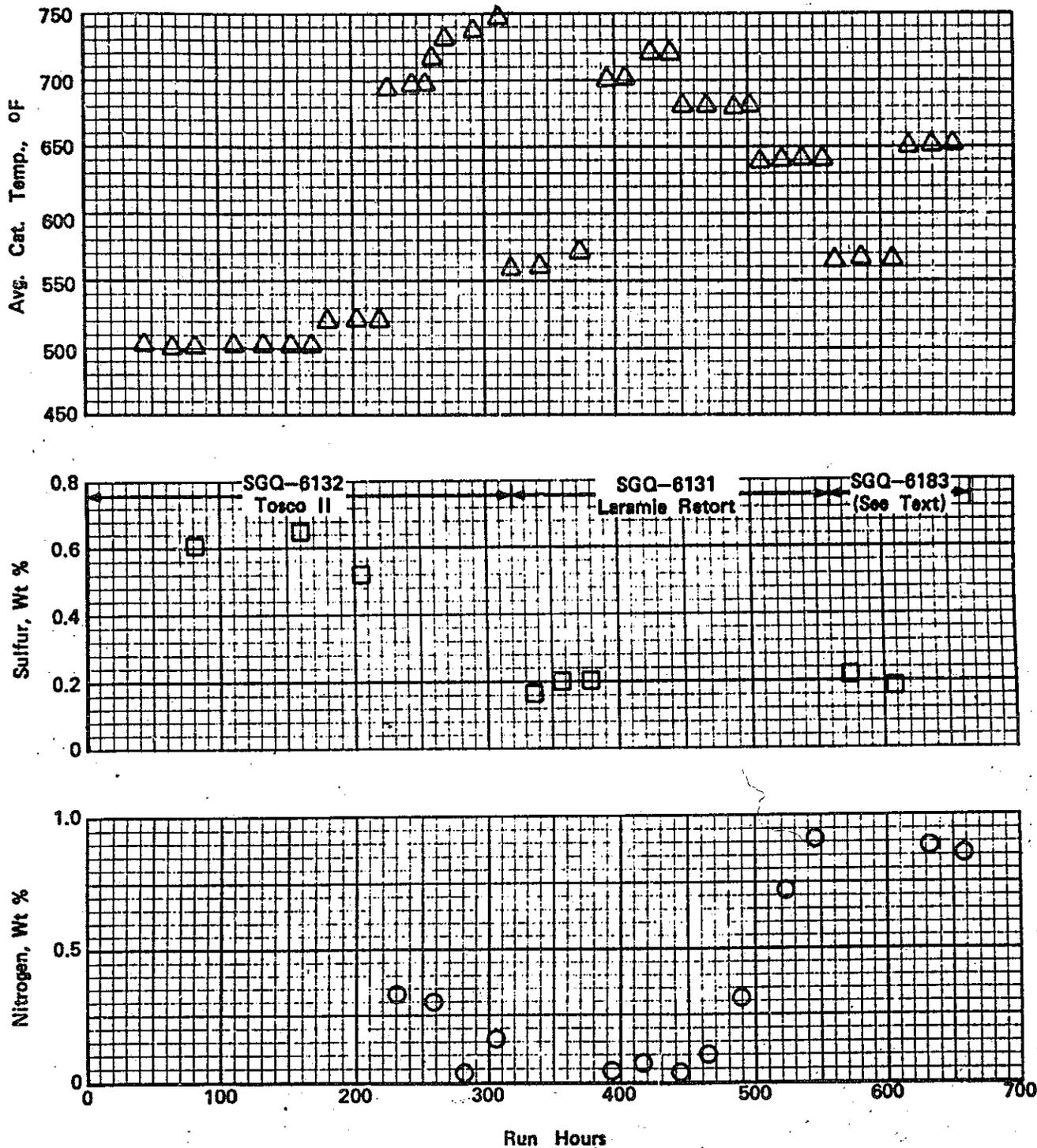
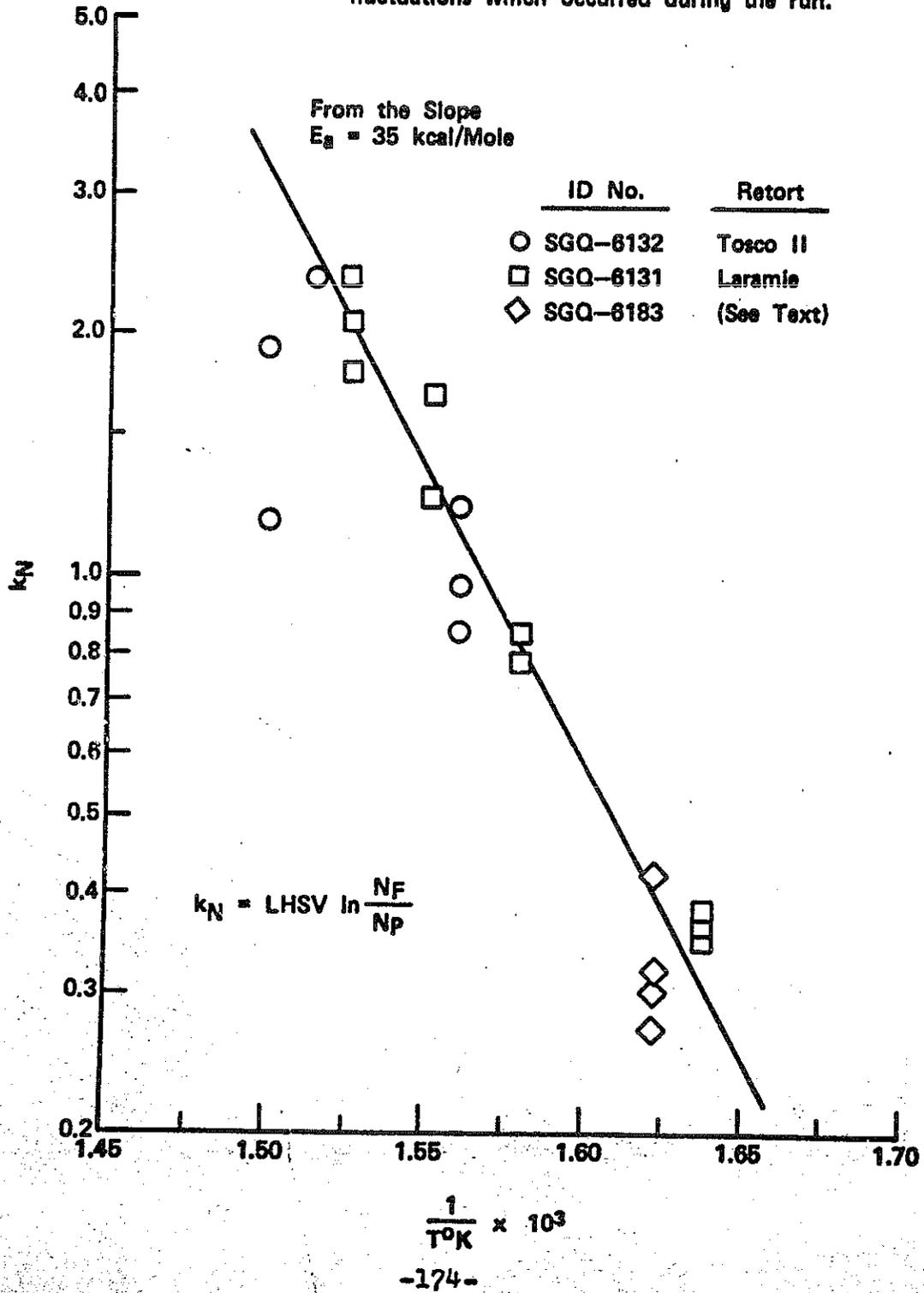


FIGURE 39

DOE CONTRACT EF-78-C-01-2315  
 HYDROTREATING OF TWO SHALE OIL DISTILLATES

DETERMINATION OF EACT FOR HDN  
 0.5 LHSV, 1900 PSIG, 5000 SCF/BBL, ONCE-THROUGH H<sub>2</sub>  
 ICR 106

Note: The scatter is partly due to pressure fluctuations which occurred during the run.



Part II

Engineering Design Studies and  
Estimated Processing Costs

By

H. A. Frumkin

DOE Contract EF-76-C-01-2315

## I. Introduction

The refining cost of a crude oil can vary widely, depending on many factors, such as crude oil properties, refinery capacity, type of processing, desired products, and refinery location, among others. The following cases were evaluated to determine the estimated range of costs for refining crude Paraho shale oil in conventional petroleum processing plants:

Case	Processing	Products	Location
1	Hydrotreating-Hydrocracking	50,000 BPCD Light Products [Liquefied Petroleum Gas (LPG), Motor Gasoline, Jet Fuel, and Diesel]	Remote Rocky Mountain Area
2	Hydrotreating-Fluid Catalytic Cracking (FCC)	Same as Case 1	Same as Case 1
3	Delayed Coking-Hydrotreating	Same as Case 1	Same as Case 1
4	Hydrotreating-Hydrocracking	100,000 BPCD Light Products	Near an Urban Center in the Mid-Continent or Rocky Mountain Area
5	Hydrotreating-FCC	Same as Case 4	Same as Case 4
6	Delayed Coking-Hydrotreating	Same as Case 4	Same as Case 4
7	Hydrotreating	100,000 BPCD Hydro-treated Synthetic Crude	At Shale Retorts in Rocky Mountain Area
8	Hydrotreating	Same as Case 7	Existing Refinery Near an Urban Center in Mid-Continent or Rocky Mountain Area

The refinery plans are based on the use of advanced, state-of-the-art, commercially proven technology. At the heart of each processing plan are Chevron-licensed processes well suited to refining this feedstock. The hydrotreating and hydrocracking plants are based on the Isocracking

process, catalytic reforming on the Rheniforming process and waste water processing for hydrogen sulfide and ammonia recovery on the Chevron WWT process. Process conditions and yields for the various refinery units are based directly on the laboratory and pilot plant data obtained for this project and contained in Part I of this report (FE-2315-25A), as well as on general petroleum processing correlations.

The primary basis for the first six cases is to produce a fixed total amount of finished transportation fuels (LPG, motor gasoline, kerosene jet fuel, and/or diesel fuel) in a "grass roots" refinery, that is, a complete, new refinery which includes all necessary supporting facilities such as utility plants, tankage, and required environmental control equipment. Two product rates have been selected for cost comparison: 50,000 and 100,000 BPCD of finished transportation fuels. The 50,000-BPCD rate would be suitable for a remote Rocky Mountain refining site and could be associated with the shale retorting facility. This is also the product rate at which the major processing plants have components near maximum size. Increasing production above this rate requires some duplication of critical pieces of equipment. The 100,000-BPCD refinery cases would fit into the larger marketing area typical of a Mid-Continent urban location. Costs for these latter cases also would be applicable to a Rocky Mountain urban location.

Cases 7 and 8 are based on producing an upgraded synthetic crude from raw shale oil. The synthetic crude should be suitable for processing in a large percentage of existing refineries. Here costs are compared for building the necessary hydrotreating facilities simultaneously and in conjunction with the shale retorting project (Case 7) versus adding the hydrotreating facilities ahead of an existing refinery (Case 8).

## II. Processing Plans

Figures 1, 2, and 3 show "grass roots" refining plans for the 50,000-BPCD product rate in the form of block flow diagrams. These diagrams are meant to point up differences in the overall refinery arrangements and, hence, do not show details within the process plants.

The first plan (Figure 1) uses a hydrotreater to remove nitrogen, sulfur, arsenic, and other contaminants from the whole shale oil, as well as to hydrogenate and partially hydrocrack it. The hydrofined product is distilled and stocks in the salable gas oil boiling range are further hydroprocessed to finished products. A recycle hydrocracker converts a major portion of the nonsalable heavy gas oil into light products.

The second plan (Figure 2) uses a similar hydrotreater and distillation for processing the whole shale oil but feeds the hydrofined nonsalable gas oil to an FCC rather than a recycle hydrocracker for conversion to light products.

In the third plan (Figure 3), the shale oil is first processed in a delayed coker. The whole coker distillate is then hydrotreated and

fractionated with suitable downstream processing to finished products. In this case, diesel hydrotreating is included to ensure product stability over a typical range of commercial operating conditions.

While the total quantity of light products remains fixed from case to case, the relative amounts of gasoline and middle distillate vary. This variation was allowed for two reasons: First, it was decided to base refining yields as closely as possible on the actual results from the laboratory runs; and, second, the product distributions shown represent typical product slates for each of the processing plans presented. Refinery units and conditions can be selected to give essentially the same gasoline to middle distillate ratios for each plan where a specific product slate is desired. Case 1 (the hydrocracking case) is by far the most flexible in allowing adjustment of the product mix. Also, Case 1 is the only one in which it is possible to efficiently maximize jet fuel production.

Each plan represents a completely feasible and reasonably efficient refining arrangement. A more detailed optimization of conditions, cut points, and other factors, while not within the scope of this study, would be carried out in the course of a normal refinery design.

Simplified flow diagrams for the hydrotreating, recycle hydrocracking, and WWT plants are shown in Figures 4, 5, and 6. The diagrams are provided for readers unfamiliar with these Chevron-licensed processes. Detailed feed properties, yields, and product qualities for the hydrocracker and hydrotreaters in each case are given in Tables I-V. This information is based directly on the laboratory results given in Part I of this report (FE-2315-25A) with some minor adjustments to represent average yields over an entire operating cycle. WWT plants are included in the processing schemes not only to allow recycling of most of the process water for use within the process plants but also to produce 200 to 400 tons/day of salable ammonia from the large quantity of nitrogen removed by hydrotreating shale oil.

III. 50,000-BPCD  
"Grass Roots" Refinery  
Cases 1, 2, and 3

Stock Balances

The following alternative processing schemes were evaluated:

- Case 1: Hydrotreating-Hydrocracking
- Case 2: Hydrotreating-Fluid Catalytic Cracking
- Case 3: Severe Coking Followed by Hydrotreating

These cases are based on producing 50,000 BPCD of transportation fuels in a new "grass roots" refinery located in the Rocky Mountain Area near the shale retorting facilities.

Detailed stock balances for each case are shown in Tables VI, VII, and VIII. Estimates of finished gasoline production were made by using correlations for catalytic reforming of naphtha and blending of gasoline components. Correlations could be used in this case because the properties of the reformer feeds distilled from the pilot plant hydro-treater products were well within Chevron's range of experience on petroleum stocks. Gasoline blending objectives were to make a single unleaded pool meeting accepted industry specifications for quality with minimum octane numbers of 93 Research (F-1 clear), 84 Motor (F-2 clear), and 89 (Research + Motor)/2. These octane numbers are typical of those projected for the early 1980's. Refining conditions were set to produce middle distillates meeting all present specifications. Table IX lists the refinery product inspections.

Another objective was to meet the refinery furnace fuel and hydrogen plant feed requirements with internally supplied clean fuels, ensuring minimum air emissions. Boiler plants, however, were assumed to be coal-fired in compliance with present DOE regulations. In Cases 1 and 2, all light gases produced were used for fuel and hydrogen plant feed; while in Case 3, the smaller hydrogen plant allowed a moderate amount of LPG to be recovered as product. If the refinery were located in or close to the shale fields, retort gas might be used as fuel, a possibility which would require further study.

Case 1, the hydrotreating-hydrocracking combination, uses 55,000 BPCD of shale oil feed to produce the target 50,000 BPCD of light products. The hydrotreating-FCC Case 2 requires a slightly lower feed rate of 52,900 BPCD; but the coking Case 3 needs a substantially higher 64,300 barrels of shale oil to produce the same 50,000 barrels of light products due to the high yield of coke. A milder coking-FCC case was also examined but proved to be significantly more costly without showing enough improvement in yield of light products to justify the higher expense.

#### Cost Estimates

Investments and utility requirements for the hydrotreater and hydro-crackers are given in Table X. These were estimated from cost correlations of actual plants constructed by Standard Oil Company of California. Hydrotreater reactors were sized to include the same fraction of guard catalyst as was used in the bench-scale catalyst life demonstration runs discussed in Part I of this report (FE-2315-25A). The guard bed catalyst removes arsenic and other catalyst poisons ahead of the normal hydrotreating catalyst. Reactor cost could be reduced if arsenic compounds were removed upstream of the hydrotreaters, a possible area for future study.

Plant investment costs for the overall refineries in Cases 1, 2, and 3 are broken down and detailed in Tables XII through XIV. Again, these estimates were developed using cost correlations based on actual plants constructed by Standard Oil Company of California. The important bases

for these estimates are summarized in Table XI. Investments are categorized as "onplot," those directly concerned with the individual refinery process plants, and "offplot," for auxiliary or supporting facilities such as utility plants, tankage, etc. The investment estimates are based on First Quarter 1978 costs. The estimating allowances shown on the tables are included to cover the cost of additional items which the history of major projects shows are encountered as the detailed project engineering proceeds.

The investments shown are estimated for a remote location which could be in the Rocky Mountain Area near the oil shale reserves. For these areas, additional costs are included to account for a cold climate with requirements for winterizing facilities, a construction camp, premium wage rates, and longer work weeks, etc. As mentioned above, the estimate for the boiler plant is based on coal-fired burners with attendant stack gas sulfur dioxide (SO<sub>2</sub>) removal facilities. As indicated on Table XI, no allowance is made for (a) shale resource costs; (b) shale mining, handling, retorting, or disposal; (c) shale oil transportation to; or (d) refined product distribution and transportation from the refinery. These additional costs are not required to evaluate refinery processing costs. However, they should be included if it is desired to determine the overall economics of a specific shale oil refining project.

In Cases 1-3 and all subsequent cases, the "waste disposal" item includes API separators and flotation followed by biological and activated carbon water treatment. In Case 3, the estimate is higher because of the additional anticipated need to remove arsenic from refinery effluent water. This is not required in Cases 1 and 2 because, as shown in the experimental work, arsenic is quantitatively removed by the hydrotreater guard catalyst before contact with water.

#### Cost Comparison

Table XV summarizes the cost estimates for Cases 1, 2, and 3 and compares the cases in a standard manner. Factors used for labor, maintenance, taxes, and insurance are typical of those used for analyzing long-term, large-scale commercial projects. The capital charge factor, the yearly rate at which the investment is charged to the project, was chosen to provide about a 15% after-tax discounted cash flow (DCF) rate of return on investment based on reasonable and commonly used assumptions for projects of this type and magnitude. These assumptions are summarized on Table XI. The sensitivity of the costs to varying rates of return in the range of 10% to 20% is discussed in a later section of this report.

Table XV shows that a "grass roots" shale oil refinery making 50,000 BPCD of distillate fuels will cost about \$500 million dollars to build now (First Quarter 1978). The lowest cost plan is Case 3 (coking-hydrotreating) followed by Case 2 (hydrotreating-FCC) and Case 1 (hydrotreating-hydrocracking). The operating costs and processing costs per barrel of product also follow the same sequence. However, the process plan selection must take into account other factors as well.

Of major importance is the significantly lower efficiency of the coking refinery (Case 3) in converting shale oil to light products. About 17-22% more shale oil is required in Case 3 to make the same volume of light products as the other cases. In essence, this additional feed ends up as coke, a product of lower value than light products. Even if the product coke were substituted for coal as the boiler plant fuel in the coking cases, cost savings would only amount to about 20 cents/bbl.

With the incentive to maximize light products, the choice may be between the more efficient hydrocracking and FCC processing schemes. The FCC case has about a 9% lower overall processing cost and a 4% higher conversion efficiency (barrels of light product per barrel of feed). On the other hand, the hydrocracking case can make more and better jet fuel and has the potential for greater flexibility in varying jet fuel and gasoline/diesel ratio. The optimum processing selection will depend on the product demands and product values specified for a given application.

#### IV. 100,000-BPCD "Grass Roots" Refinery Cases

To show the effect of refinery size and location, costs were also estimated for refineries with double the capacity of the first three cases. These larger refineries are assumed to be located in an urban area where the potential market for 100,000 BPCD of additional light products would exist. Investment costs are essentially equivalent for refineries located near any large Mid-Continent or Rocky Mountain Area city. These urban refinery investments are about 10% lower than those for an equivalent refinery in a remote location. For reference purposes, Gulf Coast refinery investments would be about 5% lower than those for the urban Mid-Continent or Rocky Mountain Areas.

We have not shown flow diagrams or stock balances for these larger refinery cases, but stock balance summaries are included on Table XXV. Processing arrangements for Cases 4, 5, and 6 would be identical to Cases 1, 2, and 3, respectively, with all rates doubled. Tables XVI through XVIII show detailed cost breakdowns, while Table XIX summarizes the cost comparison. In estimating the costs of these larger refineries, each situation was carefully examined to determine if capacity could be doubled by size increase alone or if key equipment had to be duplicated resulting in higher construction costs. All the other bases mentioned above again apply. The combination of area and capacity change causes a reduction of about \$2 in total per barrel processing costs for each refinery scheme relative to the appropriate 50,000-BPCD case.

#### V. 100,000-BPCD Synthetic Crude Cases

Another possible way to refine shale oil would be to produce an upgraded synthetic crude oil suitable as feed to a typical existing refinery. The upgrading would consist mainly of arsenic, nitrogen, and sulfur removal by hydrotreating, plus an increase in gravity due to partial

hydrocracking. The advantages of this kind of processing are discussed in a recent publication ("Catalytic Hydroprocessing of Shale Oil to Produce Distillate Fuels," by R. F. Sullivan and B. E. Stangeland, presented to the Petroleum Division of the American Chemical Society, August 1977). Briefly, it produces a synthetic crude which has less sulfur than most crude oils and essentially no vacuum residuum. As discussed in the publication, the atmospheric residuum from this synthetic crude is a particularly desirable catalytic cracking feed.

Table XX is a stock balance for this type of refining module. Figure 7 shows a simplified flow diagram. The onplot equipment consists of the same type of first-stage hydrotreater as used in the previous Case 5, a separate hydrogen plant, and appropriate sulfur recovery and effluent treating facilities. The properties of the upgraded synthetic crude are shown in Table XXI.

Costs were developed for two situations: Case 7 represents an upgrading module built in the shale area at the same time as the shale retorts and is integrated with them to an appropriate degree. Case 8 is the same basic module built near an existing urban refinery sharing some of its peripheral facilities. If sufficient sulfur recovery capacity already exists in a given refinery, the Case 8 processing costs would be reduced by about 10 cents per barrel since no new sulfur recovery facilities would be required. Tables XXII and XXIII show the investment cost and utilities breakdowns for these cases. Aside from the remote versus urban cost factor, the major difference between the cases is the boiler plant cost. This difference occurs because a new boiler plant would have to be constructed for the module at an existing refinery, and this is a more expensive alternative than adding incremental capacity to the new boiler plant being constructed for the shale retorts. The area and boiler plant factors tend to cancel one another; and the other differences are small enough that the overall costs for Cases 7 and 8 are equal, as shown on Table XXIV. No costs were included for revamping of downstream refinery facilities to accept the different synthetic crude. These could only be estimated if a specific situation were defined.

#### VI. Sensitivity of Processing Costs to Rate of Return on Investment

Shale oil processing facilities are extremely capital intensive. Therefore, the DCF rate of return on investment and corresponding capital charge factor strongly affect the processing costs. Tables XXVII and XXVIII and Figures 8, 9, and 10 show how the processing costs for all cases change as the rate of return changes over the range from 10% to 20%. Case 4, for example, the 100,000-BPCD hydrotreating-hydrocracking case, has a processing cost of only \$7.30 per barrel with a 10% rate of return; but it increases significantly to \$12.65 per barrel at 20% rate of return. The total processing costs shown in Tables XV, XIX, XXIV, and XXVI include capital charges based on a 15% DCF rate of return on

investment. In evaluating new projects, a 15% DCF rate of return is often used in industry as a checkpoint below which the justification for investment becomes increasingly questionable. Taking into account that production of oil from shale is an unproven venture with significant internal and external uncertainties and risks, the appropriate rate of return to be used in economic analysis of this new technology should be at least this high.

## VII. Overall Cost Summary

Stock balances and costs for all eight cases are summarized on Tables XXV and XXVI, respectively. Major conclusions based on the overall results of this study are:

A. A "grass roots" refinery in a remote Rocky Mountain location producing 50,000 BPCD of finished transportation fuels (LPG, motor gasoline, jet fuel, and diesel fuel) from Paraho raw shale oil will cost about \$500 million to build now and will result in a total processing cost of between \$10 and \$12 per barrel of light products, depending on the processing route selected.

B. The hydrotreating-FCC and hydrotreating-hydrocracking processing alternatives are most efficient for refining Paraho shale oil to produce distillate transportation fuels. A refinery based on coking requires about 17-22% more shale oil feed to produce the same volume of distillate products plus the lower valued coke.

C. The hydrotreating-FCC refinery shows a somewhat lower processing cost and slightly higher production of light products from feed than the hydrocracking alternative. However, the hydrotreating-hydrocracking refinery has greater flexibility for varying individual product yields and is the only scheme with the capability to efficiently maximize jet fuel production.

D. Doubling the capacity (100,000 BPCD of light products) and locating the "grass roots" shale oil refinery near an urban center (Mid-Continent or Rocky Mountain Area) decreases the total processing cost to about \$8-\$10 per barrel and increases the required total investment to between about \$800 million and \$900 million.

E. Hydrotreating to produce 100,000 BPCD of an upgraded synthetic crude, rather than finished products, will result in a processing cost of about \$6.50 per barrel and a total investment of about \$600 million.

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ENGINEERING DESIGN STUDIES AND  
ESTIMATED PROCESSING COSTS  
REFINING OF PARAHO SHALE OIL -  
DOE CONTRACT EF-76-C-01-2315

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ENGINEERING DESIGN STUDIES AND  
ESTIMATED PROCESSING COSTS  
REFINING OF PARAHO SHALE OIL -  
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**TABLE I**  
**YIELDS AND PRODUCT INSPECTIONS**  
**FIRST-STAGE HYDROTREATING - CASE 1**  
**REFINING OF PARAHO SHALE OIL**  
**DOE CONTRACT EF-76-C-01-2315**

Feedstock - Dewatered Whole Paraho Shale Oil

Process Yields				
	Weight Percent of Input Hydrocarbon	Raw Feed Basis		
		Pounds per 100 Barrels	Barrels per 100 Barrels	Chemical Consumption, BCF/Barrel
				2000
H <sub>2</sub>	0.67	225		
H <sub>2</sub> S	1.25	408		
H <sub>2</sub> O	2.57	838		
NH <sub>3</sub>	0.60	196		
C <sub>1</sub>	0.75	245		
C <sub>2</sub>	0.80	261		
iC <sub>4</sub>	0.20	65		
nC <sub>4</sub>	0.50	163		
C <sub>3</sub> /C <sub>6</sub>	1.8	587	2.6	
180-350°F	10.0	3,262	12.3	
350-650°F	48.1	15,690	53.8	
650°F <sup>+</sup>	36.0	11,743	38.3	
<b>Total</b>	<b>103.26</b>	<b>33,683</b>	<b>107.0</b>	
650-850°F	27.0	8,809	28.8	
850°F <sup>+</sup>	9.0	2,936	9.5	

Inspections							
Stock	Raw Feed	Products, Boiling Range					
		C <sub>3</sub> /C <sub>6</sub>	180-350°F	350-650°F	650°F <sup>+</sup>	650-850°F	850°F <sup>+</sup>
Gravity, °API	20.2	83.9	55.4	38.2	30.0	30.3	29.1
Aniline Point, °F			130	150	210	195	-
<u>ASTM Distillation, °F</u>	D 1160		D 86	D 86	D 1160	D 1160	
St/5	400/480		210/-	400/420	650/720	620/660	
10/30	530/670		230/245	425/450	730/760	670/695	
50	775		260	500	810	725	
70/90	870/1000		280/310	530/580	850/930	765/805	
95/EP	-/-		-/350	600/630	1000/1060	830/875	
<u>Composition, LV %</u>							
Paraffins			50				
Naphthenes			40				
Olefins			-				
Aromatics			10				
<b>Total</b>			<b>100</b>				
Sulfur, ppm	6,600		5	40	80	80	
Total Nitrogen, ppm	21,800		60	450	900	550	
Total Oxygen, ppm	11,600		100	100	150	100	
Pour Point, °F				0	+105	+85	
Ramsbottom Carbon, %					0.15	0.05	
Viscosity at 122°F, cSt					19	10	
Viscosity at 210°F, cSt					5	3.4	
<u>Octane Numbers</u>							
F-1 Clear		72	45				
F-2 Clear		73					

TABLE II

YIELDS AND PRODUCT INSPECTIONS  
 RECYCLE HYDROCRACKING - CASE 1  
 REFINING OF PARAHO SHALE OIL  
 DOE CONTRACT EF-76-C-01-2315

Feedstock - 650-850°F Product from  
 Hydrotreating of Paraho Shale Oil

Process Yields				
	Weight Percent of Input Hydrocarbon	Raw Feed Basis		
		Pounds per 100 Barrels	Barrels per 100 Barrels	Chemicals Consumption SCF/Barrel
H <sub>2</sub>				1250
H <sub>2</sub> S	0.01	3		
NH <sub>3</sub>	0.07	21		
C <sub>1</sub>	0.30	92		
C <sub>2</sub>	0.45	138		
C <sub>3</sub>	2.0	612		
iC <sub>4</sub>	2.5	765		
nC <sub>5</sub>	1.8	550		
C <sub>5</sub> /C <sub>6</sub>	10.0	3,059	13.3	
180-300°F	22.5	6,880	26.7	
300-535°F	62.54	19,125	69.0	
Total	102.17	31,244	109.0	

Inspections				
Stock	Feed (Hydro- treated Product)	Products, Boiling Range		
		C <sub>5</sub> /C <sub>6</sub>	180-300°F	300-535°F
Gravity, °API	30.3	84.1	60.3	47.1
Aniline Point, °F	195		140	165
<u>ASTM Distillation, °F</u>	D 1160		D 86	D 86
St/5	620/660		205/-	330/355
10/30	670/695		225/240	360/385
50	725		250	415
70/90	765/805		265/290	445/480
95/EP	830/875		-/310	495/515
<u>Composition, LV %</u>				
Paraffins			60	55
Naphthenes			35	40
Olefins			-	-
Aromatics			5	5
Total			100	100
Sulfur, ppm	80			
Total Nitrogen, ppm	550			
Total Oxygen, ppm	100			
Pour Point, °F				<-85
Smoke Point, Millimeters				30
Freeze Point, °F				-60
Viscosity at 122°F, cSt	10			
Viscosity at 210°F, cSt	3.4			
Ramsbottom Carbon, %	0.05			
<u>Octane Number</u>				
F-1 Clear		74	50	
F-2 Clear		75		

TABLE III  
 YIELDS AND PRODUCT INSPECTIONS  
 FIRST-STAGE HYDROTREATING - CASE 2  
 REFINING OF PARAHO SHALE OIL  
 DOE CONTRACT EP-76-C-01-2315

Feedstock - Dewatered Whole Paraho Shale Oil

Process Yields				
	Weight Percent of Input Hydrocarbon	Raw Feed Basis		
		Pounds per 100 Barrels	Barrels per 100 Barrels	Chemical Consumption, SCF/Barrel
H <sub>2</sub>				1900
H <sub>2</sub> S	0.69	225		
H <sub>2</sub> O	1.25	408		
NH <sub>3</sub>	2.54	828		
C <sub>1</sub>	0.55	179		
C <sub>2</sub>	0.70	228		
C <sub>3</sub>	0.75	245		
iC <sub>4</sub>	0.20	65		
nC <sub>4</sub>	0.46	150		
C <sub>5</sub> /C <sub>6</sub>	1.8	587	2.6	
180-350°F	10.0	3,262	12.3	
350-650°F	47.4	15,462	53.0	
650°F <sup>+</sup>	36.75	11,988	38.9	
<b>Total</b>	<b>103.09</b>	<b>33,627</b>	<b>106.8</b>	

Inspections					
Stock	Raw Feed	Products, Boiling Range			
		C <sub>5</sub> /C <sub>6</sub>	180-350°F	350-650°F	650°F <sup>+</sup>
Gravity, °API	20.2	83.9	55.2	38.0	29.0
Aniline Point, °F			130	150	210
<u>ASTM Distillation, °F</u>	D 1160		D 86	D 86	D 1160
St/5	400/480		210/-	400/420	650/720
10/30	530/670		230/245	425/450	730/760
50	750		260	500	810
70/90	870/1000		280/310	530/580	850/930
95/EP	-/-		-/350	600/630	1000/1060
<u>Composition, LV %</u>					
Paraffins			50		
Naphthenes			40		
Olefins			-		
Aromatics			10		
Total			100		
Sulfur, ppm	6,600		5	40	120
Total Nitrogen, ppm	21,800		70	500	1900
Total Oxygen, ppm	11,600		100	150	200
Pour Point, °F				0	+105
Ramsbottom Carbon, %					0.17
Viscosity at 122°F, cSt					22
Viscosity at 210°F, cSt					6
<u>Octane Numbers</u>					
F-1 Clear		73	45		
F-2 Clear		73			

TABLE IV  
 YIELDS AND PRODUCT INSPECTIONS  
 HYDROTREATING - CASE 3  
 REFINING OF PARAHO SHALE OIL  
 DOE CONTRACT EF-76-C-01-2315

Feedstock - Distillate from Severe Coking of  
 Dewatered Whole Paraho Shale Oil

Process Yields				
	Weight Percent of Input Hydrocarbon	Raw Feed Basis		
		Pounds per 100 Barrels	Barrels per 100 Barrels	Chemicals Consumption SCF/Barrel
H <sub>2</sub>				1250
H <sub>2</sub> S	0.64	193		
H <sub>2</sub> O	1.01	305		
NH <sub>3</sub>	2.10	634		
C <sub>1</sub>	0.35	106		
C <sub>2</sub>	0.75	226		
C <sub>3</sub>	0.50	151		
iC <sub>4</sub>	0.10	30		
nC <sub>5</sub>	0.30	91		
C <sub>5</sub> /C <sub>6</sub>	1.9	573	2.5	
180-350°F	17.15	5,174	19.7	
350-650°F	69.4	20,938	72.7	
650°F <sup>+</sup>	8.0	2,414	8.3	
<b>Total</b>	<b>102.20</b>	<b>30,835</b>	<b>103.2</b>	

Inspections					
Stock	Feed (Coker Distillate)	Products, Boiling Range			
		C <sub>5</sub> /C <sub>6</sub>	180-350°F	350-650°F	650°F <sup>+</sup>
Gravity, °API	32.5	84	57	40.2	38.5
Aniline Point, °F	105		130	160	175
ASTM Distillation, °F	D 86/1160		D 86	D 86	D 86
St/5	185/295		210/-	400/420	640/650
10/30	345/465		230/240	425/450	660/670
50	540		255	500	675
70/90	595/655		275/310	530/580	680/685
95/EP	-/685		-/350	600/630	-/700
<u>Composition, LV %</u>					
Paraffins			60		
Naphthenes			30		
Olefins			-		
Aromatics			10		
<b>Total</b>			<b>100</b>		
Sulfur, ppm	6,300		50	1	<10
Total Nitrogen, ppm	17,800		100	600	620
Total Oxygen, ppm	9,500		100	150	200
Pour Point, °F				0	+30
<u>Octane Numbers</u>					
F-1 Clear		70	40		
F-2 Clear		71			

TABLE V

PROPERTIES OF DEWATERED PARAHO SHALE OIL  
DOE CONTRACT EF-76-C-01-2315

Gravity, °API	20.2
Sulfur, Weight %	0.66
Total Nitrogen, Wt %	2.18
Oxygen, Weight %	1.16
Arsenic, Parts per Million	28
Pour Point, ASTM, °F	90
Carbon, Weight	84.30
Hydrogen, Weight %	11.29
Hydrogen/Carbon Atom Ratio	1.60
Chloride, Parts per Million	<0.20
Sodium, Parts per Million	5
Ash, Weight % (ASTM D 486)	0.03
Iron, Parts per Million	70
<u>Filter Residue Ash</u> <u>(0.8 µ-Filter)</u>	
Total Solids, Parts per Million	234
Ash, Parts per Million	168
<u>Filter Residue Ash</u> <u>(0.45 µ-Filter)</u>	
Total Solids, Parts per Million	252
Ash, Parts per Million	194
Sediment (Plus Trace Water), Volume %	0.1
Bromine Number	51
Average Molecular Weight	326
Ramsbottom Carbon, %	2.5
<u>Viscosity, Centistokes</u>	
At 122°F	25.45
At 210°F	5.541
Acid Neutralization No., milligrams KOH/gram	2.3
Base Neutralization No., milligrams KOH/gram (Equivalent)	38
pH	9.2
Maleic Anhydride No., milligrams per gram	40.6
Navy Heater Test	No. 1 Rating
Hot Heptane Asphaltenes (Including Any Fines), Wt %	0.17
<u>ASTM D 1160 Distillation, °F</u> <u>(Corrected to 1 Atmosphere)</u>	
	(Run at 10 millimeters)
St/5	386/456
10/30	508/659
50	776
70/90	871/995
EP	/1022
% Overhead (Excluding Trap)	94
% in Trap	1
% in Flash	5

TABLE VI

STOCK BALANCE - 50,000 BARRELS PER CALENDAR DAY LIGHT PRODUCTS\*  
 CASE 1 - "GRASS ROOTS" REFINERY (HYDROTREATING - HYDROCRACKING)  
 REFINING OF PARANO SHALE OIL  
 DOE CONTRACT EP-76-C-01-2315

Feeds and Products, Barrels per Calendar Day	Processing							Products					
	Refinery Input	First-Stage Hydro- treating	Recycle Hydro- cracking	Middle Distillate Hydro- treating	Naphtha Hydro- treating and Reforming	Hydrogen Manufacture	H <sub>2</sub> S Recovery and Sulfur Plant	Offplot Boiler Plant	Refinery Fuel	Motor Gasoline	Kerosene Jet Fuel	Diesel Fuel	By-Products
Raw Shale Oil	55,000	(55,000)											
Coal, Equivalent Fuel Oil	1,380	1,710	880	345	565	(2,360)		(1,380)	1,140	620			
Fuel Gas, Equivalent Fuel Oil		340	100		180								
Butane													
C <sub>2</sub> /C <sub>3</sub> Light Gasoline		1,410	2,100		(4,220)	(1,140)				2,370			
180-330°F Heavy Gasoline		6,765	4,220		(3,620)	(3,505)							
180-350°F Heavy Gasoline													
300-535°F Kerosene		29,505	10,940	(29,505)						10,940			
350-650°F Diesel		15,840	(15,840)						5,350				
650-850°F Heavy Gas Oil		5,350											
850°F+ Bottoms													
Reformate (93.5 F-1 Clear)					6,185					6,185		29,885	
Hydrotreated Diesel				29,885									
Total Liquid Product		60,920	18,240	30,230	6,930	7,005							
Total Liquid Feed	56,690	55,000	15,840	29,505	7,480							29,885	
Liquid Gain (Loss)		5,920	2,400	725	(550)	(7,005)							
Fuel		(890)	(200)	(185)	(470)	(4,715)	(30)						
Hydrogen, Millions of Standard Cubic Feet per Calendar Day		(125.4)/11.0	(22.5)	(5.9)	7.8	(11.0)/146.0							
Tons per Calendar Day													
Hydrogen Sulfide		61.9	0.2				(62.1)						55.5
Sulfur		230.4	1.7				55.5						232.1
Ammonia													

\*Light products include LPG, motor gasoline, jet fuel, and diesel fuel (No. 2 oil).  
 Note: Parentheses ( ) denote a negative quantity, i.e., consumption.

TABLE VII

STOCK BALANCE - 50,000 BARRELS PER CALENDAR DAY LIGHT PRODUCTS\*  
 CASE 2 - "GRASS ROOTS" REFINERY (HYDROTREATING - FLUID CATALYTIC CRACKING)  
 REFINING OF PARAGO SHALE OIL  
 DOE CONTRACT EF-76-C-01-2315

	Processing							Products				
	Refinery Input	First-Stage Hydro-treating	Fluid Catalytic Cracker	Middle Distillate Hydro-treating	Meththa Hydro-treating and Reforming	Hydrogen Manufacture	H <sub>2</sub> S Recovery and Sulfur Plant	Offplot Boiler Plant	Refinery Fuel	Motor Gasoline	Diesel Fuel	By-Products
<u>Feeds and Products, Barrels per Calendar Day</u>												
Raw Shale Oil	52,900	(52,900)										
Coal, Equivalent Fuel Oil	1,490	1,480	4,465	165	745	(4,315)		(1,490)	1,295			
Fuel Gas, Equivalent Fuel Oil		390	655		250							
Butane and/or Butene												
C <sub>3</sub> /C <sub>4</sub> Light Gasoline		1,355	7,855		(6,510)				9,210			
180-350°F Heavy Gasoline		6,510	4,440						4,440			
FCC Heavy Gasoline												
350-650°F Diesel		28,025	2,115	(28,025)								
FCC Light Cycle Oil			2,275	(1,880)								
FCC Heavy Cycle Oil			2,275									
650°F+ Bottoms		20,575	(20,575)									
Reformate (102 P-1 Clear)				30,140	4,915				4,915	30,140		
Hydrotreated Diesel												
Total Liquid Product		58,335	21,805	30,305	5,910	(4,315)			19,860	30,140		
Total Liquid Feed	54,390	(52,900)	(20,575)	(29,905)	(6,510)							
Liquid Gain (Loss)		5,435	1,230	400	1600	(4,315)						
Fuel		(845)	(225)	(185)	(410)	(3,355)	(30)					
Hydrogen, Millions of Cubic Feet per Calendar Day		(115.4)/10.6		(6.0)	7.8	(10.6)/113.6						
<u>Tons per Calendar Day</u>												
Hydrogen Sulfide		59.5	0.3				(59.8)					53.6
Sulfur							53.4					221.0
Ammonia		218.9	2.1									

\*Light products include LPG, motor gasoline, jet fuel, and diesel fuel (No. 2 Oil).

Note: Parentheses denote a negative quantity, i.e., consumption.

TABLE VIII

STOCK BALANCE - 50,000 BARRELS PER CALENDAR DAY LIGHT PRODUCTS\*  
 CASE 3 - "GRASS" REFINERY (SEVERE COKING - HYDROTREATING)  
 REFINING OF PARANO SHALE OIL  
 DOE CONTRACT EP-76-C-01-2315

Feeds and Products, Barrels per Calendar Day	Processing										Products			
	Refinery Input	Delayed Coking	Hydro- treating	Middle Distillate Hydro- treating	Naphtha Hydro- treating and Reforming	Hydrogen Manufacture	H <sub>2</sub> S Recovery and Sulfur Plant	Offplot Boiler Plant	Light Ends Recovery	Refinery Fuel	Liquidified Petroleum Gas (LPG)	Motor Gasoline	Diesel Fuel	By-Products
Raw Shale Oil	64,300	(64,300)						(1,790)						
Coal, Equivalent Fuel Oil	1,790													
Fuel Gas, Equivalent Fuel Oil		2,095	805	200	450	(2,935)			2,330	805	985			
Propane and/or Propylene		1,435	275		500					545				
Butane and/or Butylene		740	320		550						1,300			
C <sub>4</sub> /C <sub>5</sub> Light Gasoline (160-350)°F Heavy Gasoline			1,300		(10,245)									
C <sub>4</sub> + Whole Coker Distillate 350-650°P Diesel 650°P+ Bottoms		51,975	(51,975)	(37,760)	8,320				4,310					
Reformats (94 P-1 Clear) Hydrotreated Diesel												38,045		
Total Liquid Product	66,090	37,055	55,015	38,045	8,320	(2,935)				1,350	10,605	38,045		
Total Liquid Feed		(64,300)	(51,975)	(37,760)	(10,245)									
Liquid Gain (loss)		(7,245)	3,040	485	425	(2,935)								
Fuel		(2,570)	(875)	(235)	(645)	(2,285)	(30)							
Hydrogen, Millions of Standard Cubic per Calendar Day			(67.2)	(7.6)	9.6	65.1								
Tons per Calendar Day														
Coke		1,995											1995	
Hydrogen Sulfide		10.5	50.2	0.2									54.5	
Sulfur													54.5	
Ammonia		20.5	164.6										195.1	

\*Light products include LPG, motor gasoline, jet fuel, and diesel fuel (No. 2 Oil).

Note: Parentheses ( ) denote a negative quantity, i.e., consumption.

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TABLE IX

REFINERY PRODUCT INSPECTIONS  
REFINING OF PARAHO SHALE OIL  
DOE CONTRACT EF-76-C-01-2315

	Case		
	1	2	3
<u>Inspections of Motor Gasoline</u>			
Reid Vapor Pressure, Pounds	10	10	10
Docbor Test	Negative	Negative	Negative
Sulfur, Weight Percent	<0.03	<0.03	<0.03
F-1, Octane Number (Unleaded)	93.0 (Min.)	95.0	93.5 } Meets (F-1 + F-2) / 2
F-2, Octane Number (Unleaded)	85.0	84.0 (Min.)	84.5 } = 89 (Min.)
<u>Distillation Temperature, °F</u>			
At 50% Distilled	224	207	240
End Point, Maximum	420	420	420
<u>Inspections of No. 2 Oil (Diesel Fuel)</u>			
Gravity, °API	38.5	38.5	40.0
Flash Point, °F	>125 (Min.)	>125 (Min.)	>125 (Min.)
Pour Point, °F	0 (Max.)	0 (Max.)	0 (Max.)
Viscosity at 100°F, SUS	34.5	34.5	34
Sulfur, Weight Percent	<0.01	<0.01	<0.01
<u>Distillation Temperature, °F</u>			
At 50% Distilled	500	500	500
Cetane Index	46	46	45
<u>Inspections of Kerosene Jet Fuel</u>			
Gravity, °API	47.1	-	-
Flash Point, °F	>105 (Min.)	-	-
Freeze Point, °F	-60	-	-
Smoke Point, Millimeters	30	-	-
Sulfur, Weight Percent	<0.01	-	-
<u>Inspections of Coke</u>			
Sulfur, Weight Percent	-	-	0.65
Nitrogen, Weight Percent	-	-	4.1
Volatiles, Weight Percent	-	-	10

TABLE X

INVESTMENT AND UTILITY SUMMARY  
HYDROTREATERS AND RECYCLE HYDROCRACKER  
REFINING OF PARADO SHALE OIL  
DOE CONTRACT RP-76-C-01-2315

Case	1			2			3		
	Shale Oil First-Stage Hydrotreater	Distillation	Recycle Hydro- cracker	Distillation	Shale Oil First-Stage Hydrotreater	Distillation	Coker Distillate First-Stage Hydrotreater	Distillation	
Plant	61,000	67,000	18,000	38,000	59,000	65,000	58,000	60,000	
Feed Rate, Barrels per Operating Day									
Investment, Millions of Dollars <sup>1</sup>	67	15	34	11	65	15	40	17	
Total Onplot Initial Catalyst	9	0	5	0	9	0	5	0	
Utility Consumptions <sup>2</sup>									
Fuel, Barrels per Operating Day,	280	710	150	80	270	680	250	750	
Equivalent Fuel Oil <sup>3</sup>									
Power, Kilowatts <sup>4</sup>	3,800	0	600	0	3,500	0	1,000	0	
Compressors	3,300	600	1,300	300	3,800	600	2,600	500	
Steam, Thousands of Pounds per Hour <sup>5</sup>	0	(14)	0	0	0	(14)	0	(3)	
40 psig	73	(61)	11	(5)	69	(59)	57	(63)	
150 psig									
Water, Thousands of Gallons per Minute <sup>6</sup>	6	0	1	0	6	0	4	0	
Cooling Process	1	0.2	0	0	1	0.2	1	0.2	
Hydrogen Required, Millions of Standard Cubic Feet per Operating Day	122	-	22	-	112	-	74	-	
Leaks and Dissolved Bleed <sup>7</sup>	5	-	3	-	4	-	4	-	
Chemical Consumption	12	-	0	-	12	-	12	-	
Catalyst Replacement, Millions of Dollars per Year <sup>2</sup>	18	-	2	-	11	-	5	-	
Labor, Shift Positions	1	1	1	1	1	1	1	1	
Estimating Bases									
Catalyst Type	ICR 106	-	ICR 106	-	ICR 106	-	ICR 106	-	
Reactor Trains/ Reactors per Train	2/2	-	1/2	-	2/2	-	2/1	-	
Catalyst Life, Months To First Regeneration	6	-	12	-	9	-	12	-	
Total	6	-	30	-	9	-	12	-	

<sup>1</sup>Estimated for remote U.S. Rocky Mountain or Midcontinent location as of first quarter, 1978.

<sup>2</sup>Operating day basis.

<sup>3</sup>Supplied internally from refinery products.

<sup>4</sup>Gross consumptions; overall refinery requirements in later tables are net after power recovery turbines.

<sup>5</sup>Consumed (or produced) at pressure shown.

<sup>6</sup>Cooling water is recycled to cooling tower.

<sup>7</sup>Recycled through hydrogen plant.

TABLE XI  
BASES FOR COST ESTIMATES AND COMPARISONS  
REFINING OF PARAHO SHALE OIL  
DOE CONTRACT EF-76-C-01-2315

General

Estimated processing costs include annual capital charges and operating costs required to refine Paraho Shale Oil using conventional refining processes. No allowance is included in these estimates for (a) shale resource costs, (b) mining, handling, or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery. Notes apply to all subsequent cost tables. Item numbers are keyed to comparative cost summaries.

Investment Costs

Investment estimates are based on:

1. First Quarter, 1978 costs.
2. Mid-Continent or Rocky Mountain refining location.
  - a. Cases 1, 2, 3, and 7 refining facilities are assumed to be located in a remote Rocky Mountain site near the shale retorting facilities.
  - b. Cases 4, 5, 6, and 8 refining facilities are assumed to be located in either a Rocky Mountain or mid-Continent site (equivalent costs) near a large urban center.
3. Tankage equivalent to ten days of shale oil feed, ten days synthetic crude, fifteen days of motor gasoline blend components, fifteen days of light products, and ten days of intermediate products.
4. Adding an estimating allowance of about 10% of total onplot and offplot investment to account for additional cost items which the history of major projects shows are encountered as the detailed project engineering proceeds.

Working Capital Includes

5. Value of feed and product inventories in storage, assuming tanks are half full.
6. Estimated value of spare catalyst and spare parts.
7. Estimated value of accounts receivable less accounts payable for thirty days.

Capital Charge Factor

8. On Tables XV, XIX, XXIV, and XXVI, processing costs are based on a capital charge equal to 30% of total investment (including working capital and initial catalyst costs) per year. This capital charge factor is approximately equivalent to a 15% after tax discounted cash flow rate of return on investment based on:
  - a. 51% income tax.
  - b. Four-year construction period.
  - c. Investment payments equal to 10%, 15%, 25%, and 50% of total investment during the four years of construction.
  - d. 50% of design capacity during first year of operation; 100% in second year and thereafter.
  - e. 10% investment tax credit.
  - f. Sum of the years' digits depreciation; 13-year tax life of refining equipment.
  - g. 20-year project life.

Tables XXVII and XXVIII show the effect on processing costs of discounted cash flow rates of return varying from 10% to 20%. Figures 7, 8, and 9 illustrate this relationship graphically.

Utilities and Operating Costs

9. Refinery fuel and steam are internally supplied. Boiler plant fuel is purchased coal (amounts shown on stock balances).
10. Process water is conserved by treatment in the MWT plants and then returned to the refining units. ~1000 gallons per minute make-up water required. (>0.5 million dollars per year.)
11. Maintenance cost estimated at 2% of onplot plus offplot per year.
12. Taxes and insurance estimated at 2-1/2% of onplot plus offplot per year.
13. Operating labor cost estimated at 100,000 dollars per shift position per year. 65% is then added to cover support labor (technical service, administration, security, etc.) and supplies.
14. Power cost estimated at 3 cents per kilowatt-hour.
15. Coal cost estimated at 6 dollars per equivalent fuel oil barrel. (Net heating value of 6 million British thermal units.)

TABLE XII

ESTIMATES OF INVESTMENTS AND UTILITIES  
 CASE 1 - "GRASS ROOTS" REFINERY (HYDROTREATING - HYDROCRACKING)  
 REFINING OF PALMISTO SHALE OIL  
 DOE CONTRACT EP-78-C-01-2313

50,000 Barrels per Calendar Day Light Products\*

	Size	Investment, Millions of Dollars	Initial Catalyst, Millions of Dollars	Power Required, Megawatts	Catalysts and Chemicals, Millions of Dollars per Year	Labor, Shift Positions
<b>Onplot Facilities</b>						
First-stage Hydrotreater	61 Thousands of Barrels per Operating Day	82	9	5	16	2
Recycle Hydrocracker	18 Thousands of Barrels per Operating Day	45	5	1	2	2
Hydrogen Manufacture	2 x 61 Millions of Standard Cubic Feet per Operating Day	98	2	1	1	6
Middle Distillate Hydrotreater	33 Thousands of Barrels per Operating Day	19	0.5	1	-	1.5
Naphtha Hydrotreater	8 Thousands of Barrels per Operating Day	6	-	-	-	1
Reformer	8 Thousands of Barrels per Operating Day	9	0.5	-	-	1.5
MTW Plant	875 Gallons per Minute	14	-	1	-	2.5
Hydrogen Sulfide Recovery	6 Thousands of Pounds per Hour	4	-	-	-	0.5
Sulfur Production and Tail Gas Treating	62 Tons per Operating Day	6	-	-	-	1
Subtotal		285	17	9	19	16
Estimating Allowance		30				
<b>Offplot Facilities</b>						
Boiler Plant Facilities	580 Thousands of Pounds per Hour Steam	48	-	0.5	-	-
Cooling Tower	30 Thousands of Gallons per Minute	3	-	1	-	-
Electrical Distribution	30 Thousands of Kilowatt-Amperes	6	-	-	-	-
Feed and Product Storage Tanks	2.3 Millions of Barrels	29	-	0.5	-	-
Interconnecting and Tankfield Pipeways and Blending	-	28	-	-	-	-
Site Development	130 Acres	5	-	-	-	-
Relief and Flare System	-	5	-	-	-	-
Buildings and Maintenance Equipment	-	12	-	-	-	-
Waste Disposal	-	14	-	-	-	-
Land Purchase and Infrastructure Allowance	-	5	-	-	-	-
Subtotal		155				
Estimating Allowance		15				
<b>Working Capital</b>						
Feed Inventory	Half Full	14				
Intermediates and Finished Product Inventory	Half Full	14				
Spare Catalyst and Parts	-	10				
Accounts Receivable Minus Payable	30 Days	12				
Subtotal		50				

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil).  
 Notes: No allowance is included in these estimates for (a) shale source costs, (b) mining, handling, or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.

See Table XI for estimating bases.

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TABLE XIII

ESTIMATES OF INVESTMENTS AND UTILITIES  
CASE 2 REFINERY (HYDROTREATING - FLUID CATALYTIC CRACKING)  
REFINING OF PARAFFINIC SHALE OIL  
DOE CONTRACT #F-76-C-01-2315

50,000 barrels per Calendar Day Light Products\*

	Size	Investment, Millions of Dollars	Initial Catalyst, Millions of Dollars	Power Required, Megawatts	Catalysts and Chemicals, Millions of Dollars per Year	Labor, Shift Positions
<b>Capitol Facilities</b>						
First-Stage Hydrotreater	59 Thousands of Barrels per Operating Day	80	9	5	10	2
Fluid Catalytic Cracker	23 Thousands of Barrels per Operating Day	50	-	-	0.5	4.5
Hydrogen Manufacture	2 x 63 Millions of Standard Cubic Feet per Operating Day	73	2	1	0.5	6
Middle Distillate Hydrotreater	33 Thousands of Barrels per Operating Day	18	0.5	-	-	1.5
Reformer	7 Thousands of Barrels per Operating Day	6	-	-	-	1
Reformer	7 Thousands of Barrels per Operating Day	9	0.5	-	-	1.5
HRT Plant	875 Gallons per Minute	14	-	1	-	0.5
Hydrogen Sulfide Recovery	6 Thousands of Pounds per Hour	4	-	-	-	0.5
Sulfur Production and Tail Gas Treating	59 Tons per Operating Day	6	-	-	-	1
Subtotal		260	12	7	11	18.5
Estimating Allowance		25				
<b>Offplot Facilities</b>						
Sulfur Plant Facilities	628 Thousands of Pounds per Hour Steam	49	-	0.5	-	-
Cooling Tower	36 Thousands of Gallons per Minute	3	-	1	-	-
Electrical Distribution	30 Thousands of Kilovolt-Amperes	6	-	-	-	-
Feed and Product Storage Tanks	2.4 Millions of Barrels	30	-	-	-	-
Interconnecting and Tankfield Pipeways and Bleedoff		25	-	0.5	-	-
Site Development	140 Acres	5	-	-	-	-
Relief and Flare System		5	-	-	-	-
Buildings and Maintenance Equipment		12	-	-	-	-
Waste Disposal		15	-	-	-	-
Land Purchase and Infrastructure Allowance		5	-	-	-	-
Subtotal		155				
Estimating Allowance		1				
<b>Working Capital</b>						
Feed Inventory	Half Full	16				
Intermediate and Finished Product Inventory	Half Full	17				
Spare Catalyst and Parts		5				
Accounts Receivable Minus Payable		14				
Subtotal	30 Days	50				

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil).

Notes: No allowance is included in these estimates for (a) shale resource costs, (b) mining, distribution, or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.

See Table XI for estimating bases.

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TABLE XIV

ESTIMATES OF INVESTMENTS AND UTILITIES  
 CASE 3 REFINERY (DELAYED COILING - HYDROTREATING)  
 REFINING OF PALANO SHALE OIL  
 DOE CONTRACT #F-78-C-91-2113

50,000 Barrels per Calendar Day Light Products\*

	Size	Investment, Millions of Dollars	Initial Catalyst, Millions of Dollars	Power Required, Megawatts	Catalysts and Chemicals, Millions of Dollars per Year	Labor, Shift Positions
<b>Complot Facilities</b>						
First-stage Hydrotreater	58 Thousands of Barrels per Operating Day	57	5	2	5.5	2
Delayed Coker	71 Thousands of Barrels per Operating Day	69	-	1	-	4.5
Hydrogen Manufacture	72 Millions of Standard Cubic Feet per Operating Day	45	1	0.5	0.5	3
Middle Distillate Hydrotreater	47 Thousands of Barrels per Operating Day	22	0.5	0.5	-	1.5
Sulphur Hydrotreater	11 Thousands of Barrels per Operating Day	8	-	0.5	-	1
Reformer	11 Thousands of Barrels per Operating Day	12	0.5	-	-	1.5
Wet Plant	650 Gallons per Minute	13	-	0.5	-	0.5
Hydrogen Sulfide Recovery	6 Thousands of Pounds per Hour	4	-	-	-	0.5
Sulfur Production and Tail Gas Treating	61 Tons per Operating Day	6	-	-	-	1.5
Light Ends Recovery	6 Thousands of Barrels per Operating Day	5	-	-	-	1
Subtotal		249	7	5	6	17
Estimating		25				
<b>Offplot Facilities</b>						
Boiler Plant Facilities	660 Thousands of Pounds per Hour Steam	52	-	3.5	-	
Cooling Tower	30 Thousands of Gallons per Minute	3	-	1	-	
Electrical Distribution	30 Thousands of Kilovolt-Amperes	6	-	-	-	
Feed and Product Storage Tanks	2.8 Millions of Barrels	34	-	-	-	
Interconnecting and Tankfield Pipework and Blanding		24	-	-	-	
Site Development	160 Acres	6	-	0.5	-	
Relief and Flare System		5	-	-	-	
Buildings and Maintenance Equipment		12	-	-	-	
Waste Disposal		18	-	-	-	
Land Purchase and Infrastructure Allowance		5	-	-	-	
Subtotal		118				
Estimating Allowance		15				
<b>Working Capital</b>						
Feed Inventory	Half Full	18				
Intermediate and Finished Product Inventory	Half Full	13				
Spare Catalyst and Parts		3				
Accounts Receivable Minus Payable		6				
Subtotal	30 Days	40		2		9

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil).

Notes: No allowance is included in these estimates for (a) shale resource costs, (b) mining, distribution, or reorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.

See Table XI for estimating bases.

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TABLE XV

COMPARATIVE COST SUMMARY  
REFINING OF PARAHIO SHALE OIL  
DOE CONTRACT FF-76-C-01-2315

"Grass Roots" Refineries to  
Produce 50,000 Barrels per Calendar Day Light Products\*

Case	1		2		3	
	Hydrotreating - Hydrocracking		Hydrotreating - Fluid Catalytic Cracking		Severe Coking - Hydrotreating	
Refinery Type	Millions of Dollars**	Dollars per Barrel of Light Products	Millions of Dollars**	Dollars per Barrel of Light Products	Millions of Dollars**	Dollars per Barrel of Light Products
<b>Investment Costs</b>						
Onplot Investment	285		260		240	
Offplot Investment	155		155		160	
Estimating Allowance	45		40		40	
Initial Catalyst	15		10		5	
Working Capital	50		50		40	
Total Investment Costs	550	9.04	515	8.47	485	7.97
<b>Operating Costs</b>						
Catalysts and Chemicals	19		11		6	
Maintenance	10		9		9	
Taxes and Insurance	12		12		11	
Labor	4		5		4	
Power	3		2		2	
Boiler Plant Coal	3		3		4	
Total Operating Costs	51	2.79	42	2.30	36	1.97
<b>Total Processing Costs</b>		11.80***		10.80***		9.90***
Volume Ratio of Light Products to Shale Oil Feed		0.909		0.945		0.778

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil).

\*\*No allowance is included in these estimates for (a) shale resource costs, (b) mining, distribution or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.

\*\*\*Rounded to nearest \$0.10.

TABLE XVI

ESTIMATES OF INVESTMENTS AND UTILITIES  
 CASE 4 REFINERY (HYDROTREATING, HYDROCRACKING)  
 REFINING OF PANASO SHALE OIL  
 DOE CONTRACT #F-76-C-81-2315

100,000 Barrels per Calendar Day Light Products\*

	Size	Investment, Millions of Dollars	Initial Catalyst, Millions of Dollars	Power Required, Megawatts	Catalysts and Chemicals, Millions of Dollars per Year	Labor, Shift Positions
<b>Offplot Facilities</b>						
First-Stage Hydrotreater	2 x 61 Thousands of Barrels per Operating Day	132	18	10	33	4
Recycle Hydrocracker	36 Thousands of Barrels per Operating Day	68	10	2	4	2
Hydrogen Manufacture	4 x 81 Millions of Standard Cubic Feet per Operating Day	176	4	2	1	12
Middle Distillate Hydrotreater	66 Thousands of Barrels per Operating Day	27	1	1	-	1.5
Naphtha Hydrotreater	16 Thousands of Barrels per Operating Day	9	-	0.5	-	1
Reformer	16 Thousands of Barrels per Operating Day	14	1	0.5	-	1.5
Wax Plant	21 Gallons per Minute	21	-	3	-	0.5
Hydrogen Sulfide Recovery	12 Thousands of Pounds per Hour	5	-	-	-	0.5
Sulfur Production and Tail Gas Treating	124 Tons per Operating Day	8	-	-	-	1
Subtotal		468	34	19	-	24
Estimating Allowance		45				
<b>Offplot Facilities</b>						
Boiler Plant Facilities						
Cooling Tower	1,150 Thousands of Pounds per Hour Steam	70	-	0.5	-	-
Electrical Distribution	60 Thousands of Gallons per Minute	5	-	2	-	-
Feed and Product Storage Tanks	70 Thousands of Kilovolt-amperes	10	-	-	-	-
Interconnections and Tankfield	4.6 Millions of Barrels	42	-	-	-	-
Pipeways and Blending		43	-	0.5	-	-
Site Development	200 Acres	7	-	-	-	-
Relief and Flare System		7	-	-	-	-
Buildings and Maintenance Equipment		14	-	-	-	-
Waste Disposal		17	-	-	-	-
Land Purchase and Infrastructure Allowance		5	-	-	-	-
Subtotal		228	-	-	-	-
Estimate Allowance		25				
<b>Working Capital</b>						
Feed Inventory	Half Full	28				
Intermediates and Finished Product Inventory	Half Full	27				
Spare Catalyst and Parts		21				
Accounts Receivable Minus Payable	30 Days	24				
Subtotal		100				

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil).  
 Notes: No allowance is included in these estimates for (a) shale resource costs, (b) mining, distribution, or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.  
 See Table XI for estimating bases.

TABLE XVII

ESTIMATES OF INVESTMENTS AND UTILITIES  
CASE 5 REFINERY (HYDROTREATING - FLUID CATALYTIC CRACKING)  
REFINING OF PARAFFINIC CRACKING OIL  
DOE CONTRACT EF-76-C-01-2115

100,000 Barrels per Calendar Day Light Products

Facility	Size	Investment, Millions of Dollars	Initial Catalyst, Millions of Dollars	Power Acquired, Megawatts	Catalysts and Chemicals, Millions of Dollars per Year	Labor, Shift Positions
<u>Offplot Facilities</u>						
First-Stage Hydrotreater	2 x 59 Thousands of Barrels per Operating Day	130	18	8	20	4
Fluid Catalytic Cracker	46 Thousands of barrels per Operating Day	72	-	1	1	4.5
Hydrogen Manufacture	3 x 84 Millions of Standard Cubic Feet per Operating Day	122	3	1	1	9
Middle Distillate Hydrotreater	56 Thousands of Barrels per Operating Day	27	1	1	-	1.5
Naphtha Hydrotreater	16 Thousands of Barrels per Operating Day	8	-	0.5	-	1
Reformer	16 Thousands of Barrels per Operating Day	12	1	0.5	-	1.5
HMT Plant	1,750 Gallons per Minute	21	-	3	-	0.5
Hydrogen Sulfide Recovery	12 Thousands of Pounds per Hour	5	-	-	-	0.5
Sulfur Production and Tail Gas Treating	120 Tons per Operating Day	8	-	-	-	1
Subtotal		405	23	15	22	23.5
Estimating Allowance		40				
<u>Offplot Facilities</u>						
Boiler Plant Facilities	1,250 Thousands of Pounds per Hour Steam	75	-	0.5	-	
Cooling Tower	60 Thousands of Gallons per Minute	5	-	2	-	
Electrical Distribution	60 Thousands of Kilovolt-Amperes	9	-	-	-	
Feed and Product Storage Tanks	4.8 Millions of Barrels	44	-	-	-	
Interconnecting and Tankfield Pipeways and Blanding	-	48	-	1	-	
Site Development	210 Acres	7	-	-	-	
Railcar and Pallet System	-	7	-	-	-	
Buildings and Maintenance Equipment	-	14	-	-	-	
Waste Disposal	-	19	-	-	-	
Land Purchase and Infrastructure Allowance	-	5	-	-	-	
Subtotal		225				
Estimating Allowance		25				
<u>Working Capital</u>						
Feed Inventory	Half Full	27				
Intermediates and Finished Product Inventory	Half Full	32				
Spare Catalyst and Parts	-	9				
Accounts Receivable Minus Payables	30 Days	27				
Subtotal		95		3.5		12

Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (90.2 OIL).  
Notes: No allowance is included in these estimates for (a) shale resource costs, (b) mining, handling, or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.  
See Table XI for estimating bases.

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TABLE XVIII

ESTIMATES OF INVESTMENTS AND UTILITIES  
CASE 6 REFINERY (RELATED COKING - HYDROTREATING)  
REFINING OF PARANO SHALE OIL  
DOE CONTRACT EP-76-C-81-2315

100,000 Barrels per Calendar Day Light Products

	Size	Investment, Millions of Dollars	Initial Catalyst Millions of Dollars	Power Required, Megawatts	Catalysts and Chemicals, Millions of Dollars per Year	Labor, Shift Positions
<b>Onplot Facilities</b>						
First-Stage Hydrotreater	2 x 58 Thousands of Barrels per Operating Day	88	11	2	10	4
Delayed Cohers	2 x 71 Thousands of Barrels per Operating Day	118	-	1	-	9
Hydrogen Manufacture	2 x 72 Millions of Standard Cubic Feet per Operating Day	76	2	1	0.5	6
Middle Distillate Hydrotreater	84 Thousands of Barrels per Operating Day	32	1	1	0.5	1.5
Naphtha Hydrotreater	22 Thousands of Barrels per Operating Day	11	-	0.5	-	1
Reformer	22 Thousands of Barrels per Operating Day	17	1	0.5	-	1.5
WWT Plant	1,300 Gallons per Minute	18	-	2	-	0.5
Hydrogen Sulfide Recovery	12 Thousands of Pounds per Hour	5	-	-	-	0.5
Sulfur Production and Tail Gas Treating	122 Tons per Operating Day	8	-	-	-	1.4
Light Ends Recovery	8 Thousands of Barrels per Operating Day	7	-	-	-	1.5
Subtotal		380	15	8	11	36.5
Estimating Allowance		35				
<b>Offplot Facilities</b>						
Boiler Plant Facilities						
Cooling Tower	1,320 Thousands of Pounds per Hour Steam	78	-	0.5	-	-
Electrical Distribution	60 Thousands of Gallons per Minute	5	-	2	-	-
Feed and Product Storage Tanks	50 Thousands of Kilovolt-Amperes	8	-	-	-	-
Interconnecting and Tankfield Pipeways and Blending	5.6 Millions of Barrels	50	-	-	-	-
Site Development	240 Acres	38	-	0.5	-	-
Relief and Flare System	-	8	-	-	-	-
Buildings and Maintenance Equipment	-	7	-	-	-	-
Waste Disposal	-	14	-	-	-	-
Land Purchase and Infrastructure Allowance	-	22	-	1	-	-
Subtotal		5	-	-	-	-
Estimating Allowance		235				
Working Capital		25		4		12
Feed Inventory	Half Full	38				
Intermediate and Finished Product Inventory	Half Full	28				
Spare Catalyst and Parts		6				
Accounts Receivable Minus Payable		11				
Subtotal	30 Days	85				

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil).

Notes: No allowance is included in these estimates for (a) shale resource costs, (b) mining, distribution, or retorting of shale.

(c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.

See Table XI for estimating bases.

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TABLE XIX

COMPARATIVE COST SUMMARY  
REFINING OF PANASO SHALE OIL  
DOE CONTRACT EF-76-C-01-2315

"Grass Roots" Refineries to  
Produce 100,000 Barrels per Calendar Day Light Products\*

Case	Refinery Type	4		5		6	
		Hydrotreating - Hydrocracking	Dollars per Barrel of Light Products	Hydrotreating - Fluid Catalytic Cracking	Dollars per Barrel of Light Products	Severe Coking - Hydrotreating	Dollars per Barrel of Light Products
		Millions of Dollars**		Millions of Dollars**		Millions of Dollars**	
	Notes (See Table XI)						
	<u>Investment Costs</u>						
	Onplot Investment	460		405		380	
	Offplot Investment	220		225		235	
	Estimating Allowance	70		65		60	
	Initial Catalyst	35		25		15	
	Working Capital	100		95		85	
	Total Investment Costs	885	7.27	815	6.70	775	6.37
		Millions of Dollars per Year		Millions of Dollars per Year		Millions of Dollars per Year	
	<u>Operating Costs</u>						
	Catalysts and Chemicals	38		22		11	
	Maintenance	15		14		14	
	Taxes and Insurance	19		18		18	
	Labor	6		6		6	
	Power	6		5		3	
	Boiler Plant Coal	6		7		8	
	Total Operating Costs	90	2.47	72	1.97	60	1.64
	<b>Total Processing Costs</b>		<b>9.70***</b>		<b>8.70***</b>		<b>8.00***</b>
	Volume Ratio of Light Products to Shale Oil Feed		0.909		0.945		0.778

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil).  
\*\*No allowance is included in these estimates for (a) shale resource costs, (b) mining, distribution or  
retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution  
and transportation from the refinery.  
\*\*\*Rounded to nearest \$0.10.

TABLE XX

STOCK BALANCE FOR CASES 7 AND 8 MODULES  
 HYDROTREATING TO PRODUCE 100,000 BARRELS PER  
 CALENDAR DAY SYNTHETIC CRUDE  
 REFINING OF PARAHO SHALE OIL  
 DOE CONTRACT EF-76-C-01-2315

	Processing				Products			
	Input	First-Stage Hydrotreating	Hydrogen Manufacture	H <sub>2</sub> S Recovery and Sulfur Plant	Offplot Boiler Plant	Refinery Fuel	Synthetic Crude	By-Products
Barrels per Calendar Day								
Raw Shale Oil	110,000	(110,000)						
Coal or Shale, Equivalent Fuel Oil	1,900				(1,900)			
Fuel Gas, Equivalent Fuel Oil		4,100	(4,100)					
Naphtha		16,350	(7,855)				8,495	
350°F <sup>+</sup> Product		101,390					91,505	
Total Liquid Product		121,840					100,000	
Total Feed	11,900	110,000	11,955					
Gain (Loss)		11,840	(11,955)					
Fuel		(1,780)	(8,075)			9,885		
Hydrogen, Millions of Standard Cubic Feet per Calendar Day		(251)/22	(22)/251					
Tons per Calendar Day								
Hydrogen Sulfide		123.8						110.5
Sulfur								461
Ammonia		461						

Note: Parentheses denote a negative quantity, i.e., consumption.

TABLE XXI

PROPERTIES OF SYNTHETIC CRUDE FROM  
CASES 7 AND 8 HYDROTREATING MODULES  
REFINING OF PARAHO SHALE OIL  
DOE CONTRACT EF-76-C-01-2315

Gravity, °API	35.8
Aniline Point, °F	175
<u>ASTM D 1160 Distillation, °F</u>	
St/5	300/360
10/30	400/500
50	600
70/90	750/900
95/EP	950/1050
<u>Parts per Million</u>	
Sulfur	75
Nitrogen	1000
Oxygen	170
Pour Point, °F	100
Ramsbottoms Carbon, %	0.10
Viscosity at 122°F, Centistokes	10
Viscosity at 210°F, Centistokes	3

TABLE XXII

ESTIMATES OF INVESTMENT AND UTILITIES  
CASE 7 MOODLE  
REFINING OF PARABO SHALE OIL  
DOE CONTRACT EP-76-C-01-2313

Hydrotreating to Produce 100,000 Barrels per Calendar Day  
Synthetic Crude at Referts

	Size	Investment, Millions of Dollars	Initial Catalyst, Millions of Dollars	Power Required, Megawatts	Catalysts and Chemicals, Millions of Dollars per Year	Labor, Shift Positions
<u>Onplot Facilities</u>						
First-Stage Hydrotreater	2 x 61 Thousands of Barrels per Operating Day	144	18	9	20	4
Hydrogen Manufacture	4 x 70 Millions of Standard Cubic Feet per Operating Day	166	3	1	1	9
MWT Plant	1,700 Gallons per Minute	21	-	2	-	0.5
Hydrogen Sulfide Recovery	12 Thousands of Pounds per Hour	5	-	-	-	0.5
Sulfur Production and Tail Gas Treating	120 Tons per Operating Day	9	-	-	-	1.0
Subtotal		345	21	12	21	15
Estimating Allowance		35				
<u>Offplot Facilities</u>						
Boiler Plant Facilities	1,130 Thousands of Pounds per Hour Steam	48	-	0.5	-	-
Cooling Tower	50 Thousands of Gallons per Minute	4	-	1	-	-
Electrical Distribution	50 Thousands of Kilovolt-Amperes	8	-	-	-	-
Feed and Product Storage Tanks	2.3 Millions of Barrels	24	-	-	-	-
Interconnecting and Tankfield Pipeways	-	28	-	0.5	-	-
Site Development	110 Acres	4	-	-	-	-
Relief and Flare System	-	6	-	-	-	-
Building and Maintenance Equipment	-	5	-	-	-	-
Waste Disposal	-	10	-	-	-	-
Land Purchase and Infrastructure Allowance	-	3	-	-	-	-
Subtotal		140				
Estimating Allowance		15				
<u>Working Capital</u>						
Feed Inventory	Half Full	13				
Intermediate and Finished Product Inventory	Half Full	14				
Spare Catalyst and Parts	-	9				
Accounts Receivable Minus Payable	30 Days	9				
Subtotal		45		2		6

Notes: No allowance is included in these estimates for (a) shale resource costs, (b) mining, handling, or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.

See Table XI for estimating bases.

TABLE XXIII

ESTIMATES OF INVESTMENTS AND UTILITIES

Case 8 MODULE  
REFINING OF PARAHIO SHALE OIL  
DOE CONTRACT EP-76-C-01-2315

Hydrotreating to Produce 100,000 Barrels per Calendar Day  
Synthetic Crude Ahead of Existing Refinery

	Size	Investment, Millions of Dollars	Initial Catalyst, Millions of Dollars	Power Required, Megawatts	Catalysts and Chemicals, Millions of Dollars per Year	Labor, Shift Positions
<b>Onplot Facilities</b>						
First-Stage Hydrotreater	2 x 61 Thousands of Barrels per Operating Day	133	18	9	20	4
Hydrogen Manufacture	4 x 70 Millions of Standard Cubic Feet per Operating Day	153	3	1	1	9
WWT Plant	1,700 Gallons per Minute	21	-	2	1	0.5
Hydrogen Sulfide Recovery	12 Thousands of Pounds per Hour	5	-	-	-	0.5
Sulfur Production and Tail Gas Treating	120 Tons per Operating Day	8	-	-	-	1.0
Subtotal		320	21	17	21	15
Estimating Allowance		30				
<b>Offplot Facilities</b>						
Boiler Plant Facilities	1,130 Thousands of Pounds per Hour Steam	73	-	0.5	-	-
Cooling Tower	50 Thousands of Gallons per Minute	4	-	-	-	-
Electrical Distribution	60 Thousands of Kilovolt-Amperes	8	-	1	-	-
Feed and Product Storage Tanks	2.3 Millions of Barrels	21	-	-	-	-
Interconnecting and Tankfield Pipeways	-	25	-	-	-	-
Site Development	110 Acres	4	-	1	-	-
Relief and Flare System	-	6	-	-	-	-
Buildings and Maintenance Equipment	-	7	-	-	-	-
Waste Disposal	-	14	-	-	-	-
Land Purchase and Infrastructure Allowance	-	3	-	-	-	-
Subtotal		165				
Estimating Allowance		15				
<b>Working Capital</b>						
Feed Inventory	Half Full	13				
Intermediate and Finished Product Inventory	Half Full	14				
Spare Catalyst and Parts	-	9				
Accounts Receivable Minus Payable	30 Days	9				
Subtotal		45				

Notes: No allowance is included in these estimates for (a) shale resource costs, (b) mining, handling, or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.

See Table XI for estimating bases.

TABLE XXIV

COMPARATIVE COST SUMMARY  
REFINING OF PARAHO SHALE OIL  
DOE CONTRACT EF-76-C-01-2315

Incremental Facilities to  
Produce 100,000 Barrels per Calendar Day Synthetic Crude

Case	7		8		
	Hydrotreating at Retorts		Hydrotreating Ahead of Existing Refinery		
	Notes (See Table XI)	Millions of Dollars*	Dollars per Barrel of Light Products	Millions of Dollars*	Dollars per Barrel of Light Products
<u>Investment Costs</u>					
Onplot Investment	(1), (2), (3)	345		320	
Offplot Investment	(1), (2), (3)	140		165	
Estimating Allowance	(4)	50		50	
Initial Catalyst		20		20	
Working Capital	(5), (6), (7)	45		45	
Total Investment Costs	(8)	600	4.93	600	4.93
		Millions of Dollars per Year		Millions of Dollars per Year	
<u>Operating Costs</u>					
Catalysts and Chemicals		21		21	
Maintenance	(11)	11		11	
Taxes and Insurance	(12)	13		13	
Labor	(13)	3		3	
Power	(14)	4		4	
Boiler Plant Coal	(15)	4		4	
Total Operating Costs		56	1.53	56	1.53
Total Processing Costs			6.50**		6.50**
Volume Ratio of Light Products to Shale Oil Feed			0.909		0.909

\*No allowance is included in these estimates for (a) shale resource costs, (b) mining, distribution or retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution and transportation from the refinery.

\*\*Rounded to nearest \$0.10.

3-28-78

HAF RD 781599

TABLE XIV

STOCK BALANCE SUMMARY  
REFINING OF PIRABO SHALE OIL  
DDE CONTRACT EP-78-C-81-2115

Case	1	2	3	4	5	6	7	8
Coal	50,000 Barrels per Calendar Day	100,000 Barrels per Calendar Day	100,000 Barrels per Calendar Day	100,000 Barrels per Calendar Day	100,000 Barrels per Calendar Day	100,000 Barrels per Calendar Day	100,000 Barrels per Calendar Day	100,000 Barrels per Calendar Day
Processing	Hydrotreating - Hydrocracking	Hydrotreating - Fluid Catalytic Cracking	Severe Coking - Hydrotreating	Hydrotreating - Hydrocracking	Hydrotreating - Fluid Catalytic Cracking	Severe Coking - Hydrotreating	Hydrotreating	Hydrotreating
Shale Oil Input	55,000	52,900	64,300	110,000	103,800	128,600	110,000	110,000
Coal Input for Boiler Plant Fuel, Barrels per Calendar Day of Equivalent Fuel Oil	1,388	1,490	1,790	2,760	2,980	3,580	1,900	1,900
Processing, Barrels per Operating Day**								
First-Stage Hydrotreating	61,100	58,800	57,800	122,200	117,600	115,600	122,200	122,200
Recycle Hydrocracking	17,600	-	-	35,200	45,800	-	-	-
Fluid Catalytic Cracking	-	22,900	71,400	-	-	142,000	-	-
Delayed Coking	-	-	-	-	-	-	-	-
Residue Hydrotreating and Reforming	8,300	7,200	11,400	16,600	14,400	22,800	-	-
Middle Distillate Hydrotreating	32,800	33,200	42,800	65,600	64,400	84,000	-	-
Light Ends Recovery	-	-	3,600	-	-	7,200	-	-
Hydrogen Manufacture, Standard Cubic Feet per Operating Day	162 · 10 <sup>6</sup>	126 · 10 <sup>6</sup>	72 · 10 <sup>6</sup>	324 · 10 <sup>6</sup>	252 · 10 <sup>6</sup>	244 · 10 <sup>6</sup>	279 · 10 <sup>6</sup>	279 · 10 <sup>6</sup>
Hydrogen Sulfide Recovery, Pounds per Hour	6,000	6,000	6,000	12,000	12,000	12,000	12,000	12,000
Sulfur Plant, Short Tons per Day	62	59	61	124	118	122	123	123
NSR Plant, Gallons per Minute	875	875	650	1,750	1,750	1,300	1,700	1,700
Light Products, Barrels per Calendar Day								
Liquid Petroleum Gas (LPG)	-	-	1,358	-	-	2,700	-	-
Motor Gasoline	9,175	19,860	16,685	17,350	39,720	21,210	-	-
Kerosene Jet Fuel	10,540	-	-	11,770	-	-	-	-
Distillate Fuel (No. 2 Oil)	29,885	30,140	38,045	52,770	60,280	76,090	-	-
	50,000	50,000	50,000	100,000	100,000	100,000	100,000	100,000
Other Products, Short Tons per Calendar Day								
Sulfur	55.5	53.4	54.5	111	106.8	109	110.5	110.5
Ammonia	232.1	221.0	185.1	464.2	442	378.2	461	461
Other	-	-	1,995	-	-	3,990	-	-
Refinery Fuel, Barrels per Calendar Day (Internally Supplied)	(6,490)	(5,050)	(6,640)	(12,980)	(10,100)	(13,280)	(9,865)	(9,855)

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil). \*\*Total capacity shown. Number of parallel units, where necessary, shown on cost breakdown tables for individual cases.

TABLE LXVI

OVERALL COST SUMMARY  
REFINING OF PARADO SHALE OIL  
DOE CONTRACT EP-78-C-01-1313

Case	Products	Location	Processing	Total Investment,** Millions of Dollars	Operating Cost,** Millions of Dollars per Year	Total Cost, Dollars per Barrel of Light Products or Synthetic Crude***	Volume Ratio of Light Products to Shale Oil Feed
1	50,000 Barrels per Calendar Day Light Products*	Remote Rocky Mountain Area	Hydrotreating - Hydrocracking	550	51	11.80	0.909
2			Hydrotreating - Fluid Catalytic Cracking	515	42	10.80	0.945
3			Delayed Coking - Hydrotreating	485	36	9.90	0.778
4	100,000 Barrels per Calendar Day Light Products*	Midcontinent or Rocky Mountain Near Urban Center	Hydrotreating - Hydrocracking	885	90	9.70	0.909
5			Hydrotreating - Fluid Catalytic Cracking	815	72	8.70	0.945
6			Delayed Coking - Hydrotreating	775	60	8.00	0.778
7	100,000 Barrels per Calendar Day Hydrotreated Synthetic Crude	Rocky Mountain at Shale Retorts Midcontinent or Rocky Mountain Near Urban Center	Hydrotreating	600	56	6.50	0.909
8			Hydrotreating	600	56	6.50	0.909

\*Light products include LPG, motor gasoline, kerosene jet fuel, and diesel fuel (No. 2 Oil).  
\*\*No allowance is included in these estimates for (a) shale resource costs, (b) mining, distribution or  
retorting of shale, (c) shale oil transportation to the refinery, or (d) refined product distribution  
and transportation from the refinery.

\*\*\*Processing costs are based on a capital charge at 15% value of money and  
are rounded to the nearest \$0.10/barrel.

**TABLE XXVII**  
**EFFECT OF RATE OF RETURN ON PROCESSING COST**  
**REFINING OF PARAHO SHALE OIL**

Grass Roots Refineries Producing Finished Light Products

Assumed Rate of Return	Costs in Dollars per Barrel					
	Case 1 Hydrotreating Hydrocracking, 50,000 BPCD	Case 2 Hydrotreating FCC, 50,000 BPCD	Case 3 Delayed Coking Hydrotreating, 50,000 BPCD	Case 4 Hydrotreating Hydrocracking, 100,000 BPCD	Case 5 Hydrotreating FCC, 100,000 BPCD	Case 6 Delayed Coking Hydrotreating, 100,000 BPCD
10%	8.80	7.90	7.30	7.30	6.40	5.90
15%	11.80	10.80	9.90	9.70	8.70	8.00
20%	15.50	14.20	13.10	12.70	11.40	10.60

**TABLE XXVIII**  
**EFFECT OF RATE OF RETURN ON**  
**PROCESSING COSTS**  
**REFINING OF PARAHO SHALE OIL**  
  
**Synthetic Crude Module at Retorts or**  
**Ahead of Existing Refinery**  
**Processing Costs in Dollars/Barrel of Product**

<u>Assumed</u> <u>Rate of Return</u>	<u>Cases 7 and 8 for</u> <u>Synthetic Crude,</u> <u>100,000 BPCD</u>
10%	4.80
15%	6.50
20%	8.40

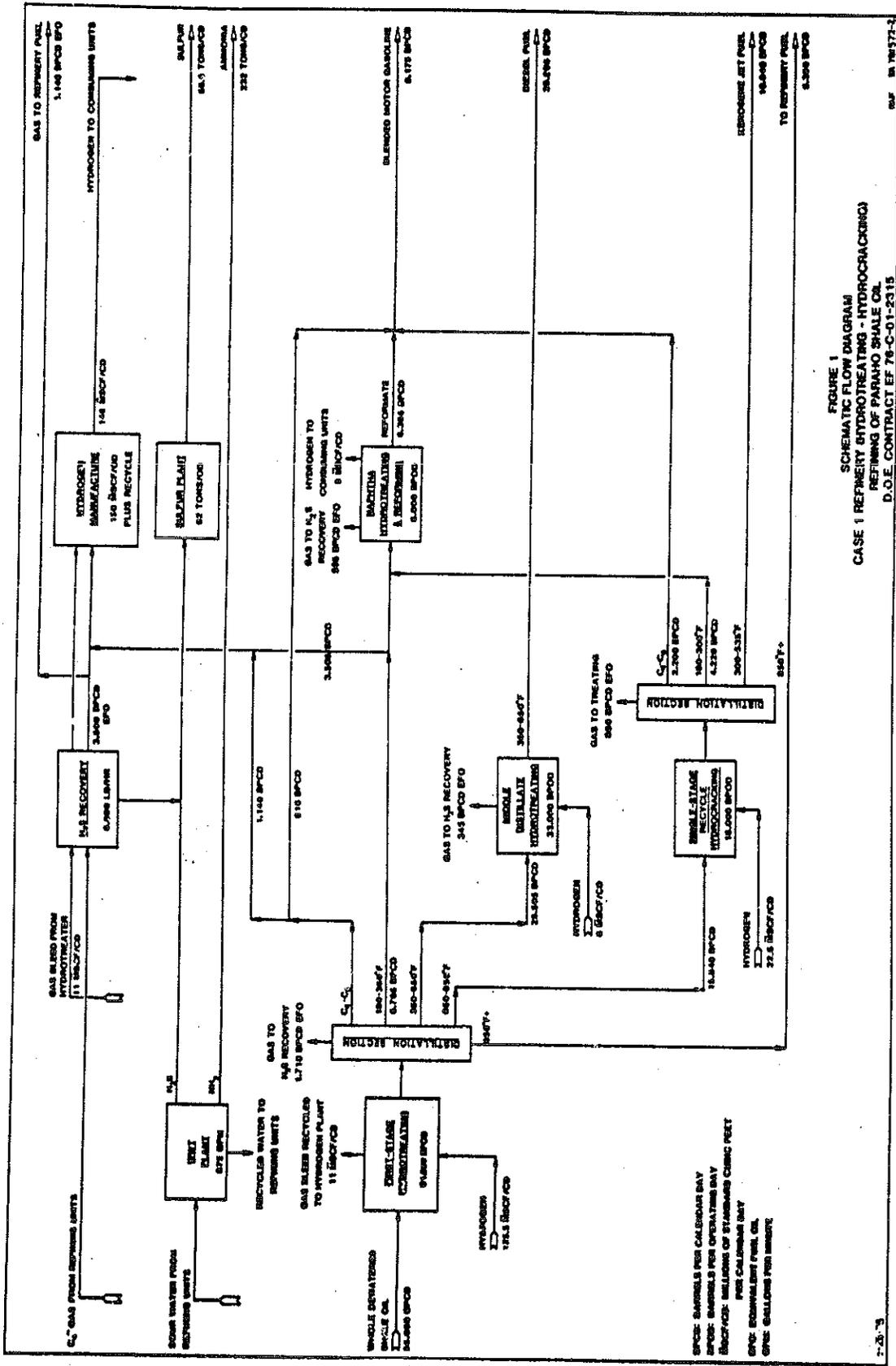
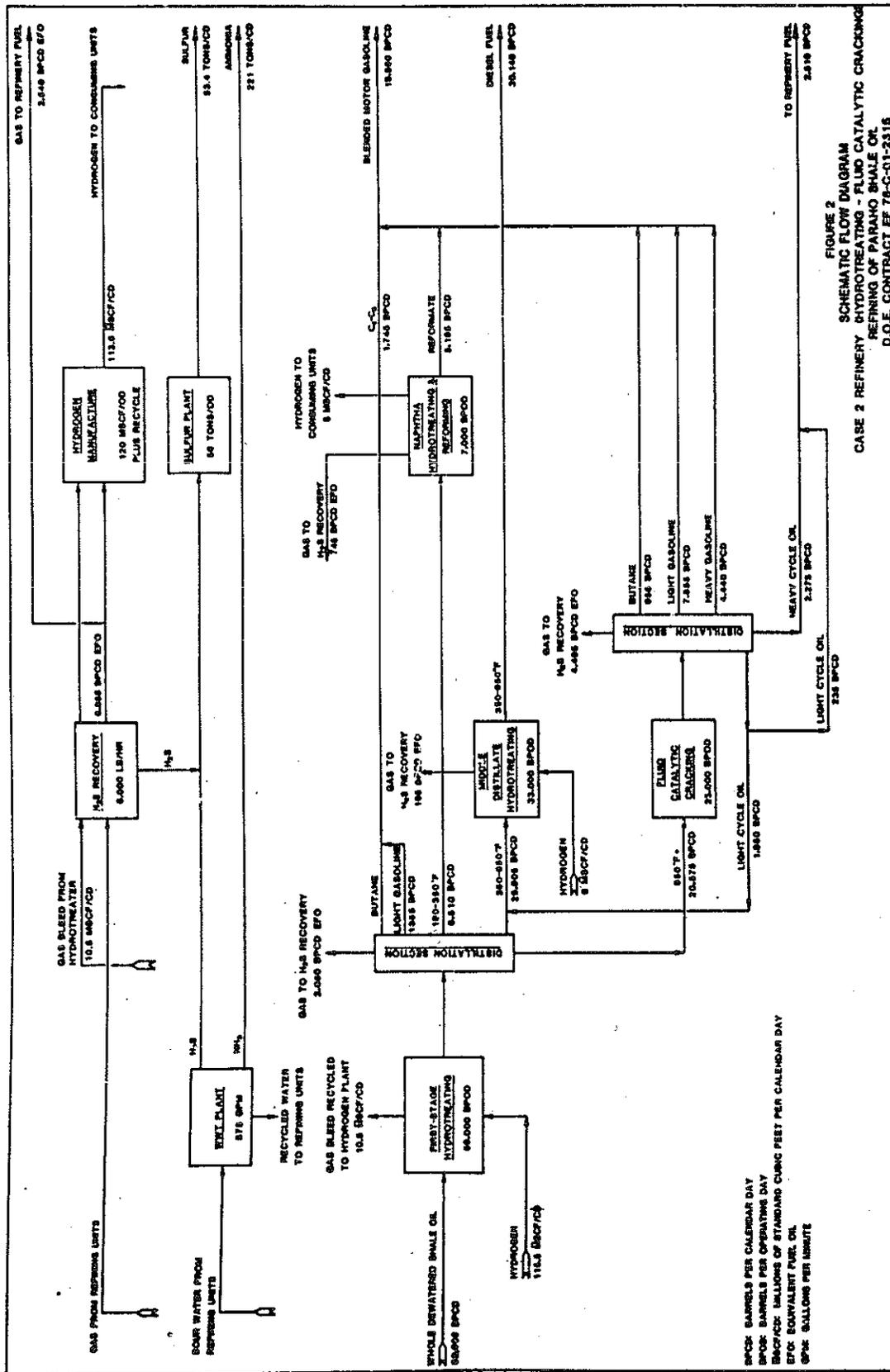


FIGURE 1  
 SCHEMATIC FLOW DIAGRAM  
 CASE 1 REFINERY HYDROTREATING - HYDROCRACKING  
 REFINING OF PARANO SHALE OIL  
 D.O.E. CONTRACT EF 78-C-01-2315



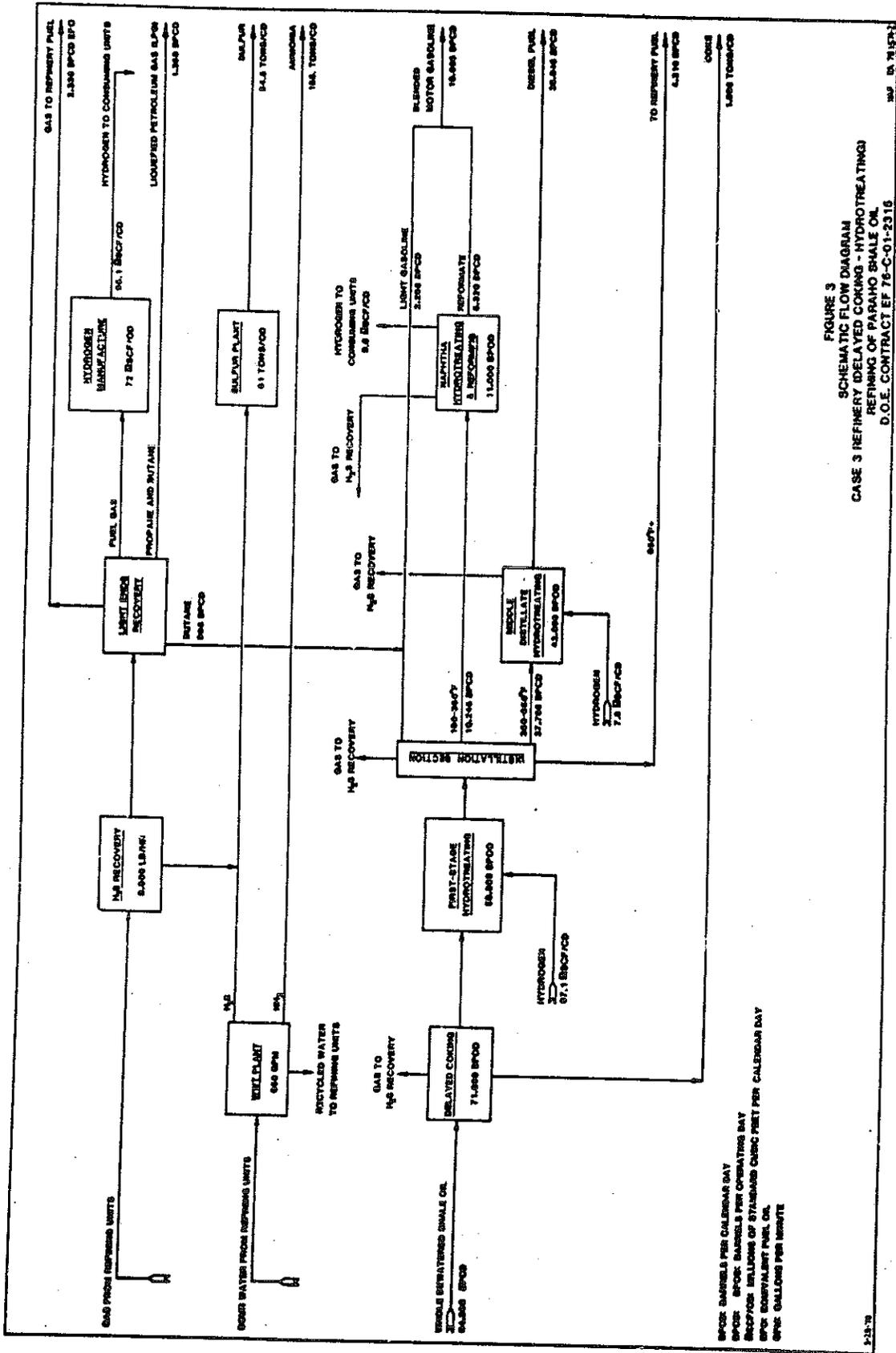
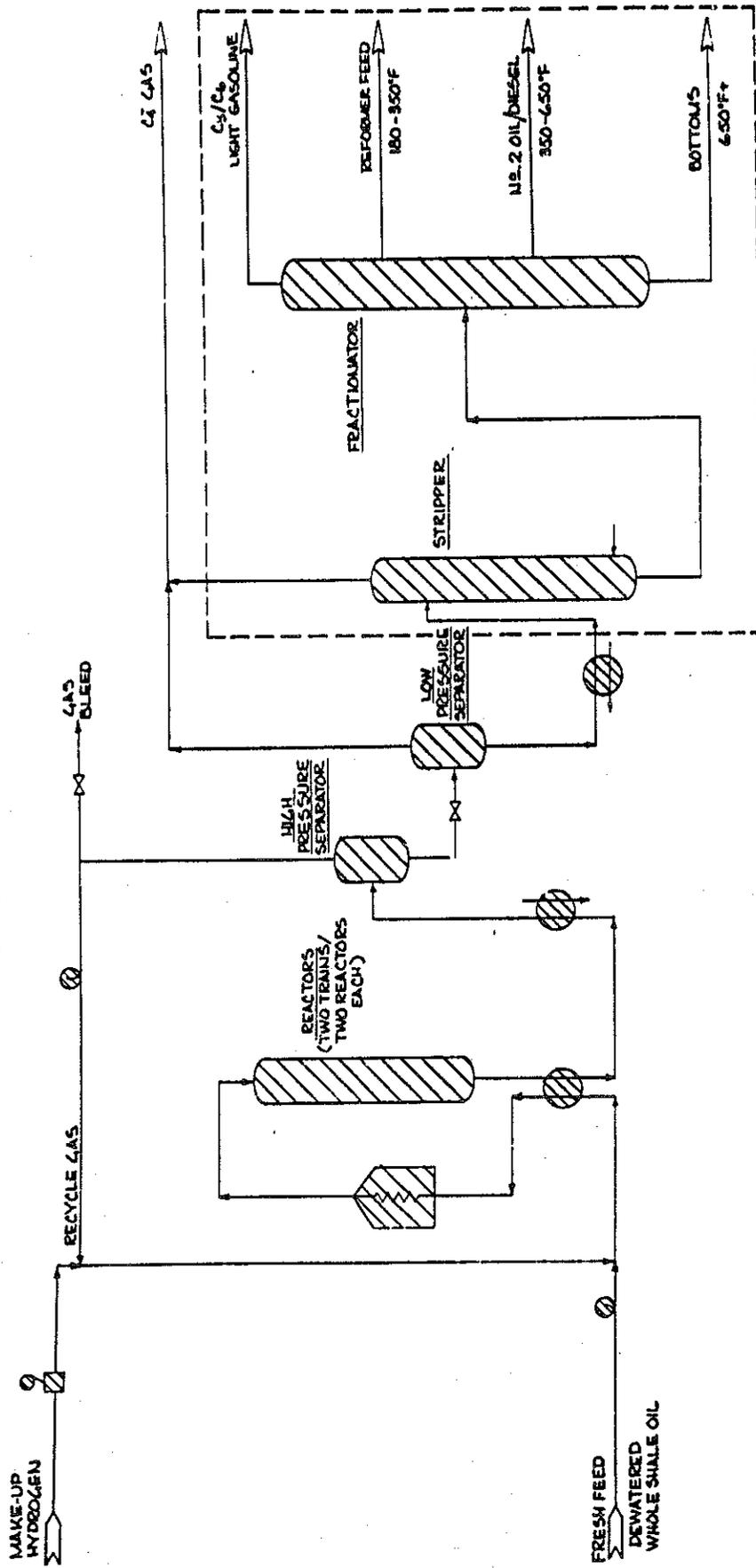


FIGURE 3  
SCHEMATIC FLOW DIAGRAM  
CASE 3 REFINERY (DELAYED COKING - HYDROTREATING)  
REFINING OF PARANO SHALE OIL  
D.O.E. CONTRACT EF 76-C-01-2316

**FIGURE 4**  
**SIMPLIFIED FLOW DIAGRAM OF CHEVRON FIRST-STAGE HYDROTREATING**  
**REFINING OF PARAHO SHALE OIL**  
**D.O.E. CONTRACT EF-76-C-01-2315**

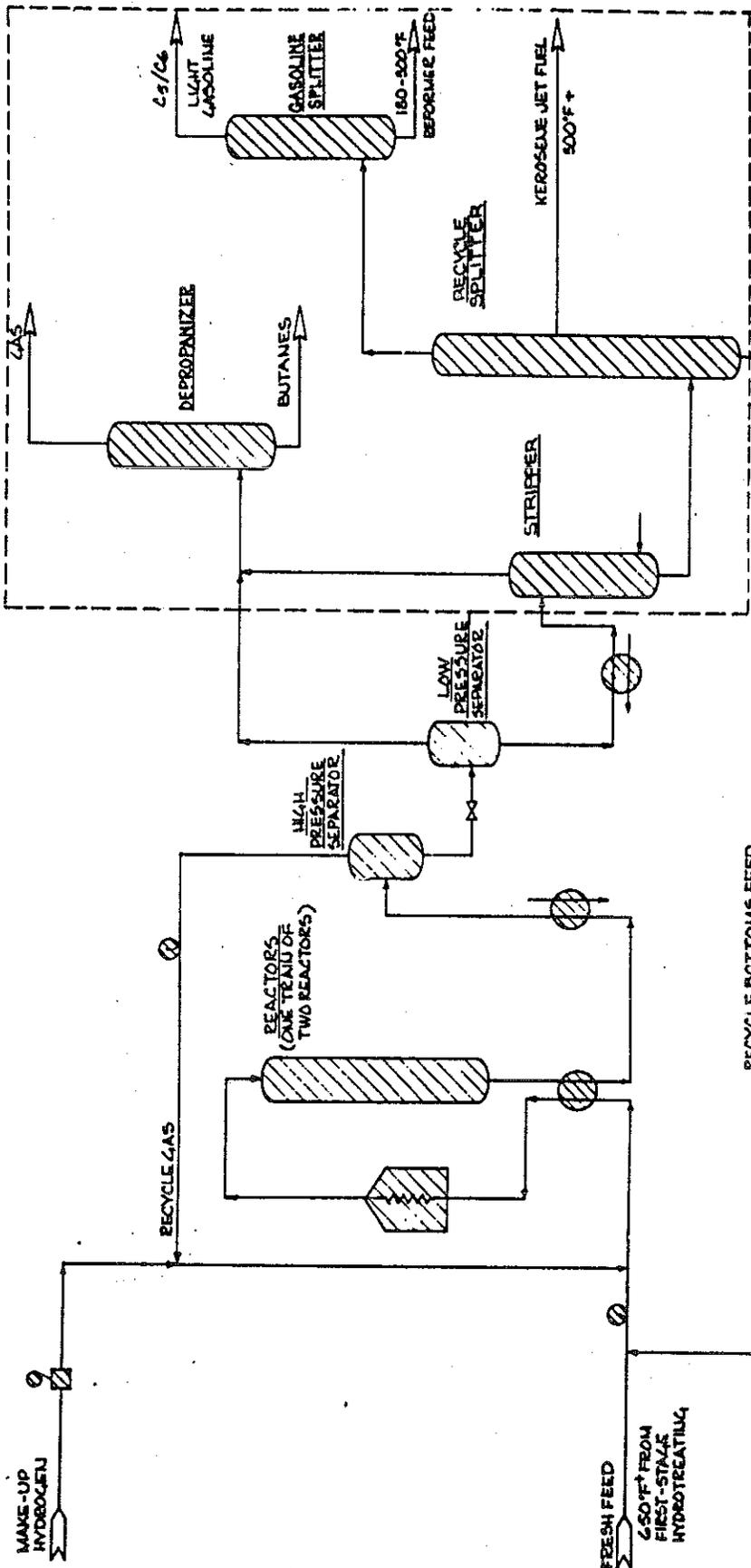


**DISTILLATION SECTION FOR CASE 2**

NAF RC 780388

1-19-76

FIGURE 5  
 SIMPLIFIED FLOW DIAGRAM OF CHEVRON SINGLE-STAGE RECYCLE ISOCRACKING  
 REFINING OF PARAHO SHALE OIL  
 D.O.E. CONTRACT EF-76-C-01-2315

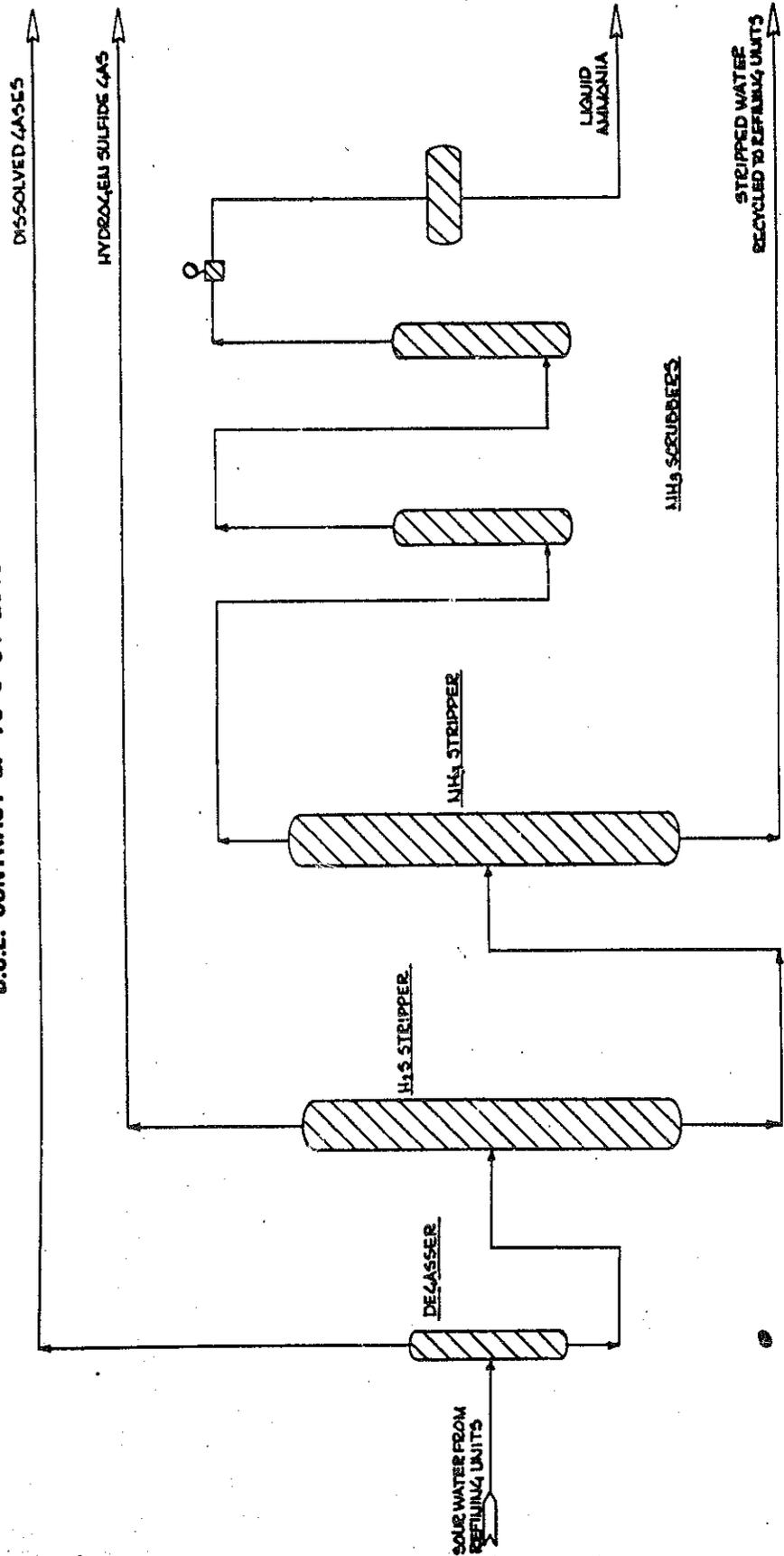


DISTILLATION SECTION FOR CASE 1

1-19-78

HAF RC780569

FIGURE 6  
 SIMPLIFIED FLOW DIAGRAM OF CHEVRON WWT PROCESS  
 REFINING OF PARAHO SHALE OIL  
 D.O.E. CONTRACT EF-76-C-01-2315



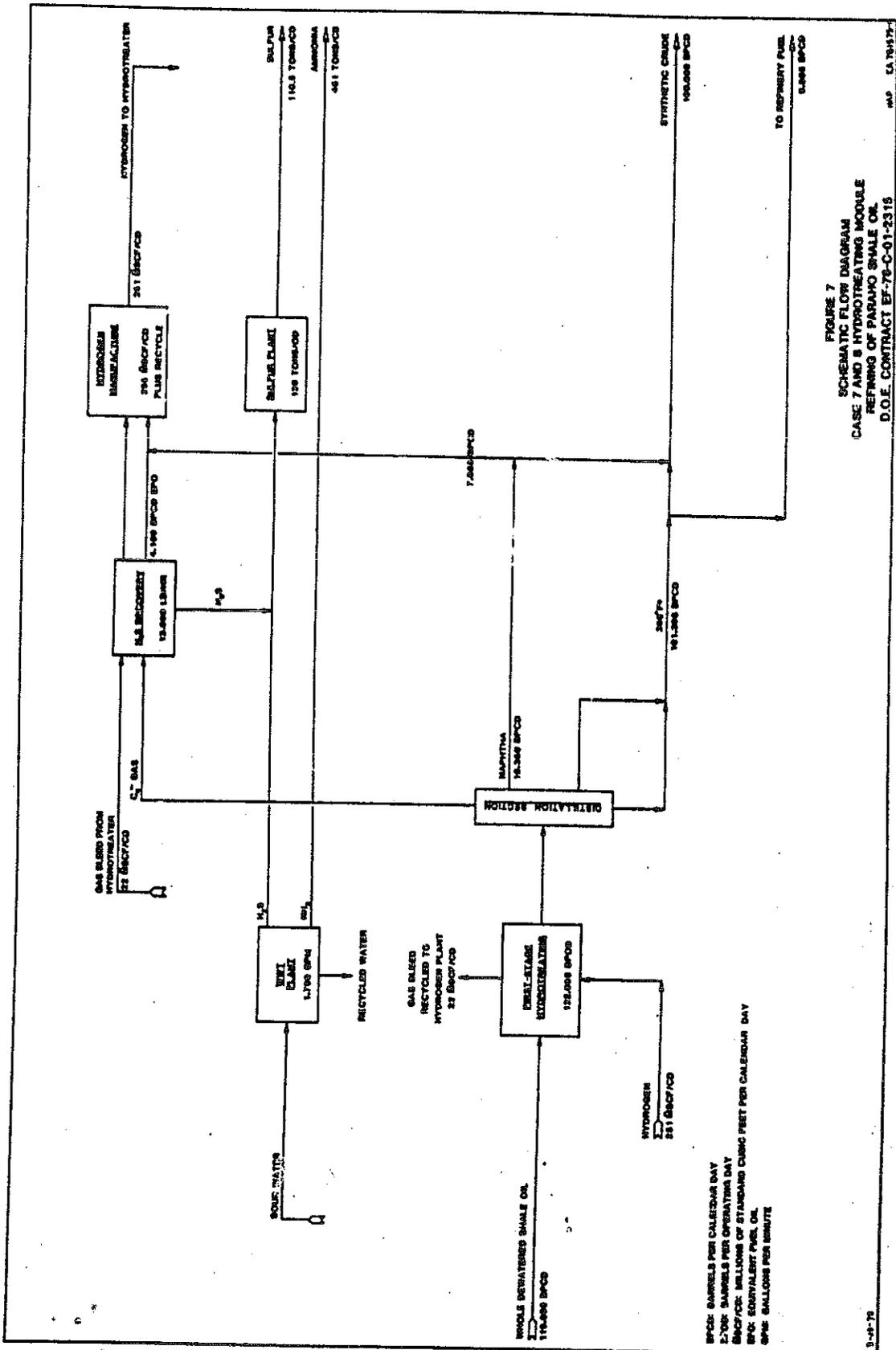


FIGURE 7  
 SCHEMATIC FLOW DIAGRAM  
 CASE 7 AND 8 HYDROTREATING MODULE  
 REFINING OF PARANO SHALE OIL  
 D.O.E. CONTRACT EF-78-C-01-2316

FIGURE 8  
 EFFECT OF RATE OF RETURN ON PROCESSING COSTS FOR  
 REFINING OF CRUDE PARAHO SHALE OIL INTO  
 FINISHED LIGHT PRODUCTS

Grass Roots 50,000 BPCD Refinery

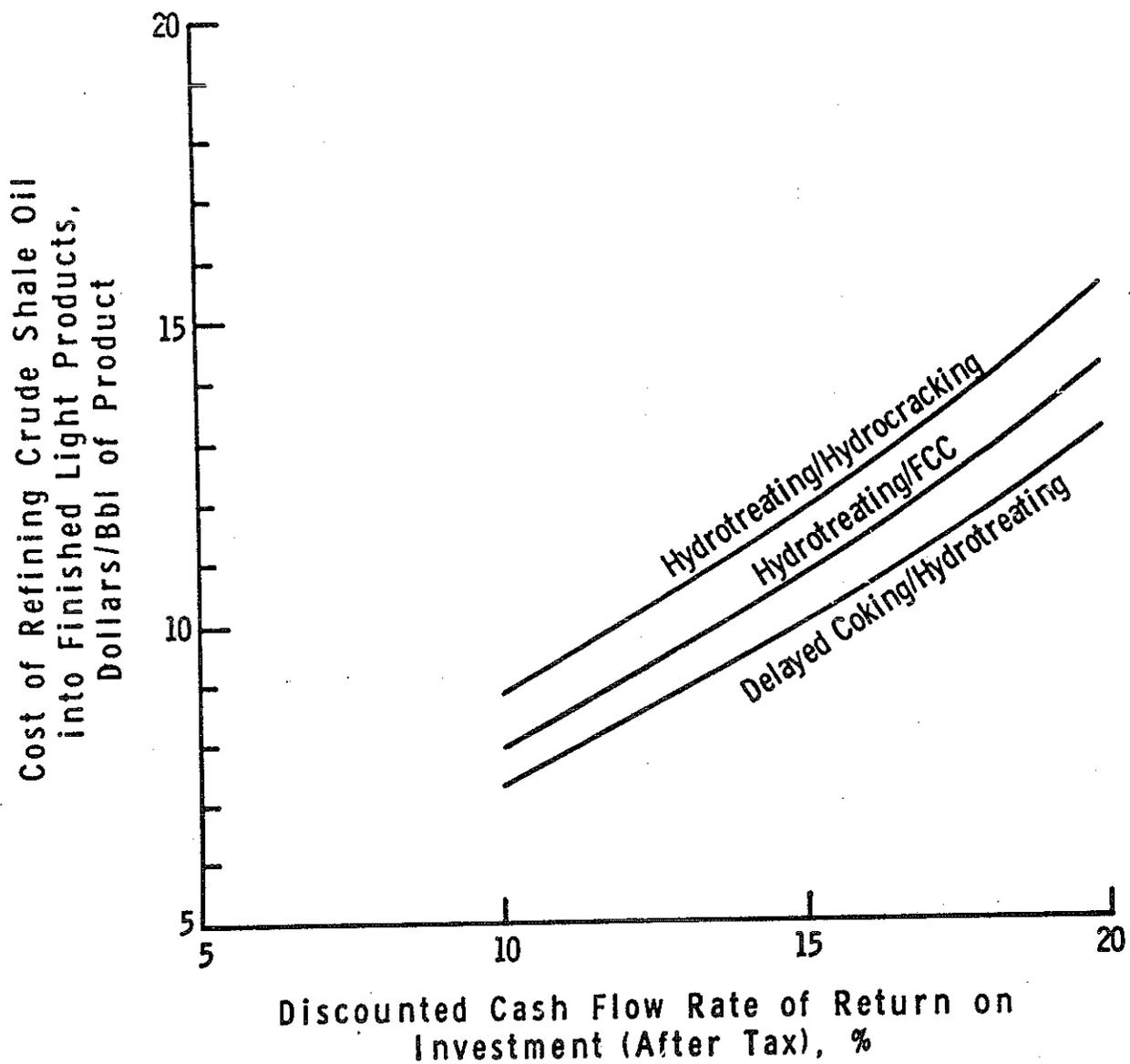


FIGURE 9  
EFFECT OF RATE OF RETURN ON PROCESSING COSTS FOR  
REFINING OF CRUDE PARAHO SHALE OIL INTO  
FINISHED LIGHT PRODUCTS

Grass Roots 100,000 BPCD Refinery

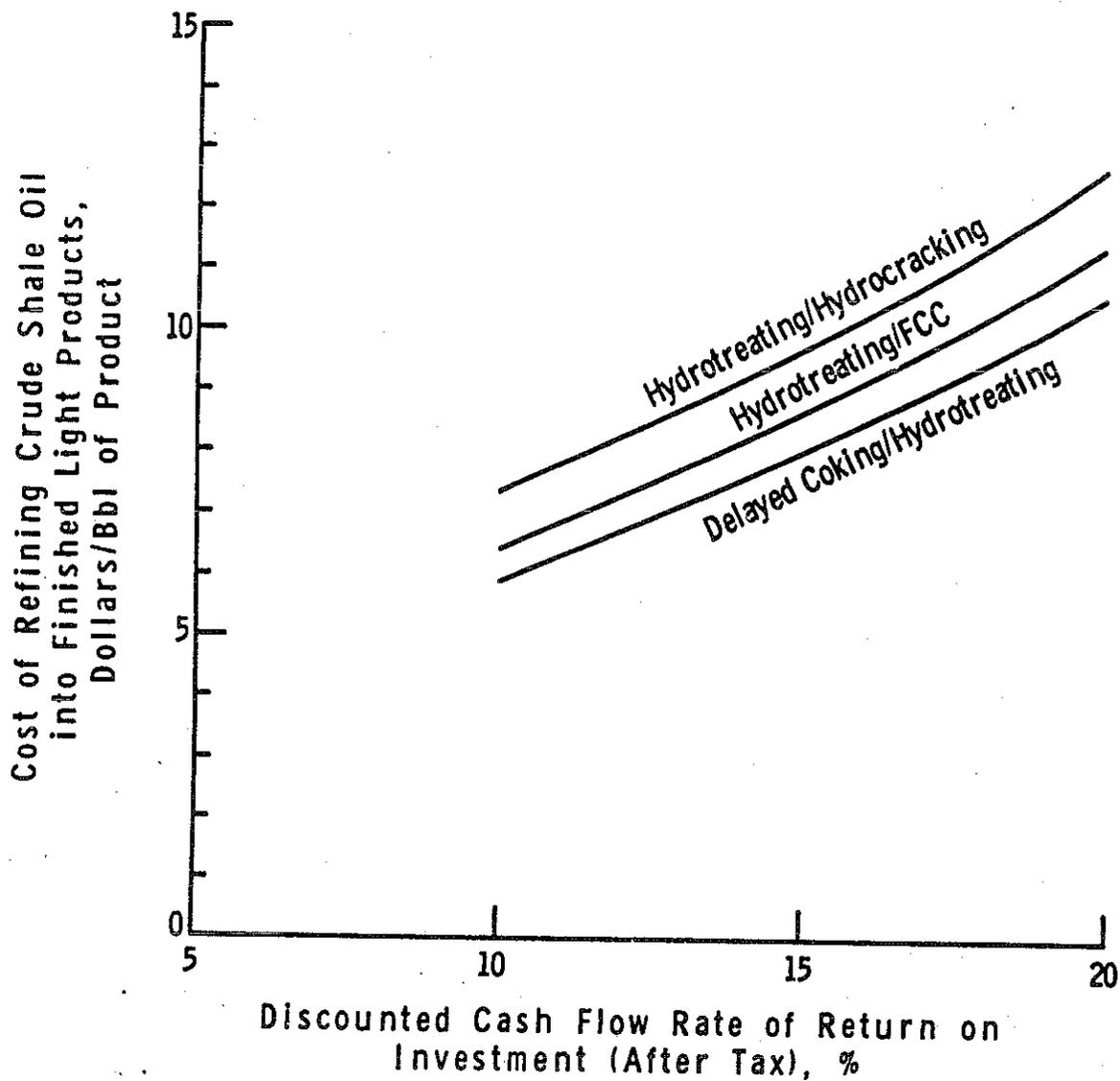
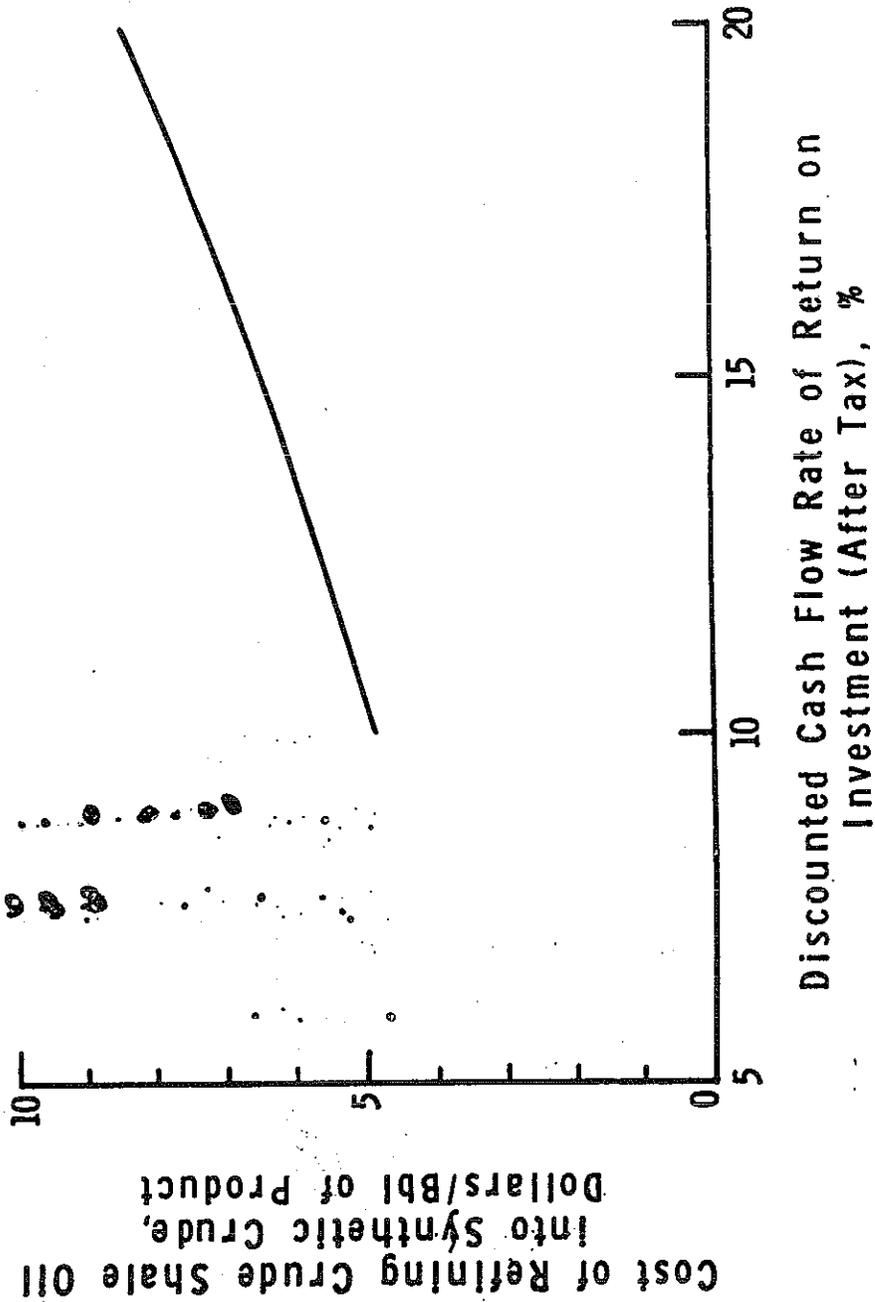


FIGURE 10  
 EFFECT OF RATE OF RETURN ON PROCESSING COSTS FOR  
 PRODUCING SYNTHETIC CRUDE FROM PARAH0 SHALE OIL  
 100,000 BPCD



\*U.S. GOVERNMENT PRINTING OFFICE: 1979-640-092 729.