

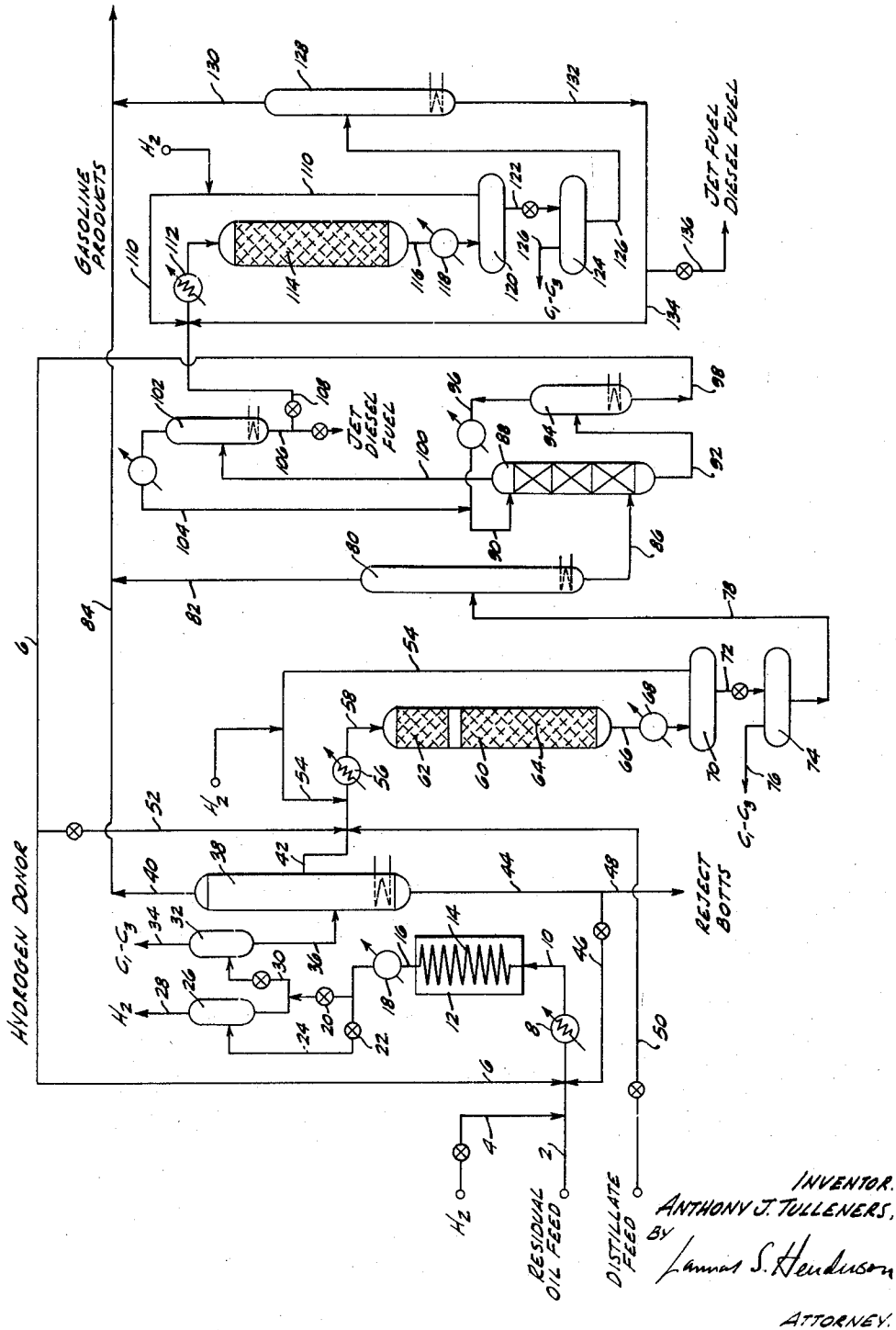
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HYDROCRACKING PROCESS WITH THE USE OF A HYDROGEN DONOR

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This invention relates to catalytic hydrocracking, and more particularly is concerned with methods for hydrocracking residual oil feedstocks, i.e., undistilled crude oil residua boiling above about 800° F. In broad aspect, the process consists in first subjecting the residual oil feed to a non-catalytic hydrogen donor treatment at elevated temperatures, using as the hydrogen donor a high boiling fraction of the subsequent hydrocracker effluent, to thereby effect hydrogen transfer and boiling point reduction of the residual oil feed components, and then subjecting a selected gas oil fraction of the product from the hydrogen donor treatment to catalytic hydrocracking. The remaining heavy material from the hydrogen donor treatment, boiling above the gas oil fraction which is subjected to hydrocracking, may then be recycled at least in part to the hydrogen donor treatment along with fresh feed. The effluent from the hydrocracking operation is fractionated to recover a desired gasoline fraction, and remaining gas oil boiling above the gasoline range may be recycled to the hydrogen donor treatment as such, or it may be subjected to a separation step, as by solvent extraction, to recover a more concentrated aromatic hydrogen donor fraction which is recycled to the hydrogen donor treatment.

Where the solvent extraction step is employed, the raffinate therefrom is highly saturated in character, and may be utilized as jet fuel or diesel fuel after suitable fractionation. Or, in cases where further conversion to gasoline is desired, part or all of the raffinate may be subjected to a second stage of catalytic hydrocracking, advantageously under relatively mild conditions which are particularly adapted for the hydrocracking of non-aromatic feedstocks.

According to another embodiment of the invention, a distillate mineral oil feedstock may be used in conjunction with the residual oil feedstock, the former being admitted directly to the initial hydrocracking step in admixture with the gas oil fraction derived from the hydrogen donor treatment. According to a still further and preferred embodiment of the invention, the initial hydrocracking step is preceded by an "integral" hydrofining operation, wherein the hydrocracker feedstock is first contacted with a relatively non-cracking hydrofining catalyst in the presence of hydrogen to effect partial saturation and conversion of sulfur and nitrogen compounds. The entire effluent from the hydrofining step is then transferred directly to the initial hydrocracking step without intervening condensation or purification.

The invention is based essentially upon my discovery that the hydrocracked product from an initial stage of hydrocracking of an aromatic feedstock contains a high proportion of very desirable types of hydrogen donor compounds, i.e., polycyclic condensed-ring hydrocarbons containing at least one benzene ring and at least one six-membered naphthenic ring. A prime example of such hydrocarbons are those of the tetralin type, and alkylated derivatives thereof. It has further been found that an even more concentrated fraction of these desirable hydrogen donor type compounds can be isolated by solvent

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extraction of the hydrocracking effluent boiling above the gasoline range. This extract from the hydrocracked gas oil can hence be used very effectively for the upgrading of residual oil feedstocks, so that a substantial portion of the refractory components in such residual oil feedstocks can be partially hydrogenated by non-catalytic hydrogen transfer, and thermally cracked simultaneously to produce suitable feed for hydrocracking. During the subsequent hydrocracking, a portion of the dehydrogenated donor compounds is re-hydrogenated, and a portion is hydrocracked to form monocyclic hydrocarbons in the gasoline range. But at the same time, more donor compounds are synthesized from the residual oil feed components, so that a desired donor balance can be maintained in the process.

In its present state of development, catalytic hydrocracking has been adapted almost exclusively for the conversion of distillate mineral oil feeds to gasoline boiling range hydrocarbons. Such distillate feeds may be either straight-run gas oils or cracked gas oil fractions derived from catalytic cracking, coking or the like. However, a considerable proportion of the total crude oils being refined appears ultimately as undistillable residua which are ordinarily diverted to low cost fuel oils, or converted to coke and asphalt. Such residua have never been considered practical feedstocks for hydrocracking in that they contain metal contaminants and extremely heavy, condensed-ring aromatic hydrocarbons of the asphaltene and carboid types, all of which rapidly deactivate hydrocracking catalysts. Economical methods for converting such residua to practical hydrocracking feedstocks are hence highly to be desired as a means of increasing the proportion of crude oil ultimately convertible to more valuable products of the gasoline-jet fuel-diesel fuel type.

For a more detailed description of the invention, reference is made to the accompanying drawing, which is a flow sheet illustrating several specific modifications of the process. The initial residual oil feedstock is brought in through feed line 2, mixed with hydrogen, if desired, from line 4, and blended with about 0.1 to 4 volumes of recycle hydrogen donor fraction from line 6, and the entire mixture is preheated to the desired hydrogen donor processing temperature in preheater 8. The use of hydrogen during the donor processing step is optional, and may be entirely omitted if desired. Suitable donor treatment temperatures may range between about 650–950° F., and preferably between about 750–850° F. The preheated mixture of feed and hydrogen donor is transferred via line 10 to hydrogen donor treating unit 12, which may be an insulated case containing an elongated, tubular reaction coil 14. The hydrogen donor treatment is normally carried out at pressures of about 25–1,500 p.s.i.g., preferably about 200–750 p.s.i.g. The liquid hourly space velocity may range between about 0.2 and 10 volumes of combined liquid feed per volume of reactor space per hour. Preferably the liquid hourly space velocity is about 0.5 to 3. Any other type of conventional hydrogen donor processing equipment and conditions may be employed in lieu of the specific modification illustrated.

The effluent from donor processing unit 12 is withdrawn via line 16, and cooled to about 50–200° F. in heat exchanger 18, and the cooled effluent is then treated in either of two alternative modes, depending upon whether hydrogen was used in the processing. If hydrogen was used, valve 20 is closed and valve 22 opened, thereby diverting the mixture, still substantially at reactor pres-

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sures, through line 24 into high pressure separator 26, from which the hydrogen is removed via line 28, and may be recycled to the unit. The liquid product in high pressure separator 26 is withdrawn via line 30 and flashed into low pressure separator 32, from which light hydrocarbon gases are exhausted via line 34. The liquid phase in separator 32 is then withdrawn via line 36 and transferred to fractionating column 38. If hydrogen is not employed in the donor processing, high pressure separator 26 may be omitted, as by closing valve 22 and opening valve 20, whereby the effluent flows directly through line 30 into low pressure separator 32, the liquid phase in separator 32 then being transferred as before to fractionating column 38 via line 36.

In column 38, gasoline synthesized in donor treating unit 12 is withdrawn as overhead via line 40, a gas oil side-cut is withdrawn via line 42, constituting hydrocracker feedstock, and a heavy bottoms fraction boiling above about 750° F. is withdrawn via line 44. This bottoms fraction, containing most of the metallic impurities in the initial feed, and a high proportion of refractory polycyclic aromatic hydrocarbons may be entirely withdrawn from the process, or if desired a substantial portion thereof may be recycled via line 46 for further hydrogen donor treatment in admixture with the fresh feed. Ordinarily it is not feasible to recycle all of this material because of the resultant buildup of metallic impurities and highly refractory materials. However, as an alternative to the partial recycle shown (with a bleed stream being taken off via line 48), the entire bottoms product in line 44 may be subjected to a conventional deasphalting step using propane or other light hydrocarbons, and the asphaltic raffinate withdrawn from the process while the extract is recycled in its entirety to donor processing unit 12 via line 46. This deasphalting procedure effects a concentration of metallic impurities in the asphaltic residue which is withdrawn from the process.

The primary hydrocracking feedstock in line 42 normally comprises a gas oil fraction boiling between about 350–900° F., and may be utilized as such, or may be blended with certain auxiliary feedstocks or recycle streams from the process. In particular, it is contemplated that this primary feedstock may be blended with an extraneous distillate feedstock from line 50, and may also be blended with a portion of the hydrogen donor material which is being recycled in line 6, a portion thereof being diverted to line 42 via line 52. It will be understood that the hydrogen donor fraction itself constitutes excellent hydrocracking feedstock, and the desirability of diverting some of it directly via line 52 depends upon the ultimate donor balance in the process. If more hydrogen donor material is recovered from the hydrocracking step than is necessary to upgrade the primary residual oil feedstock, the excess is preferably recycled directly to the hydrocracker via line 52. In cases where an auxiliary distillate feed is employed via line 50, there will ordinarily be more hydrogen donor fraction recovered than is necessary for upgrading the residual feedstock. Suitable auxiliary feedstocks may comprise for example straight-run gas oils, recycle oils from catalytic cracking units, coker distillate gas oils or the like. Any of these auxiliary feedstocks will also produce during the subsequent hydrocracking step a considerable proportion of valuable hydrogen donor compounds.

The combined hydrocracking feedstock in line 42 is then blended with recycle and fresh makeup hydrogen admitted via line 54, and the mixture is then preheated to suitable hydrofining or hydrocracking temperatures in preheater 56, and the preheated mixture is then transferred via line 58 to hydrocracking unit 60. In the modification illustrated, hydrocracking unit 60 comprises an upper bed of hydrofining catalyst 62, and a lower bed of hydrocracking catalyst 64. However, in cases where the combined feedstock in line 58 is sufficiently low in nitrogen compounds, it is feasible to omit the hydrofining

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catalyst, and fill the entire reactor 60 with hydrocracking catalyst. Suitable hydrofining conditions for processing in catalyst bed 62 are as follows:

TABLE 1
Hydrofining Conditions

	Operative	Preferred
Temperature, ° F.....	600–850	650–750
Pressure, p.s.i.g.....	500–3,000	800–2,000
LHSV, v./v./hr.....	0.5–10	1–5
H ₂ /oil ratio, s.c.f./b.....	500–20,000	1,000–10,000

Suitable processing conditions for use in lower hydrocracking bed 64 are as follows:

TABLE 2
First-Stage Hydrocracking Conditions

	Operative	Preferred
Temperature, ° F.....	600–850	600–800
Pressure, p.s.i.g.....	400–2,500	800–2,000
LHSV, v./v./hr.....	0.5–10	1–5
H ₂ /oil ratio, s.c.f./b.....	500–20,000	1,000–10,000

Those skilled in the art will readily understand that when ranges of operating conditions are specified as above, a large number of determinative factors are involved. Thus, highly active catalysts, or fresh catalysts at the beginning of a run, will be used in conjunction with lower temperatures than will less active or partially deactivated catalysts. The lower limit of pressure to be utilized in a given operation will normally depend upon the desired run length. Lower pressures generally result in a more rapid deactivation of the catalyst, and hence where extremely long run lengths are desired, pressures of above about 1,000 p.s.i.g. are mandatory. However, economically feasible run lengths are normally obtainable with most catalysts and feedstocks within the 600–2,000 p.s.i.g. pressure range.

It is further to be noted in reference to hydrocracking conditions, that it is desirable to correlate such conditions so that about 20–60% conversion to gasoline per pass will be obtained. At higher conversion levels, the proportion of desirable hydrogen donor compounds tends to be reduced because of conversion to gasoline, e.g., to monocyclic alkyl aromatic compounds. This is brought about mainly by the cracking of the naphthene rings in tetralin-type compounds. It is of course necessary to obtain a sufficient volume of hydrogen donor fraction to treat effectively the raw residual oil feed. Moreover, to obtain the optimum type of hydrogen donor compounds, it is desirable to use the relatively low hydrocracking pressures above specified, since higher pressures would tend to produce completely saturated hydrocarbons such as decalin, which are relatively ineffective as hydrogen donors.

The effluent from hydrocracker 60 is withdrawn via line 66, cooled in condenser 68 and transferred to high pressure separator 70, from which recycle hydrogen is withdrawn via line 54. The liquid condensate in separator 70 is transferred via line 72 to low pressure separator 74, from which light hydrocarbon gases are exhausted via line 76. The liquid phase in separator 74 is withdrawn via line 78 and transferred to fractionating column 80, from which overhead gasoline product vapors are withdrawn via line 82, and if desired blended with the gasoline from lines 40 and 84. The bottoms from fractionating column 80 constitutes a mixture of highly saturated and partially saturated hydrocarbons boiling above the gasoline range, and if desired a portion of this fraction may be recycled as such to the donor treating unit 12, the remainder being recycled to hydrocracking unit 60.

Preferably however, the bottoms fraction from column 80 is separated into a relatively aromatic and a relatively non-aromatic fraction, the former constituting a preferred hydrogen donor fraction for recycle, and the latter constituting a high quality jet fuel-diesel fuel fraction which may be used as such or subjected to a second stage of hydrocracking for further conversion to gasoline. Any conventional method for separating aromatic from non-aromatic hydrocarbons may be employed. Suitable methods include for example azeotropic distillation, selective adsorption, extractive distillation and the like. In the modification illustrated, the bottoms fraction from column 80 is transferred via line 86 to the bottom of a counter-current solvent extraction column 88, which is preferably packed with a suitable material such as Raschig rings or the like to facilitate contact between countercurrently flowing immiscible liquids. The solvent employed for the extraction may comprise any of the well known polar compounds which exhibit a selective solvency for aromatic hydrocarbons as opposed to non-aromatic hydrocarbons. Suitable solvents include for example phenols, aniline, nitrobenzene, benzyl alcohol, furfural, glycerol mono ethers, ethylene glycol, glycol monoethers, aliphatic dinitriles, benzonitrile, acetonitrile, sulfur dioxide, and the like. Conditions for the countercurrent extraction of aromatic hydrocarbons from non-aromatics are well known in the art and hence need not be described in detail.

In the solvent extraction step illustrated, the solvent is admitted at the top of the column via line 90 and passes downwardly countercurrently to the rising hydrocarbon stream. The aromatic extract is withdrawn at the bottom of the column via line 92, and transferred to fractionating column 94, from which the volatile solvent is removed as overhead via line 96, and recycled to extraction column 88. The stripped aromatic fraction, comprising the preferred hydrogen donor hydrocarbons is then taken off via line 98 and recycled via line 6 as previously described. The raffinate from extraction column 88, comprising non-aromatic hydrocarbons containing a small amount of dissolved solvent, is withdrawn via line 100, and sent to a raffinate stripping column 102 from which solvent is recovered overhead via line 104 and recycled to line 90 for reuse in extraction column 88. The bottoms from stripping column 102 is withdrawn via line 106, and if further conversion to gasoline is not desired, may be sent to further fractionating equipment not shown to recover high-quality jet fuel and/or diesel fuel therefrom.

However, in cases where further conversion to gasoline is desired, the bottoms fraction from column 106 is transferred via line 108, admixed with recycle and fresh hydrogen from line 110, preheated to hydrocracking temperatures in preheater 112, and sent to second-stage hydrocracker 114. Due to the substantial absence of aromatic hydrocarbons in the feed to hydrocracker 114, substantially milder hydrocracking conditions can be employed, as indicated in the following table.

TABLE 3

Second-Stage Hydrocracking Conditions

	Operative	Preferred
Temperature, ° F.	400-850	450-700
Pressure, p.s.i.g.	400-2,500	800-2,000
LHSV, v./v./hr.	0.5-15	1-10
H ₂ /oil ratio, s.c.f./b.	500-20,000	2,000-12,000

The conditions set forth above can be so correlated as to obtain a relatively high crack per pass if desired, e.g., about 50-90%, if the concomitant production of light hydrocarbons in the C₁-C₄ range can be tolerated. Alternatively, where light gas make is to be minimized, a lower crack per pass may be more desirable, i.e., in the

30-60% range ("crack per pass" here refers to volume percent conversion to 400° F. end-point gasoline per pass).

The effluent from second-stage hydrocracker 114 is withdrawn via line 116 condensed in exchanger 118 and transferred to high pressure separator 120, from which recycle hydrogen is withdrawn via line 110. The liquid condensate in separator 120 is flashed via line 122 into low pressure separator 124, from which light hydrocarbon gases are exhausted via line 126. The liquid phase in separator 124 is then transferred via line 126 to fractionating column 128, from which the desired gasoline fraction is taken off overhead via line 130, and may if desired be blended with the total gasoline production in line 84. The bottoms from column 128 is withdrawn via line 132, and may either be recycled entirely to hydrocracker 114 via line 134, or may be withdrawn via line 136 and further fractionated to obtain a desired proportion of high-quality jet fuel and/or diesel fuel.

The initial residual feedstocks employed herein may comprise any residue from the topping, fractionation or vacuum distillation of crude oils. Specifically, it is preferred to treat residual oils having an API gravity of about 2° to 15°, and boiling above about 850° F., and normally above about 1,000° F. Where lighter residua are found, it is normally preferable to subject them to further distillation to recover the light components rather than to subject the entire fraction to the donor treatment.

Suitable hydrocracking catalysts for use in the hydrocracking operations described above may comprise any desired combination of a refractory cracking base with a suitable hydrogenating component. Suitable cracking bases include for example mixtures of two or more difficultly reducible oxides such as silica-alumina, silica-magnesia, silica-zirconia, alumina-boria, silica-titania, silica-zirconia-titania, acid treated clays and the like. Acidic metal phosphates such as aluminum phosphate may also be used. The preferred cracking bases comprise composites of silica and alumina containing about 50-90% silica; coprecipitated composites of silica, titania, and zirconia containing between 5 and 75% of each component; partially dehydrated, zeolitic, crystalline molecular sieves, e.g., of the "X" or "Y" crystal types, having relatively uniform pore diameters of about 8 to 14 Angstroms, and comprising silica, alumina and one or more exchangeable zeolitic cations.

A particularly active and useful class of molecular sieve cracking bases are those having a relatively high SiO₂/Al₂O₃ ratio, e.g., between about 2.5 and 6.0. The most active forms are those wherein the exchangeable zeolitic cations are hydrogen and/or a divalent metal such as magnesium, calcium or zinc. In particular, the "Y" molecular sieves, wherein the SiO₂/Al₂O₃ ratio is about 5, are preferred, either in their hydrogen form, or a divalent metal form. Normally, such molecular sieves are prepared first in the sodium or potassium form, and the monovalent metal is ion-exchanged out with a divalent metal, or where the hydrogen form is desired, with an ammonium salt followed by heating to decompose the zeolitic ammonium ion and leave a hydrogen ion. It is not necessary to exchange out all of the monovalent metal; the final compositions may contain up to about 6% by weight of Na₂O, or equivalent amounts of other monovalent metals. Molecular sieves of this nature are described more particularly in Belgian Patents Nos. 577,642, 598,582, 598,683 and 598,682.

As in the case of the X molecular sieves, the Y sieves also contain pores of relatively uniform diameter in the individual crystals. In the case of X sieves, the pore diameters may range between about 6 and 14 Å., depending upon the metal ions present, and this is likewise the case in the Y sieves, although the latter usually are found to have crystal pores of about 9 to 10 Å. in diameter.

The foregoing cracking bases are compounded, as by impregnation, with from about 0.5% to 25% (based on

free metal) of a Group VIB or Group VIII metal promoter, e.g., an oxide or sulfide of chromium, tungsten, cobalt, nickel, or the corresponding free metals, or any combination thereof. Alternatively, even smaller proportions, between about 0.05% and 2% of the noble metals, e.g., platinum, palladium, rhodium or iridium, may be employed. The oxides and sulfides of other transitional metals may also be used, but to less advantage than the foregoing.

In the case of zeolitic type cracking bases, it is desirable to deposit the hydrogenating metal thereon by ion exchange. This can be accomplished by digesting the zeolite with an aqueous solution of a suitable compound of the desired metal, wherein the metal is present in a cationic form, and then reducing to form the free metal, as described for example in Belgian Patent No. 598,686.

A particularly suitable class of hydrocracking catalysts is composed of about 75-95% by weight of a coprecipitated hydrocracking base containing 5-75% SiO_2 , 5-75% ZrO_2 , and 5-75% TiO_2 , and incorporated therein from about 5-25%, based on free metal, of a Group VIII metal or metal sulfide, e.g., nickel or nickel sulfide.

Where a hydrofining treatment is desired, as in contacting zone 62, substantially any conventional hydrofining catalyst may be employed. Suitable hydrofining catalysts include for example mixtures of the oxides and/or sulfides of cobalt and molybdenum, or of nickel and tungsten, preferably supported on a carrier such as alumina, or alumina containing a small amount of coprecipitated silica gel. Other suitable catalysts include in general the oxides and/or sulfides of the Group VIB and/or Group VIII metals, preferably supported on adsorbent oxide carriers such as alumina, silica, titania, and the like.

As a more specific illustration of the process as described above, the following example is cited, which should not however be construed as limiting in scope:

EXAMPLE

Feedstock

A residual oil obtained by vacuum distillation of mixed Los Angeles basin and Kuwait crude oils, and having the following properties:

Gravity, ° API	6.8.
Total nitrogen, wt. percent	1.0.
Total sulfur, wt. percent	3.0.
Carbon residue wt. percent	13.3.
Ash, p.p.m.	900.
Boiling range	93% above 1,000° F.

Hydrogen Donor Treatment

Conditions:

Temperature, ° F.	800.
Pressure, p.s.i.g.	500.
Hydrogen added	None.
LHSV	1-2 (av. 1.5).
Volume ratio, donor/feed	1/1.

Feed to Donor Treatment (Donor Plus Residual Oil Feed)

Gravity, ° API	13.4
Volume percent boiling below 435° F.	4
435-750° F. fraction, vol. percent	50
750° F.+fraction, vol. percent	46

Product of Donor Treatment

Gravity, ° API	15.2
Wt. percent boiling below 435° F.	22.2
435-750° F. fraction, wt. percent	48.4
750° F.+fraction, wt. percent	28.7

Process Conditions in Hydrofining-Hydrocracking Reactor 60

Hydrofining:

Feed—435-750° F. product fraction from donor treatment.	
Catalyst—3% CoO, 15% MoO ₃ on 5% SiO ₂ , 95% Al ₂ O ₃ carrier; catalyst presulfided.	
Temperature, av. bed, ° F.	720
Pressure, p.s.i.g.	1,500
H ₂ /oil ratio, s.c.f./b.	8,000
LHSV	2

Hydrocracking:

Feed—Total effluent from hydrofining.	
Catalyst—0.5% Pd on decationized Y molecular sieve diluted with an equal weight of 25% NiO-75% activated Al ₂ O ₃ .	
Temperature, ° F.	725
Pressure, p.s.i.g.	1,500
H ₂ /oil ratio, s.c.f./b.	8,000
LHSV	2
Conversion per pass to 435° F. end-point gasoline, vol. percent	40

Solvent Extraction of 435° F.+Hydrocracker Effluent

Solvent	Furfural
Solvent/oil ratio	1/1
Extract fraction (hydrogen donor to be recycled to donor treatment):	
Vol. percent of feed to extraction	39
Gravity, ° API	17.3
Acid solubility, vol. percent	95
Boiling range, ° F.	430-705
Raffinate fraction (feed to second-stage hydrocracker 114):	
Vol. percent of feed to extraction	61
Gravity, ° API	40.6
Acid solubility, vol. percent	5
Boiling range, ° F.	430-700

Process Conditions in Second Stage Hydrocracker 114

Feed—Raffinate fraction from solvent extraction plus 435° F.+recycle oil from hydrocracker 114.	
Catalyst—Same as in hydrocracker 60.	
Temperature, ° F.	615
Pressure, p.s.i.g.	1,500
H ₂ /oil ratio, s.c.f./b.	8,000
LHSV	2
Conversion per pass to 435° F. end-point gasoline, vol. percent	60

Results analogous to those indicated in the foregoing example are obtained when other hydrocracking catalysts and conditions, other feedstocks and other process conditions within the broad purview of the above disclosure are employed. It is hence not intended to limit the invention to the details of the example or the drawing, but only broadly as defined in the following claims.

I claim:

1. A process for converting a residual mineral oil feedstock boiling above about 800° F. to lower boiling hydrocarbons, which comprises:

(A) subjecting said feedstock to a non-catalytic hydrogen donor treatment at a temperature between about 650° and 950° F. in admixture with a recycle stream of hydrogen donor hydrocarbons derived as hereinafter defined, said hydrogen donor treatment being carried out at a liquid hourly space velocity of about 0.2 to 10 volumes of combined liquid feed per volume of reactor space per hour;

(B) fractionating the effluent from said hydrogen donor treatment to recover a distillate gas oil fraction and a high-boiling bottoms fraction;

(C) subjecting said distillate gas oil fraction plus added hydrogen to catalytic hydrocracking in contact with a hydrocracking catalyst under conditions of temperature between about 600° and 850° F.

and pressure between about 400 and 2,500 p.s.i.g., said conditions being correlated to give about 20-60 volume-percent conversion to gasoline per pass, said hydrocracking catalyst comprising a Group VIB and/or Group VIII metal hydrogenating component deposited upon a solid cracking base selected from the class consisting of coprecipitated composites of silica-alumina, silica-magnesia, silica-zirconia, alumina-boria, silica-titania, and silica-zirconia-titania; acid-treated clays, acidic aluminum phosphates, and zeolitic alumino-silicate molecular sieves;

(D) fractionating effluent from said hydrocracking step to recover a gasoline product fraction and a higher boiling fraction comprising partially hydrogenated aromatic fused-ring polycyclic hydrogen donor hydrocarbons; and

(E) recycling at least a portion of said hydrogen donor hydrocarbons to said hydrogen donor treatment step (A).

2. A process as defined in claim 1 wherein an extraneous distillate gas oil feed is blended with the feed to said hydrocracking step (C).

3. A process as defined in claim 1 wherein at least a portion of said bottoms fraction from step (B) is recycled to said hydrogen donor treatment step (A).

4. A process as defined in claim 1 wherein a selected portion of said hydrogen donor hydrocarbons recovered from step (D) is recycled to said hydrocracking step (C), the remainder thereof being recycled to said hydrogen donor treatment step (A), so as to maintain a desired ratio of hydrogen donor hydrocarbons to said residual mineral oil feedstock.

5. A process as defined in claim 1 wherein said distillate gas oil recovered from step (B) is subjected to an intervening catalytic hydrofining treatment, and the total effluent from said hydrofining treatment is passed directly to said hydrocracking step (C).

6. A process as defined in claim 1 wherein said hydrocracking catalyst comprises a crystalline, zeolitic molecular sieve cracking base wherein the zeolitic cations are selected from the class consisting of hydrogen and divalent metals, and intimately compounded therewith a minor proportion of a Group VIII metal hydrogenating component.

7. A process as defined in claim 6 wherein said Group VIII metal is selected from the class consisting of palladium and platinum.

8. A process for converting a sulfur-, nitrogen- and metal-contaminated residual mineral oil feedstock boiling above about 800° F. to lower boiling hydrocarbons, which comprises:

(A) subjecting said feedstock to a non-catalytic hydrogen donor treatment at a temperature between about 650° and 950° F. in admixture with a recycle stream of hydrogen donor hydrocarbons derived as hereinafter defined, said hydrogen donor treatment being carried out at a liquid hourly spaced velocity of about 0.2 to 10 volumes of combined liquid feed per volume of reactor space per hour;

(B) fractionating the effluent from said hydrogen donor treatment to recover a distillate gas oil fraction and a high-boiling bottoms fraction;

(C) subjecting said distillate gas oil fraction plus added hydrogen to catalytic hydrocracking in contact with

a hydrocracking catalyst under conditions of temperature between about 600° and 850° F. and pressure between about 400 and 2,500 p.s.i.g., said conditions being correlated to give about 20-60 volume-percent conversion to gasoline per pass, said hydrocracking catalyst comprising a Group VIB and/or Group VIII metal hydrogenating component deposited upon a solid cracking base selected from the class consisting of coprecipitated composites of silica-alumina, silica-magnesia, silica-zirconia, alumina-boria, silica-titania, and silica-zirconia-titania; acid-treated clays, acidic aluminum phosphates, and zeolitic alumino-silicate molecular sieves;

(D) fractionating effluent from said hydrocracking step to recover a gasoline product fraction and a higher boiling fraction comprising partially hydrogenated aromatic fused-ring polycyclic hydrogen donor hydrocarbons; and

(E) subjecting said higher boiling fraction from step (D) to a separation step to segregate (a) an aromatic fraction rich in said hydrogen donor hydrocarbons and (b) a non-aromatic fraction rich in paraffinic and naphthenic hydrocarbons; and

(F) recycling at least a portion of said aromatic fraction (a) to said hydrogen donor treatment step (A) to supply said recycle hydrogen donor hydrocarbons.

9. A process as defined in claim 8 wherein said non-aromatic fraction (b) from step (E), plus added hydrogen, is subjected to a second stage of catalytic hydrocracking in contact with a hydrocracking catalyst under conditions effective to give a substantial conversion to gasoline, but at lower hydrocracking temperatures than were employed in said hydrocracking step (C).

10. A process as defined in claim 8 wherein said separation step (E) is solvent extraction.

11. A process as defined in claim 8 wherein an extraneous distillate gas oil feed is blended with the feed to said hydrocracking step (C).

12. A process as defined in claim 8 wherein at least a portion of said bottoms fraction from step (B) is recycled to said hydrogen donor treatment step (A).

13. A process as defined in claim 8 wherein a selected portion of said aromatic fraction recovered from step (E) is recycled to said hydrocracking step (C), the remainder thereof being recycled to said hydrogen donor treatment step (A), so as to maintain a desired ratio of hydrogen donor hydrocarbons to said residual mineral oil feedstock.

14. A process as defined in claim 8 wherein said distillate gas oil recovered from step (B) is subjected to an intervening catalytic hydrofining treatment, and the total effluent from said hydrofining treatment is passed directly to said hydrocracking step (C).

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