Economic Enhancement of Western Shale Oil Upgrading

Topical Report

**J.W. Bungler
H. Ryu
S-Y. Jeong**

July 1989

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**For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia**

**By
Western Research Institute
Laramie, Wyoming
and
University of Utah
Salt Lake City, Utah**

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ECONOMIC ENHANCEMENT OF WESTERN SHALE OIL UPGRADING

by

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EXECUTIVE SUMMARY

A proof-of-concept study for a novel shale oil refining process was undertaken. This project promises reduced upgrading costs, thereby making shale oil development more feasible for commercialization. The process consists of distillation of raw shale oil into a distillate and residue portion, cracking of the residue by hydropyrolysis, and selective hydrotreating of narrow boiling cuts from the total distillate. Based on models and experimental data, the end product slate is projected to be 34% naphtha, 57% middle distillate, and 10.3% atm residue + coke. Hydrogen addition is 1.3% or 800 scf/bbl. These results are considerably improved over conventional processing, which gives 14% naphtha, 41% middle distillate, and 48.2% residue + coke and hydrogen addition of 3.2% or 2000 scf/bbl. More quantitative data and preliminary economics will be obtained in the next phase of study.

DEFINITION OF THE PROBLEM AND OBJECTIVES

Upgrading of raw shale oil to refinery-acceptable crude oil adds significantly to the cost of shale oil production. The high cost of upgrading contributes to the marginal economics for oil shale commercialization. An improvement in the efficiency of shale oil upgrading and a reduction in costs would help advance the development of oil shale in the United States.

Currently practiced upgrading approaches encounter several difficulties which must be addressed in any advanced upgrading process. Raw shale oil contains appreciable quantities of particulates and metals such as vanadium, nickel, iron, and arsenic. The presence of iron and

arsenic is particularly significant because catalysts used in conventional petroleum processing have not been designed for the presence of these metals.

Shale oil also contains high concentrations of nitrogen which must be removed catalytically before the shale oil is acceptable as a feedstock to a conventional petroleum refinery. Shale oil also contains large quantities of conjugated olefins, such as indenenes, indoles, and pyrroles, which are difficult to completely hydrogenate but which, when left in the shale oil, cause long-term instability of the shale-oil product.

Much of the difficulty with nitrogen removal relates to the catalytic hydroprocessing of the heavy ends of the shale oil. Atmospheric residue typically constitutes 50% or more of the raw shale oil. To effect nitrogen removal from this material, high-process severities are required which result in high hydrogen consumption and reduced catalyst life due to coking reactions. High severities result in poor control over product slates.

The literature contains numerous articles on shale oil upgrading and processing.¹⁻³ Various schemes are proposed for dealing with the inherent difficulties of shale oils. Some schemes involve prior coking of the raw shale oil to reduce the molecular weight and to sequester undesirable inorganics in the coke. While such schemes make the down-stream catalytic processing easier, coking results in a significant loss in yield; the advantages obtained are offset by the higher costs in the mining and retorting processes. Larger amounts of material must be handled to obtain a given output of product.

Upgrading and converting the entire raw shale oil distillation profile to high value marketable products has not yet been achieved. Most often, a significant portion of the final product is sold for fuel oil, because of the cost and difficulty in converting the residual end of shale oil to transportation fuels. At the present time conventional refineries will not accept a raw shale

oil and catalytic hydroprocessing must be effected prior to refining.

Shale oil from Colorado is currently being upgraded at the several-thousand bbl/day level by UNOCAL.⁴ The upgrading sequence consists of passing raw shale oil over a catalyst/guard bed to remove metals and particulates. The shale oil then passes over hydrotreating catalysts for nitrogen and sulfur removal and for hydrogenation of olefins. Mild hydrocracking of paraffins may be included for pour point control.

For sequences which do not involve hydrocracking, the hydrogen requirements for denitrogenation typically exceed 2,000 scf/bbl.⁵ For these approaches, less than 15% of the finished product boils in the naphtha range (80 to 200°C). This means that additional cracking and hydroprocessing is required at the refinery for production of gasoline and middle distillates (diesel or aviation turbine fuels). Thus, currently practiced technology results in high hydrogen consumption, low yields of products in the desirable boiling range, and less than desirable catalyst life.

Calculations show that complete removal of nitrogen, sulfur, and oxygen to form NH_3 , H_2S , and H_2O , respectively, and saturation of olefinic bonds requires approximately 800 scf/bbl. Additionally, ring hydrogenation is required for certain structures to allow removal of ring heteroatoms. Theoretical thermodynamic considerations suggest that high selectivity for HDN, HDS/HYD cannot be accomplished when operating at high process severities. Theoretical considerations have also shown that high selectivity is extremely difficult when processing higher molecular weight aromatics. Condensed aromatics of two rings or larger are more susceptible to hydrogenation than are single-ring aromatics.

The high rate of catalyst deactivation and the inability to project a cost effective refining and market option for the residual portion make the assessment of upgrading costs very difficult. There are three basic characteristics of raw shale oil which contribute to the high costs of upgrading and refining which must be addressed in any advanced scheme for processing these materials. The first of these is the metals and particulates content; the second of these is the extraordinarily high nitrogen content, much of which is pyrrolic in nature, and these types are not found in the same level of concentration in conventional petroleums; the third characteristic is the relatively large fraction of atmospheric resid in comparison to light and middle distillate. Ordinarily, the atmospheric resid or vacuum gas oil is charged to catalytic crackers, but the high nitrogen content of these high boiling materials makes the catalytic cracking option unacceptably expensive. An advanced upgrading and refining approach must account for these three critical factors.

An advanced shale oil upgrading process would involve cost-effective removal of inorganics and metals, accomplish selective utilization of hydrogen, and maximize the yield of low-to-middle boiling distillates. The upgrading process studied under this project promises to accomplish these objectives. The approach maximizes liquid yields and minimizes hydrogen consumption while meeting criteria for refinery acceptability. The process promises higher process efficiency and is expected to result in significantly reduced shale oil upgrading costs.

PROCESS DESCRIPTION

Hydropyrolysis is a process which has been under development at the University of Utah for more than a decade.^{6,9} The Utah hydropyrolysis process is a short-residence time (1-3 sec.) moderate temperature (500-600°C) process for the cracking of residual materials to distillate products. Feedstock is introduced to the reaction zone containing 1,200-2,000 psig H₂ by use of

a spray nozzle. Hydrogen participates in the reaction by metathesis reactions with hydrocarbon free radicals. Hydrogen atoms produced in the metathesis reactions accelerate cracking, dealkylation, cracking of naphthenic compounds, and partial hydrogenation of PNA's.

During hydropyrolysis, reduction of molecular weight occurs with virtually no production of coke. Coke-forming reactions are almost completely inhibited by the action of hydrogen and the dilute phase of the reaction regime.¹⁰⁻¹² Reduction of the molecular weight is achieved with very modest net hydrogen consumption. A reduction of average molecular weight by 50% is typically accompanied by a net hydrogen consumption of only 200-300 scf/bbl. Theoretical calculations show that this is very close to the minimum hydrogen requirements for one cracking event per molecule, assuming that there is no net dehydrogenation of naphthenes to aromatics.

An important feature of hydropyrolysis is its relative insensitivity to inorganics and metals. These species do not adversely affect the process and tend to report to the unreacted residual portion of the products. The process may be operated at low severities to minimize gas production and unreacted material can be recycled to improve the conversion. A small drag stream of the heaviest components containing the majority of the particulates and metals is removed from the reaction system. Thus, hydropyrolysis is a process which promises to convert heavy shale oil to distillate products in high yields and with low consumption of hydrogen.

An equally important discovery has been made in the catalytic processing research conducted. In a study to determine the effect of adsorption thermodynamics on catalytic reaction, it has been learned that the value of the adsorption constant K_{ads} is a strong function of molecular weight. Structural features, including the presence of most heteroatoms, played a secondary role in determining the magnitude of K_{ads} . In the first order reaction rate equation:

$$r = kK_a A / (1 + \sum K_i X_i).$$

The adsorption constant for non-reacting species appears as an inhibition term in the denominator. If $K_i \gg K_a$, then strong inhibition is expected. If a feedstock exhibiting a broad range of molecular weights is passed over a catalytic surface, the highest molecular weight species will successfully compete for adsorption on the surface, displacing lower molecular weight species and thereby inhibiting the opportunity for these lower molecular weight species to react.

This phenomenon is illustrated graphically in Figure 1. Figure 1a depicts a feedstock which has a broad molecular weight range with an average of 300 amu. When this material is passed over a catalyst surface, under steady state conditions, the molecular weight distribution of the adsorbed species is shown by the "dashed" line. At a temperature of 350°C, all but the highest molecular weight species are effectively excluded from the surface. An increase in temperature to 410°C improves the situation, but it is still clear that a large fraction of the feedstock is inhibited from adsorption. Figure 1b illustrates the case in which a narrower molecular weight range material is passed over the catalyst. In this case, a much larger fraction of the feedstock participates in the adsorption process and thereby is afforded an opportunity to react.

The implications of these findings to shale oil processing are obvious. If shale oil upgrading is to be achieved with high selectivity and acceptable conversions at moderate process severities, the feedstock to the catalytic process must be of a narrow molecular weight range. A narrow molecular weight range is most conveniently achieved through distillation into narrow boiling cuts (here it is recognized that the effect of structural differences in molecules will limit how narrow the range can be). The practical limit in how narrow the fraction should be and how many parallel hydrotreaters are required is, in fact, an economic question which would constitute a significant portion of the second phase of the program.

The SOR sequence is illustrated in the flow diagram shown in Figure 2. The initial step in the process is a fractionation of raw shale oil into an atmospheric distillate and an atmospheric resid. The resid, which contains the vast majority of the inorganics, is fed to the hydropyrolysis reactor designed to operate at moderate, e.g., 70%, once-through conversions. Unreacted material is recycled to improve the conversions. By operating at modest severities, high conversions to distillable liquids with low hydrogen consumption is achieved.

The distillate material, boiling in the range of 80-350°C is recombined with the overhead in the original distillation step. This material is fractionated in a high efficiency column to produce "n" fractions for hydrotreating. Hydrotreating is conducted in parallel reactors under conditions specifically tailored for the distillate cut. The hydrotreaters may or may not contain guard beds, depending upon the metals content of the fuel.

RESULTS AND DISCUSSION

Theoretical Development for Catalytic Process Model

Conversion in catalytic processing will be a function of four variables: (1) α^* , (2) (a^*-a) , (3) $a^\dagger$, and (4) m_o . These variables are defined in the following equations:

$$K_{abs} = K^* e^{a^* m}$$

$$K_{rate} = K_o e^{-a m}$$

$$MW \text{ distribution} = e^{-a^\dagger (m - m_o)^2}$$

where in each of these, m = molecular weight. Also,

$$\alpha^* = \frac{A S L K_o^*}{M}$$

where: A = cross-section area of reactor;
 L = length of reactor;
 S = number of active sites on the catalyst per cc of reactor; and
 M = hydrocarbon flow rate, moles/sec.

Since a conversion for a set of the four variables must be obtained by computer numerical integration, much time will be needed to fill out a table of conversion over appropriate ranges of the four variables. Then an empirical fit to that set of numbers would give the desired general equation. So the equation can be obtained, but it will take a good deal of computer analysis.

In our mathematical model for hydrotreating a given feed over a given catalyst, we let the equilibrium absorption constant, K_a , be a function only of molecular weight, m , and the reaction rate constant for absorbed species, K_r , is likewise just a function of m :

$$K_a = k^* a^* m$$

$$K_r = K_o e^{-am}$$

In our experimental work, we will know the molecular weight distribution of the feed, a distillate cut from the shale oil. So in our mathematical model, the conversion will be a function only of two as yet unknown parameters, (a^*-a) and α^* as defined above.

Note that the unknown part of α^* is:

$$\alpha_o^* = S K_o K^*$$

Thus we have two unknown parameters, (a^*-a) and α_o^* , and numerous data points of known conversion for known feed and residence time. We have worked out an effective strategy to obtain (a^*-a) and α_o^* through a least-squares type analysis of this data. Such a result will now allow us to predict conversion for any distillate cut from the shale oil reacting over the given catalyst in any reactor configuration and for any residence time. We note that we could improve the prediction, if needed, by letting K_a and K_r be functions of z and i as well as m (see Z-BASIC method).¹³

Shale Oil Characteristics

The shale oil characteristics are shown in Table 1. Samples of shale oil were obtained from the Western Research Institute (WRI) and the New Paraho Shale Oil Corporation. The samples have been subjected to elemental analysis and certain physical properties measurements. The data in Table 1 show that the Paraho shale oil has a slightly higher density and viscosity and lower volatility. In selecting a sample for use, consideration was given to the two principal experimental tasks of the program, namely, the hydropyrolysis and hydrotreating. The Paraho shale oil is a better test sample for the hydropyrolysis because of its higher residue content. The WRI sample is a better test sample for the hydrotreating portion because of its higher yield of distillate. Ultimately, the value of the concept is expected to be greatest for those processes which provide a high yield of shale oil based on Fischer assay, but which also may contain higher percentages of atmospheric resid. The data shown in Table 1 are typical for shale oils, and reveal the large percentage of material boiling in the atmospheric and vacuum distillate and residue range. Both of these samples represent good cases for study for the proof-of-concept of the shale oil refining process.

Distillation

The WRI shale oil was subjected to thin-film distillation for purposes of topping the shale and removing the most volatile components. These components are easily hydrotreatable, but their presence in the samples results in increasing difficulties with material balances and their removal will improve the final results. The fractional distillation compositions of the as-received shale oil and the topped shale oil are shown in Table 2. Approximately 12.4 percent of the shale oil was distilled and condensed. The light end condensate exhibited an API gravity of 40.8°. The topped shale oil, representing nearly 16 kg of material and 87.6 percent of the original material exhibited an API gravity of 22°. The simulated distillation data for these two samples is shown in Table 2.

The degassed and topped WRI shale oil was further distilled by semi-cal fractional distillation to obtain three narrow fractions and a residue. The three fractions are representative of the 200-275, 275-365, and 365-455°C boiling ranges. The efficiency of the separation was determined by simulated distillation analysis of the three fractions. These results are shown in Table 3.

PROCESS SIMULATION developed by Simulation Sciences, Inc., was used to calculate the important physical and thermodynamic properties of these three fractions based on a pseudo-component concept. These results, taken at the ambient temperature case and the 400°C case, are shown in Table 4. These distillates were used for the hydrotreating studies. Also used for the hydrotreating studies was a blended sample of the three distillates blended in equal weight proportions.

Catalyst Characteristics

For purposes of the proof of concept studies, the particular selection of catalyst is not critical. However, due to the design and optimization tasks to be conducted in the second year of the study, it was important that a meaningful catalyst be selected because the results could ultimately form the baseline for comparison with other candidate catalysts.

Two catalysts were selected and obtained from Shell Development Company. The properties of these two 1/16 inch catalysts are listed in Table 5. The data in the table show that the chemical composition is nearly identical with the 344 catalyst possessing a lower specific surface area compared to the 444 catalyst. Catalysts of lower specific surface area typically exhibit higher average pore diameters. For purposes of this study, it was desirable to limit the diffusion effects as much as possible so that the conversions and rates measured are more representative of the intrinsic catalytic reaction rates. Therefore, the 344 catalyst was selected for initial study.

Hydropyrolysis

Hydropyrolysis is a short residence time, non-catalytic thermal hydrocracking process for reducing the molecular weight of residual components without the formation of coke. Hydropyrolysis may ideally crack atmospheric and vacuum resid to atmospheric distillate, rendering the nitrogen and sulfur compounds amenable to conventional catalytic hydrotreating. Experimental difficulties with the hydropyrolysis unit and delays in the generation of distillate samples by the high efficiency distillation have prevented hydropyrolysis processing of shale oil resid to be completed in the present reporting period.

In order to obtain data and an estimate of the hydropyrolysis yields for shale oil, characterization data for the shale oil resid were used as input data for our kinetic model of the

hydropyrolysis process to predict the yield distribution as a function of temperature and residence time.¹³

The complex mixture is separated into specified boiling ranges to provide individual products of commercial importance such as residue (R), atmospheric and vacuum distillate (AVD), naphtha and middle distillate (NMD), and gas (G). The knowledge that molecular weight is a strong function of its boiling point and material of high molecular weight cracks to lower molecular weight species suggests the following relatively simple scheme for representing the overall kinetics:

where R is the weight fraction of residue exhibiting a nominal boiling point above 538°C, AVD is the weight fraction of atmospheric and vacuum distillate with a nominal boiling point of 275-538°C, NMD is the weight fraction of naphtha and middle distillate boiling at 1 atm up to 275°C (C₅-275°C), and G represents the weight fraction of gases (C₄'s, dry gas).

The predicted yield pattern is shown in Figure 3 A-D and revealed that at high severity of temperature and residence time, high conversions of residue to light distillate and gas is predicted. At more moderate (and probably more optimum) conditions, approximating 525°C and 5 seconds residence time, the predicted yield structure is 8 percent gas, 38 percent naphtha plus middle distillate, 18 percent atmospheric and vacuum distillate, and 18 percent residue. These compositions represent the outlet composition at nominal ambient conditions and includes recycle gas. The unreacted heavy materials and gases at the reactor outlet will be continuously recycled. The concentration of each lump at zero time was taken to be equal to the concentration of reactor inlet.

With regard to the residue, it may be fairly noted that the process model is based on tar sand bitumen which possesses typically over 50 percent vacuum residue and is expected to be considerably less tractable to hydrolysis than would be a shale oil resid. Thus, it is expected that actual hydrolysis data will show considerably greater conversions of the residue than is predicted by the currently available model. Properly configured, it is thought possible to recycle the residue portion to extinction, thereby dramatically improving the yield of distillate in comparison to that contained in the raw shale oil.

Catalytic Hydrotreating

The equipment used in this study is a trickle bed reactor charged with about 50 g of catalyst. Process conditions were selected to provide mild conversions, thereby providing high sensitivity to changes in feedstock composition and process conditions. Process conditions used for the line-out studies were:

$$T = 375^{\circ}\text{C};$$

$$P = 1500 \text{ psig } \text{H}_2;$$

$$\text{WHSV} = 0.86/\text{hr};$$

$$\text{H}_2 \text{ flow} = 3000 \text{ scf/bbl}; \text{ and}$$

$$\text{catalyst} = \text{Shell 344 T/L 1/16}.$$

Under these conditions, approximately 70% HDN is achieved.

Experimental results of processing of distillates are too preliminary to report at this time. Calculations suggest that by processing narrow boiling distillates under identical conditions that conversions of 85% would be achieved. However, the presence of the residual "tail" in the whole shale oil tends to make the calculations appear worse than would actually be experienced.

Quantitative data will be obtained in the early stages of the next phase.

Discussion of Conventional and SOR Results

Calculated results from the SOR process are compared with conventional upgrading results as shown in Table 6. As can be seen, a major difference in product slate is calculated and lower hydrogen consumption is expected. These results clearly demonstrate the major differences expected from the refining approach. Further experimental verification of these projected results will be conducted during the next phase of study.

SECOND PHASE WORK

The second phase of this research will consist of developing more quantitative data required for preliminary optimization of the process. Experimental work will be conducted to systematically determine important pre-design data for engineering purposes. The tasks will involve the development of the reaction engineering model which will include the effect of process variables on yield, product composition, hydrogen consumption, and level of hydrodenitrogenation. This model will include estimates of the effect of operating conditions such as residence times, catalyst to oil ratios, temperatures, throughput rates, distillation conditions, and pressure. The reaction engineering model will form the basis for a process simulation and optimization exercise.

Patentable innovations are anticipated and will be disclosed to the Office of Technology Transfer at the University of Utah, which will make the technology available for license for commercial development. Subsequent activities (Phase III and beyond) will be required to complete engineering design data, to test the general applicability to shale oils, and to ascertain preferred catalyst compositions and structures. Phase III funds will be sought for these purposes, both through the Western Shale Oil Consortium and from other government and private sources.

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Table 1
*Elemental Analysis and Physical Properties
of Raw Shale Oil*

	Paraho	WRI
<hr/>		
<u>Element</u>		
C	82.02	83.08
H	11.07	11.67
N	1.87	1.47
S	0.71	1.036
O (diff)	4.34	2.74
C/H	0.617	0.593
<u>Property</u>		
Sp. Gr. 60/60°F	0.9330	0.9155
API	20.16	23.06
Viscosity (cp @ 25°C)	540	45
 Simulated Distillation Data (wt.%)		
< 305°C	22.3	35.5
305-350°C	11.5	15.2
350-538°C	49.6	45.8
> 538°C	<u>16.6</u>	<u>3.5</u>
	100	100

Table 2

*Boiling Point Distribution of the WRI
As Received and Topped Shale Oil*

	As Received Shale Oil (wt.%)	Topped Shale Oil (wt.%)
Gasoline (<200°C)	2.6	0.2
Kerosene (200-275°C)	16.1	11.5
Gas Oil (275-325°C)	15.8	17.0
Hvy GO (325-400°C)	27.8	30.9
VGO (400-538°C)	32.5	33.0
Vac. Residue (>538°C)	5.2	7.4
Total	100	100

Table 3

*Boiling Point Distribution of Fractional
Distillation Cuts from WRI Shale Oil*

Fraction	1 (200-275°C)	2 (275-365°C)	3 (365-455°C)
Gasoline (<200°C)	11.4 wt. %	--	--
Kerosene (200-275°C)	86.4	4.6	--
Gas Oil (275-325°C)	2.2	41.0	--
Heavy Gas Oil (325-400°C)	--	54.4	12.8
Vacuum Gas Oil (400-538°C)	--	--	87.2
Total	100	100	100
Fractionation Efficiency (%)	97.8	95.4	94.5

Table 4
Physical and Thermodynamic Properties of Distillate Fraction @ 20 and 400°C

Stream ID	C1A	C1B	C2A	C2B	C3A	C3B
Stream Phase	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Mole Frac Liquid	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature, °C	20.0000	400.0000	20.0000	400.0000	20.0000	400.0000
Pressure, psia	14.7000	1500.0000	14.7000	1500.0000	14.7000	1500.0000
Enthalpy, mm Btu/Hr	-0.0003	0.0087	-0.0002	0.0084	-0.0003	0.0080
Entropy, Btu/lb C	0.3874	1.2701	0.3054	1.1613	0.1956	1.0120
Act. Dens lb/ft ³	56.0335	35.3676	58.2766	40.5101	57.4169	44.1880
Crit. Temp, °C	412.3242	412.3242	490.5108	490.5108	583.9207	583.9207
Crit. Press., psia	319.4544	319.4544	255.5638	255.5638	191.2883	191.2883
Crit. Volume, ft ³ /lb mole	10.4114	10.4114	13.9730	13.9730	20.0213	20.0213
Crit. Compress. (°C)	0.2491	0.2491	0.2414	0.2414	0.2302	0.2302
Reduced Temperature	0.4277	0.9820	0.3839	0.8815	0.3420	0.7854
Reduced Pressure	0.0460	4.6955	0.0575	5.8694	0.0768	7.8416
Acentric Factor	0.5239	0.5239	0.6205	0.6205	0.7596	0.7596
Std. Spgr, (H ₂ O, 60°F)	0.8199	0.8199	0.8702	0.8702	0.9323	0.9383
Std. API Gravity	41.0727	41.0727	31.1093	31.1093	20.2780	20.2780
UOP K (mole basis)	11.7803	11.7803	11.6323	11.6323	11.4497	11.4497
Molecular Weight	178.6985	178.6985	241.1974	241.1974	333.4063	333.4063
NML. Boil Pt. (Mole), C	225.3356	225.3356	302.0174	302.0174	401.1645	401.1645
Flash Point, °C	66.8794	66.8794	120.5196	120.5196	178.7225	178.7225
CP, Btu/lb C	0.8084	1.4026	0.7628	1.3736	0.7179	1.3802
CP/(CP-4) Ratio	1.0254	1.0145	1.0198	1.0109	1.0152	1.0078
Thermal Conductivity Btu/Hr. Ft.F	0.0679	0.0779	0.0639	0.0549	0.0605	0.0401
Viscosity, cp	2.0372	0.1138	7.3194	0.1596	133.6496	0.2733
Surf. Tens., Dynes/cm	29.1007	0.8666	32.2362	4.4358	35.4442	9.1950

1 = 200-275; 2 = 275-365; 3 = 365-455

Table 5
Catalyst Properties

Catalyst	344 T/L 1/16	444 T/L 1/16
Nominal size, in (mm)	1/16 (1.5)	1/16 (1.5)
Chemical composition, wt.%		
Cobalt	2.52	3.22
Molybdenum	9.85	9.59
Physical properties		
Surface area, m ² /g	190	230
Pore volume, cc/g	0.56	0.57
Side plate crush strength, lb	22.5	19.4
Attrition index	98.8	98.8
Compacted bulk density, lbs/ft ³	45.0	45.4
Loss on ignition at 900°F, wt.%	1.24	0.99

Table 6
Projected Results

(percent)	Feed	Conventional Upgrading	Proposed Upgrading
<i>Liquids</i>			
60-200°C	12	14	34
200-350°C	38	41	57
200-350°C	24	24	7
350-425°C	26	24	2
Coke	--	0.2	1.3
	—	—	—
	100	103.2	101.3
<i>H Consumption</i>			
wt%	--	3.2	1.3
scf/bbl (feed)	--	2000	800

(a) Broad M.W. Range

(b) Narrow M.W. Range

Figure 1

Composition of Feed Streams and Adsorbed Species

Figure 2
Shale Oil Upgrading Process
Flow Diagram and Material Balance Projected

a) Gas

b) Naphtha+Middle Distillate

c) Atmospheric and Vacuum Distillate

d) Residue

Figure 3

Yield of Products from Hydropyrolysis of Shale Oil as a Function of Time and Temperature (Modeled Results)