

[54] **HYDROCRACKING EXTRACTION** 3,702,817 11/1972 Cummins et al. 208/18
PROCESS FOR LUBES 3,766,055 10/1973 Cummins et al. 208/96
3,806,445 4/1974 Henry et al. 208/18

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Related U.S. Application Data

[63] Continuation of Ser. No. 285,203, Aug. 31, 1972,
abandoned.

[52] U.S. Cl. **208/96; 208/18**

[51] Int. Cl.² **C10G 37/00**

[58] Field of Search. **208/96, 18**

[56] **References Cited**

UNITED STATES PATENTS

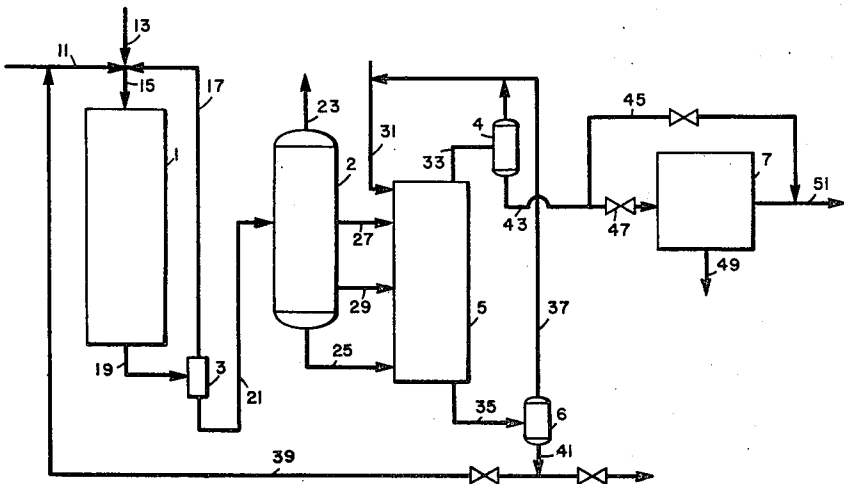
3,256,175 6/1966 Hozlowski et al. 208/87

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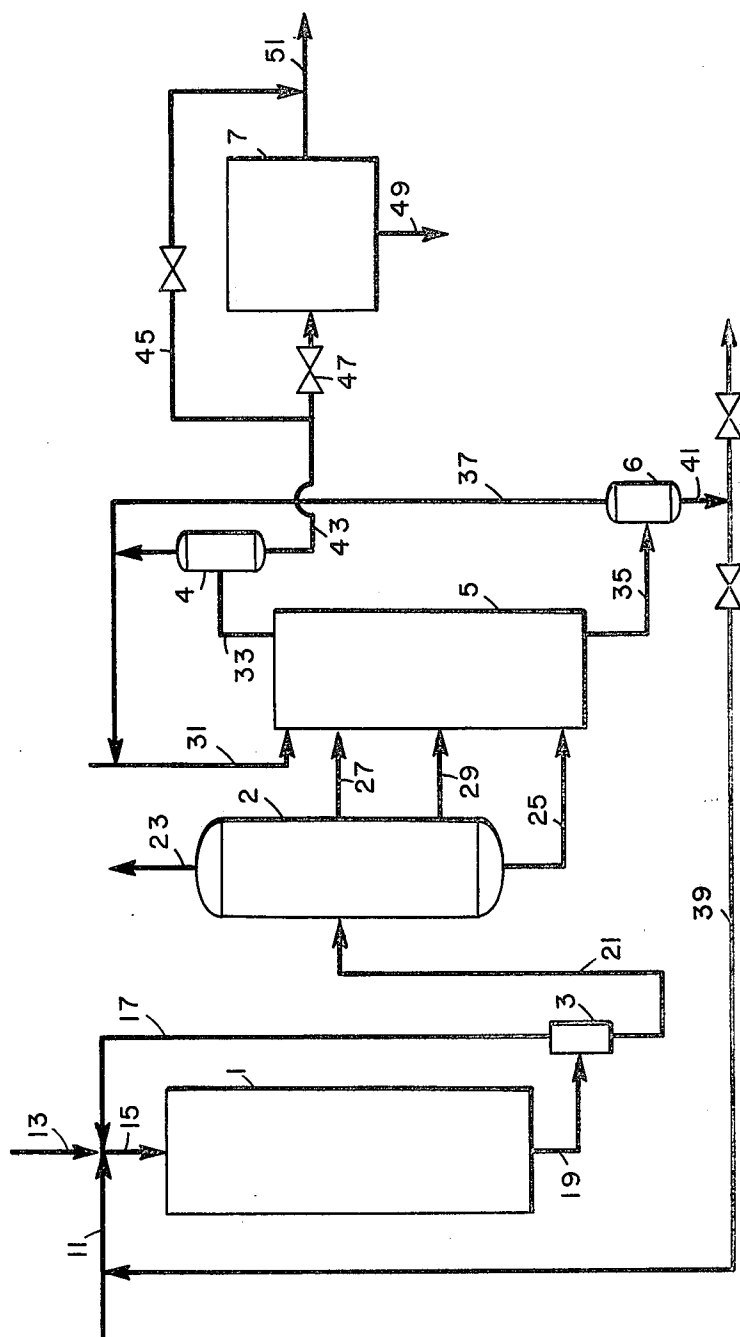
[57] **ABSTRACT**

High quality UV stable lubricating oil stocks are prepared by hydrocracking a hydrocarbon feedstock under mild hydrocracking conditions to increase the viscosity index of the feedstock. The hydrocrackate product is subsequently solvent-extracted with a solvent having preferential solubility for aromatics, thereby forming extract and raffinate phases. The extract phase is stripped of solvent, and at least a portion of the substantially solvent-free extract phase is recycled to the hydrocracking step.

13 Claims, 1 Drawing Figure



IMPROVED HYDROCRACKING-EXTRACTION PROCESS FOR LUBES



HYDROCRACKING EXTRACTION PROCESS FOR LUBES

This is a continuation, of application Ser. No. 285,203 filed Aug. 31, 1972 now abandoned.

BACKGROUND OF THE INVENTION

1. Field of the Invention

The present invention relates to a process for the upgrading of lubricating oil stocks. More specifically, the process relates to improvements in the viscosity index and UV stability of lubricating oil stocks.

2. Description of the Prior Art

Several processes have been used in the past for upgrading lubricating oil stocks. These processes have generally included either solvent extraction and/or hydrogenation including hydrocracking. Generally, the combination hydrocrackingsolvent extraction processes have involved subjecting hydrocarbon feedstocks, such as distillates and deasphalted oils, to hydrocracking at temperatures on the order of about 650°F in the presence of a catalyst and at relatively high hydrogen partial pressures.

Recent developments in hydrocracking processes have led to increased interest in commercialization of such processes as a method for upgrading lubricating oil stocks. This has been due in part to a decline in the availability of high quality feedstocks from which high VI, UV stable lubricating oils could be derived by conventional operations, such as solvent extraction. This, in turn, has led to the perfection and utilization of hydrocracking-solvent extraction combination processes on lower grade feedstocks. Typical of such processes is that disclosed in U.S. Pat. No. 3,660,273 issued May 2, 1972.

The prior art processes have generally suffered from several disadvantages including the use of relatively high hydrogen partial pressures in the hydrocracking operation. Specifically, pressures above 1500 pounds per square inch and preferably above 2000 pounds per square inch have been employed. This is undesirable from an economic standpoint and methods by which lower hydrocracking pressures could be used would be desirable. In addition, other disadvantages, such as high hydrogen consumption rates have been noted in prior art processes. Finally, it has been found that the overall yield of lubricating oils with the desired UV stability has been generally low.

In summary, then, it would be economically desirable to provide a process for the upgrading of low quality crude sources to high quality lubricating oil.

SUMMARY OF THE INVENTION

In accordance with this invention, it has now been discovered that an economically feasible process for the production of high VI, UV stable lubricating oil products can be achieved. Specifically, the process comprises (a) contacting a hydrocarbon feedstock, in a hydrocracking zone, with hydrogen in the presence of a hydrocracking catalyst, at mild hydrocracking conditions, thereby forming a hydrocrackate product, (b) solvent extracting at least a portion of the hydrocrackate with a solvent having preferential solubility for aromatics, thereby forming extract and raffinate phases, and (c) recycling at least a portion of the extract phase, substantially free of solvent, to step (a) wherein it is introduced into the hydrocracking operation either

separately or in combination with the hydrocarbon feedstock.

Feedstocks that are suitable for use in the subject process include hydrocarbons, mixtures of hydrocarbons, and, particularly, hydrocarbon fractions, the predominant portions of which exhibit initial boiling points above about 650°F. Unless otherwise indicated, boiling points are taken at atmospheric pressure. Nonlimiting examples of useful process feedstocks include crude oil vacuum distillates from paraffinic or naphthenic crudes, e.g., deasphalted residual oils, the heaviest fractions of catalytic cracking cycle oils, coker distillates and/or thermally cracked oils, heavy vacuum gas oils and the like. These fractions may be derived from petroleum crude oils, shale oils, tar sand oils, coal hydrogenation products and the like. Preferred feedstocks include deasphalted petroleum oils that exhibit initial boiling points in the range of from about 930° to about 1050°F and a Conradson Carbon Residue number less than about 3 and heavy gas oils that boil predominantly between about 650° and 1050°F and exhibit viscosities ranging from about 35 to 200, preferably 40 to 100 SUS at 210°F. It is desirable that the feedstocks to the subject process be of sufficiently low asphalt and metal content to prevent possible catalyst fouling in the hydrocracking operation. Any conventional deasphalting operation may be used which fulfills these objectives. Although not critical to the efficiency of the process, the feedstock should have a viscosity index above 0 and preferably above about 30. The viscosity index referred to herein is the extended viscosity index (VI_E) (ASTM D2270-64).

The instant process produces high quality lubricating oils from starting feedstocks derived from both good quality lube crudes, such as light Arabian and North Louisiana crudes and low quality "non-lube" crudes, such as West Texas Sour and Hawkins crudes. The "non-lube" crudes are generally characterized by having low VI ceilings, high sulfur content (e.g., greater than 2 wt. % based on total feed) and producing low yields of the desired lubricating oil products by solvent refining. It should be pointed out that the process feedstocks may not only be derived from high sulfur content crudes, but may themselves be of high sulfur content, e.g., greater than 2 wt. % sulfur, based on total feed.

The hydrocarbon feedstock hereinabove described is introduced into a hydrocracking zone or zones wherein it is contacted with hydrogen in the presence of one or more hydrocracking catalysts. Illustrative but nonlimiting examples of the hydrocracking procedure and modifications used are disclosed in U.S. Pat. Nos. 3,617,476; 3,242,068 and 3,579,437, the disclosures of which are incorporated herein by reference.

The preferred hydrocracking procedure is characterized by having overall low hydrogen consumption and comprises hydrocracking the raffinate obtained above at temperatures ranging between about 500° and 825°, preferably between about 650° and 750°F. Pressures which may be employed in the hydrocracking process are generally below about 1500 psig, preferably below about 1000 psig, and most preferably ranging between about 400 psig and about 1000 psig. In addition, liquid hourly space velocities can range between about 0.2 and 4 V/V/Hr and more preferably between about 0.4 and 1.5 V/V/Hr.

The hydrocracking operation is conducted, as previously indicated, in the presence of one or more hydro-

cracking catalysts, i.e., having both aromatic saturation and ring scission activity. The catalyst may be any conventional hydrocracking catalyst, such as, for example, that described in U.S. Pat. Nos. 3,535,230 and 3,287,252, the disclosures of which are incorporated herein by reference. Thus, the catalyst comprises a mixture of a major amount of an amorphous component and a minor amount of a hydrogenation component preferably comprising one or more transitional metals selected from Groups VIB and/or VIII of the Periodic Table and the oxides and sulfides thereof. The catalyst may also contain a minor amount of P_2O_5 which acts to stabilize the catalyst and increase its overall activity.

Representative of these metals are molybdenum, chromium, tungsten, nickel, cobalt, palladium, iron, rhodium, and the like, as well as combinations of these metals and/or their oxides and/or sulfides. Preferred metals are nickel, molybdenum and mixtures thereof. One or more of the metals, metal oxides or sulfides, alone or in combination, may be added to the support in minor proportions ranging from 1 to 25 wt.% based on the total catalyst.

The amorphous component, i.e., support, can be one or more of a large number of noncrystalline materials having high porosity. The porous material is preferably inorganic but can be organic in nature if desired. Representative porous materials that can be employed include metals and metal alloys; sintered glass; firebrick; diatomaceous earth; inorganic refractory oxides; organic resins, such as polyesters, phenolics and the like; metal phosphates such as boron phosphate, calcium phosphate and zirconium phosphate; metal sulfides such as iron sulfide and nickel sulfides; inorganic oxide gels and the like. Preferred inorganic oxide support materials include one or more oxides of metals selected from Groups IIA, IIIA and IV of the Periodic Table. Nonlimiting examples of such oxides include aluminum oxide, titania, zirconia, magnesium oxide, silicon oxide, titanium oxide, silica-stabilized alumina and the like.

Preferably, the starting catalyst composition comprises a silica/alumina support containing molybdenum trioxide and nickel oxide hydrogenation components. The silica:alumina weight ratio in the amorphous support can range from 20:1 to 1:20 and preferably from 1:4 to 1:6. The molybdenum trioxide: nickel oxide weight ratio in the amorphous support can range from about 1:25 to 25:1 and preferably from 2:1 to 4:1. Finally, the weight ratio of the support to the hydrogenation component can range from about 20:1 to 1:20 and preferably from 4:1 to 6:1. A particularly preferred starting catalyst composition comprises:

NiO	4.5 wt. %
MoO ₃	13.0 wt. %
SiO ₂	14.0 wt. %
Al ₂ O ₃	68.4 wt. %

The catalyst is preferably presulfided by conventional methods such as by treatment with hydrogen sulfide or carbon disulfide prior to use. The precise chemical identity of the hydrogenation constituents present on the support during the course of the hydrocracking operation is not known. However, the hydrogenation components probably exist in a mixed elemental metal/metal oxide/metal sulfide form.

Additionally, low sieve-content catalysts consisting of a mixture of a major amount of an amorphous component and minor amounts of (1) a crystalline aluminosilicate component comprising less than about 25 wt.%, preferably less than about 5 wt.% of the total catalyst and (2) a hydrogenation component, can be used as hydrocracking catalysts. The catalyst may also contain a small amount of P_2O_5 , which acts to stabilize the catalyst against decomposition. The amorphous component (support) is similar to that described above. The hydrogenation component is preferably a transitional metal selected from Groups VIB and VIII of the Periodic Table and/or the oxides and/or sulfides thereof. Useful catalyst metals include chromium, molybdenum, tungsten, platinum, palladium, cobalt, nickel, etc. One such catalyst comprises 95 wt.% based on total catalyst of NiO/MoO₃ on a SiO₂/Al₂O₃ base (stabilized with P_2O_5) and 5 wt.% based on total catalyst of nickelexchanged faujasite.

The crystalline aluminosilicate (sieve component) employed in the preparation of the crystalline component of the catalyst comprises one or more natural or synthetic zeolites. Representative examples of particularly preferred zeolites are zeolite X, zeolite Y, zeolite L, faujasite and mordenite. Synthetic zeolites have been generally described in U.S. Pat. Nos. 2,882,244, 3,130,007 and 3,216,789, the disclosures of which are incorporated herein by reference.

The silica:alumina mole ratio of useful aluminosilicates is greater than 2.5 and preferably ranges from about 2.5 to 10. Most preferably this ratio ranges between about 3 and 6. These materials are essentially the dehydrated forms of crystalline hydrous siliceous zeolites containing varying quantities of alkali metal and aluminum with or without other metals. The alkali metal atoms, silicon, aluminum and oxygen in the zeolites are arranged in the form of an aluminosilicate salt in a definite and consistent crystalline structure. The structure contains a large number of small cavities, interconnected by a number of still smaller holes or channels. These cavities and channels are uniform in size. The pore diameter size of the crystalline aluminosilicate can range from 5 to 15 Å and preferably from 5 to 10 Å.

The aluminosilicate component may comprise a sieve of one specific pore diameter size or, alternatively, mixtures of sieves of varying pore diameter size. Thus, for example, mixtures of 5 Å and 13 Å sieves may be employed as the aluminosilicate component. Synthetic zeolites such as type-Y faujasites are preferred and are prepared by well-known methods such as those described in U.S. Pat. No. 3,130,007.

The aluminosilicate can be in the hydrogen form, in the polyvalent metal form, or in the mixed hydrogen-polyvalent metal form. The polyvalent metal or hydrogen form of the aluminosilicate component can be prepared by any of the well known methods described in the literature. Representative of such methods is ion-exchange of the alkali metal cations contained in the aluminosilicate with ammonium ions or other easily decomposable cations such as methyl-substituted quaternary ammonium ions. The exchanged aluminosilicate is then heated at elevated temperatures of about 570° to 1112°F to drive off ammonia, thereby producing the hydrogen form of the material. The degree of polyvalent-metal or hydrogen exchange should be at least about 20%, and preferably at least about 40% of the maximum theoretically possible. In any event, the

crystalline aluminosilicate composition should contain less than about 6.0 wt.% of the alkali metal oxide based on the final aluminosilicate composition and, preferably, less than 2.0 wt.%, i.e., about 0.3 to 0.5 wt.% or less.

The resulting hydrogen aluminosilicates can be employed as such, or can be subjected to a steam treatment at elevated temperatures, i.e., 800° to 1300°F, for example, to effect stabilization, thereof, against hydrothermal degradation. The steam treatment, in many cases, also appears to effect a desirable alteration in crystal structures resulting in improved selectivity.

The mixed hydrogen-polyvalent metal forms of the aluminosilicates are also contemplated. In one embodiment the metal form of the aluminosilicate in ion-exchanged with ammonium cations and then partially back-exchanged with solutions of the desired metal salts until the desired degree of exchange is achieved. The remaining ammonium ions are decomposed later to hydrogen ions during thermal activation. Here again it is preferred that at least about 40% of the monovalent metal cations be replaced with hydrogen and polyvalent metal ions.

Suitably, the exchanged polyvalent metals are transition metals and are preferably selected from Groups VIB and VIII of the Periodic Table. Preferred metals include nickel, molybdenum, tungsten and the like. The most preferred metal is nickel. The amount of nickel (or other metal) present in the aluminosilicate (as ion-exchanged metal) can range from about 0.1 to 20% by weight based on the final aluminosilicate composition.

In addition to the ion-exchanged polyvalent metals, the aluminosilicate may contain as nonexchanged constituents one or more hydrogenation components comprising the transitional metals, preferably selected from Groups VIB and VIII of the Periodic Table and their oxides and sulfides. Such hydrogenation components may be combined with the aluminosilicate by any method which gives a suitably intimate admixture, such as by impregnation. Examples of suitable hydrogenation metals, for use herein, include nickel, tungsten, molybdenum, platinum, and the like, and/or the oxides and/or sulfides thereof. Mixtures of any two or more of such components may also be employed. Particularly preferred metals are tungsten and nickel. Most preferably, the metals are used in the form of their oxides. The total amount of hydrogenation components present in the final aluminosilicate composition can range from about 1 to 50 wt.%, preferably from 10 to 25 wt.% based on the final aluminosilicate composition.

The amorphous component and the crystalline aluminosilicate component of the low sieve content catalyst may be brought together by any suitable method, such as by mechanical mixing of the particles thereby producing a particle form composite that is subsequently dried and calcined. The catalyst may also be prepared by extrusion of wet plastic mixtures of the powdered components followed by drying and calcination. Preferably the complete catalyst is prepared by mixing the metal-exchanged zeolite component with alumina or silica-stabilized alumina and extruding the mixture to form catalyst pellets. The pellets are thereafter impregnated with an aqueous solution of nickel and molybdenum or tungsten materials to form the final catalyst.

The hydrocrackate is subsequently removed from the hydrocracking zone and at least a portion of the hydro-

crackate is contacted with a solvent or mixture of solvents having preferential solubility for aromatics. It is noted that the hydrocrackate product may first be separated into various lube fractions followed by individual extraction of each fraction; alternatively, the hydrocrackate product may first be extracted as a broad cut, containing substantially all of the lube oil components in the hydrocrackate product, say, for example, the 650°F+ fraction, followed by separation of the raffinate into the individual lube fractions.

The particular solvent which is used in the extraction operation depends upon several considerations, such as for example, the basic economics. Any solvent selective for aromatics, particularly polycyclic aromatics, may be used such as furfural, acetophenone, liquid SO₂, acetonitrile, phenol, nitrobenzene, aniline, 2,2-dichlorodiethylether, dimethyl sulfoxide, dimethyl formamide, n-methyl, 2-pyrrolidone, and mixtures thereof. In addition, any of the aforementioned solvents may be used in combination with an anti-solvent, such as water. In general, the most preferred solvents are phenol, furfural, and phenol/water. The latter solvent is most preferred and will contain from about 0.1 to 15LV% water, preferably 3 to 8LV % water, based on the total solvent mixture.

In general, the various means customarily utilized in extraction processes to increase the contact area between the oil stock and the solvent can be employed. Thus, the apparatus used in the instant process can comprise a single extraction zone or multiple extraction zones equipped with (a) shed rows or other stationary devices to encourage contacting; (b) orifice mixers; or (c) efficient stirring devices such as mechanical agitators, jets of restricted internal diameter, turbo mixers and the like. The operation may be conducted as a batch or continuous-type operation with the latter being preferred. A continuous countercurrent operation is most preferred. Known techniques for increasing selectivity for aromatics can be employed. Examples of these are the use of small amounts of anti-solvents, e.g. water, during the extraction with the solvent, operating at fairly low temperatures sufficient to carry out the extraction objectives, and using low-solvent-to-oil ratios. The solvents are generally used at dosages of about 100 to 600%, preferably about 100 to 300%.

The equipment employed in the extraction operation is not critical to the invention and can comprise rotating disc contactors, podbielniak contactors, countercurrent packed bed extraction columns and countercurrent tray contactors.

The temperatures of the extraction and the amount of solvent used are interdependent, and are, in turn, dependent upon the composition of the particular oil stock to be extracted. With this in mind the following extraction process points are noted. First, the extraction temperature is preferably maintained at about 40°F below the temperature of miscibility of the oil and solvent in order to obtain the desired extraction effect and to conduct a highly efficient extraction operation with good yields of oil. If the extraction temperature is too low, the extraction will be too selective and will require application of compensating features, such as additional amounts of solvent and extraction stages. The extraction temperature range is generally between about 0° and 350°F, preferably between about 80° and 250°F, depending upon the oil-solvent miscibility temperature. In the case of the preferred phenol-water solvent systems, the temperature ranges between about

120° and 200°F.

It is noted that high solvent-oil ratios tend to reduce operational efficiency, producing lower yields of raffinate as hydrocracker feed and are to be avoided. Thus, for the most part, solvent-oil ratios (defined as volume of solvent added per volume of oil) range between about 6:1 and about 0.25: 1. Particularly preferred ratios range between about 4:1 to about 0.8:1. For feedstocks derived from low lube quality crudes, such as vacuum gas oils and deasphalted oil derived from West Texas Sour crudes, typical extraction temperatures ranging between about 140° and 200°F may be used with solvent-to-oil ratios of about 6.0:1 to about 1.0:1.

The resulting extract and raffinate phases are separated and at least a portion of the solvent is removed from the extract and raffinate phases and high quality lube oil recovered from the raffinate phase. The extract phase, substantially free of solvent, can be recycled in whole or in part to the hydrocracking operation wherein it is introduced alone or in combination with the hydrocracker feed.

The recycle feature of the subject invention results in a higher overall lube oil yield and in the production of higher quality lubricating oil product. While the specific mechanism of the recycle process is not known, it is speculated that the extract phase comprises a small amount of UV unstable components formed in the hydrocracking operation, which are believed to be partially hydrogenated polycyclic aromatic compounds, together with a larger amount of other components, chiefly aromatics and naphthenes which are of fairly good quality in terms of VI, color, stability and the like. Recycle of this extract to the hydrocracking stages enables a large part of This fraction to be recovered as lube oil product, thus increasing the overall lube oil yield. In addition high value fuels byproducts are produced during hydrocracking at the expense of low value extract yield.

It is noted that lube oil product recovered from the raffinate may be further treated such as by catalytic or ketone dewaxing, for example, in order to obtain further upgrading of the lube oil product.

BRIEF DESCRIPTION OF THE DRAWING

The FIGURE is a flow diagram of the overall process for the production of high quality lubricating oils.

suitable hydrocracking catalyst as hereinabove defined, for a time sufficient to obtain the desired conversion. The hydrocrackate product is removed from zone 1 via line 19 and passed through separator 3 wherein volatile gases formed during the hydrocracking process are removed. Recovered hydrogen is recycled via line 17 to hydrocracking zone 1 while the remaining hydrocrackate is introduced into pipestill 2 via line 21. The hydrocrackate is distilled and a fuels fraction recovered via line 23. Lubricating oil sidestreams are taken off through lines 27 and 29 and a bottoms fraction removed via line 25. The lubricating oil fractions are solvent extracted in extraction zone 5 at conditions hereinabove defined using such solvents as phenol/water and the like. An extract phase is removed from zone 5 via line 35 and passed through stripper 6 wherein solvent is separated and recycled to extraction zone 5 via lines 37 and 31. The extract phase is recovered from stripper 6 via line 41 and recycled in whole or in part via line 39 to hydrocracking zone 1. In this embodiment, the extract is shown as being introduced and mixed with the fresh feed. However, as indicated previously, the extract phase could be introduced into the hydrocracking zone separately from the fresh feed. The raffinate from the extraction zone 5 is removed via line 33 and passed through stripper 4 wherein excess solvent is removed and recycled via lines 37 and 31 to the extraction zone. Raffinate product is withdrawn from stripper 4 and, if desired, introduced into dewaxing zone 7 via lines 43 and 47. Wax is recovered via line 49 and dewaxed lube oil via line 51. Alternatively, the lubricating oil can bypass the dewaxing operation and go to storage via lines 45 and 51. Makeup solvent can be introduced into zone 5 via line 31.

DESCRIPTION OF THE PREFERRED EMBODIMENTS

The present invention will be more clearly understood by reference to the following nonlimiting examples:

A series of tests were performed to illustrate the hydrocracking-extraction process with light vacuum gas oil (LVGO) from West Texas Sour crude and heavy vacuum gas oil (HVGO) from Light Arabian crude. In addition, a broad cut blend from Light Arabian crude was studied. Properties of these feedstocks are shown in the following tabulation:

Feedstock	West Texas Sour LVGO	Light Arabian HVGO	Light Arabian Blend (Arma-A)
Boiling range, °F.	700-925	850-1050	750-1400
Gravity, °API	23.1	19.9	21.4
Refractive Index at 60°C.	1.4949	1.5098	1.5022
Viscosity 210°F, SUS	48.0	75.8	65.9
Sulphur, wt. %	2.0	2.6	2.4
Wax, wt. %	11	9	—
Dewaxed Viscosity 210°F, SUS	51	96	71.9
VI _E	50	54	64
Color, ASTM	D8.0	D8.0	D8.0

Turning to the drawing in detail, fresh feed such as a heavy vacuum gas oil boiling between about 650° and 1100°F and/or a deasphalted oil boiling above about 1000°F is introduced into hydrocracking zone 1 via lines 11 and 15. Makeup hydrogen is introduced into the hydrocracking zone via lines 13 and 15. The feedstock is contacted with hydrogen in the presence of a

In each of the following seven examples the solvent used in the countercurrent extraction stages was aqueous phenol and the catalyst used in the hydrocracking stage a conventional sulphided catalyst having the following composition (prior to sulphiding):

Composition	Wt. %
NiO	6
MoO ₃	13
SiO ₂	14
Al ₂ O ₃	Balance
Properties	
Surface Area, m ² /g	360
Pore Volume, ml./g.	0.5
Bulk Density, gm./ml.	0.56

EXAMPLE 1

West Texas Sour LVGO was extracted with 235 vol. % treat of phenol plus 3 vol. % water at 140°F to produce a 100 viscosity index dewaxed lube having a viscosity of 202 SUS at 100°F (Table I). This extraction operation produced 47 wt. % yield of UV stable raffinate on extraction with a color of approximately 3 ASTM. The extract exhibited a refractive index at 60°C of 1.523.

EXAMPLE 2

The feedstock of example 1 was mildly hydrocracked at 650°F, 1.0 v/v/hr., 600 psig H₂ and 1500 SCF H₂/B and then topped to 660°F by vacuum distillation (Table I). A 250 vol. % treat of aqueous phenol then produced the same overall yield of dewaxed UV stable lube as above with viscosity and viscosity index almost identical to that of Example 1. Compared to extraction alone the mild hydrocracking operation increased the raffinate yield on extraction from 47 to 55 wt. % and improved considerably the raffinate color, thereby eliminating the need for a hydrofining step after extraction as used in a conventional processing sequence. The yield of extract, based on LVGO, from the hydrocracked feed to extraction was reduced compared to extraction alone, but the quality was considerably improved. The reduction in extract refractive index of 0.015 resulting from the prehydrocracking is equivalent to approximately a 25 viscosity index improvement. Recycle of this upgraded extract to the hydrocracker feed results in approximately a 10 wt. % overall yield gain of dewaxed oil and production of valuable fuels by-products at the expense of low-value extract.

EXAMPLE 3

The feedstock of Example 1 was hydrocracked now at 725°F, 1.0 v/v/hr., 600 psig H₂ and 1500 SCF H₂/B and then topped to 750°F by vacuum distillation (Table I). A 130 vol. % aqueous phenol treat was adequate to produce a UV stable, light colored lube basestock of similar viscosity and viscosity index to those produced in Examples 1 to 2. The more severe prehydrocracking thus reduced considerably the treat requirements compared to extraction alone and mild prehydrocracking-extraction. In addition producing the same overall yield of basestock, based on LVGO feed, the raffinate yield on extraction was increased to 69 wt. % compared to 55 wt. % from low severity prehydrocracking. Thus, prehydrocracking a lube feedstock to extraction is an effective means of debottlenecking existing throughput-limited extractors. As was the case in Example 2, considerable lube yield gains result from recycling part or all of the upgraded extract to the hydrocracker feed. Alternatively, this extract may be mildly hydrocracked alone to produce low pour naphthenic lubricating oils.

EXAMPLE 4

Light Arabian HVGO was extracted with 160 vol. % treat of phenol plus 3 vol. % water at 170°F. to produce a 95 viscosity index dewaxed lube having a viscosity of 590 SUS at 100°F (Table 2). This extraction operation produced 38 wt. % yield of UV stable raffinate on extraction with a color of 5.0 ASTM. The extract exhibited a refractive index at 60°C of 1.535.

EXAMPLE 5

The feedstock of Example 4 was mildly hydrocracked at 650°F, 1.0 v/v/hr., 600 psig H₂ and 1500 SCF H₂/B and then topped to 700°F by vacuum distillation to remove non-lube products (Table II). A 155 vol. % treat of aqueous phenol produced 47 wt. % yield of dewaxed UV stable lube of 95 viscosity index and 510 SUS viscosity at 100°F. Thus, minimum topping of the prehydrocracked feed to extraction resulted in markedly improved overall dewaxed oil yield on HVGO as well as improved raffinate yield on extraction compared to extraction alone. The viscosity loss resulting from prehydrocracking can be compensated for by using a higher viscosity distillate feedstock. It can be noted that prehydrocracking produced lubes of improved color compared to extraction alone. Recycle of the upgraded extract from the prehydrocracked feedstock to the hydrocracker results in approximately 10 wt. % overall yield gain of dewaxed oil from the combination process.

EXAMPLE 6

The feedstock from example 4 was hydrocracked now at 750°F, 1.0 v/v/hr., 600 psig H₂ and 1500 SCF H₂/B and then topped to 700°F by vacuum distillation to remove non-lube products (Table II). A 70 vol. % aqueous phenol treat was adequate to produce a UV stable, light colored lube basestock of 95 VI and 410 SUS viscosity at 100°F. It can be seen that significantly improved overall dewaxed oil yields and raffinate yields on extraction have resulted from the prehydrocracking operation at both severity levels compared to extraction alone. Thus the prehydrocracking operation is beneficial for application with existing extractors which are throughput and/or feedstock limited as well as in new facilities to optimize lube throughput and minimize equipment costs.

EXAMPLE 7

For a broad cut feedstock, the hydrocracking-extraction processing sequence produces a UV stable lube slate of uniform viscosity index-viscosity distribution in a continuous, single pass operation. To illustrate such a process, a broad-cut lube feedstock blend from Light Arabian crude (Aram-A) was prehydrocracked at 723°F, 0.5 v/v/hr., 1500 psig H₂ and 5000 SCF H₂/B and topped to 700°F by vacuum distillation. The 700°F+ cut from hydrocracking was extracted with phenol plus 3 vol. % water at 165°F at three treat levels 150, 260 and 400 vol. % and then vacuum distilled into three lube cuts; a 700°-925°F Light Neutral lube, a 925°-1050°F Heavy Neutral lube and a 1050°F+ Bright Stock. Inspections on these products are shown in Table III. The good UV stability and the desirable uniform viscosity index-viscosity distribution of the lube products are shown in the following tabulation for the 260 vol. % treat case.

Dewaxed Lube °F.	Viscosity at 100°F. SUS	Viscosity Index	UV Stability ⁽¹⁾ of Hydrocracked Extracted Lubes	UV Stability ⁽¹⁾ of Equivalent Solvent Extracted Base
700-925	179	103	14	7-21
925-1000	569	98	>25	>25
1050+	2993	97	>25	>25

⁽¹⁾days in window to form sludge

Again, marked lube yield advantages result from recycling all or part of the extract from the prehydrocracked feedstock to the hydrocracker.

EXAMPLE 8

For West Texas Sour LVGO, the surprising results obtained by using prehydrocracking before extraction, in Examples 1 to 4, have also been obtained with the following catalysts:

Composition wt. %

NiO	2.8	3.5	3.5	3.5
MoO ₃	14	18	15	15
P ₂ O ₅	—	—	trace	trace
SiO ₂	1.5	—	5	5
Al ₂ O ₃	Balance	Balance	Balance	Balance
Faujasite (Ni stabilized)	—	—	—	5
Properties				
Surface Area, m ² /gm	270	158	178	173
Pore Volume, cc/gm	0.51	0.43	0.45	0.41
Bulk Density, gm/cc	0.73	0.77	0.79	0.77

EXAMPLE 9

For prehydrocracking Light Arabian HVGO over the catalyst used in Examples 1 to 7, hydrogen consumption can be minimized by using a low hydrogen pressure in the hydrocracking step, while still maintaining the benefits in improved raffinate yield, reduced sol-

below, for fixed hydrocracking temperature and space velocity, hydrogen consumption is reduced from 524 SCF/B to 295 SCE/B by reducing the reaction pressure from 1350 psig to 600 psig, the viscosity index of the 700°F.+ product remaining approximately constant.

Hydrocracking Conditions

Temperature, °F.	750	750
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LHSV, v/v/hr	1.0	1.0
Pressure, psig.	600	1350
H ₂ consumption, SCF/B	295	524
Dewaxed Oil Inspections		
700°F + yield on HVGO, wt. %	79	79
Viscosity 100°F, SUS.	442	442
VI _E	79	82

TABLE I

Feedstock Example	West Texas Sour LVGO		
	1 Extraction Alone	2 Hydrocracking	3 Extraction
Hydrocracking			
Temperature, °F.	—	650	725
H ₂ Consumption, SCF/B	—	324	318
Hydrocrackate Boiling Range, °F.	—	660+	750+
Yield on Hydrocracking, wt. %	100	95	68
DWO Viscosity 100°F, SUS	364	256	250
VI _E	50	61	76
UV stability	—	Fail	Fail
Extraction			
Water in Phenol, vol. %	3	3	3
Temperature, °F.	140	140	140
Treat, vol. %	235	250	130
Raffinate Yield on Extraction, wt. %	47	55	69
DWO Yield on LVGO, wt. %	37	37	38
Viscosity 100°F, SUS	202	199	201
VI _E	100	100	100
Color, ASTM	L3.0	L1.0	1.5
UV stability	Pass	Pass	Pass
Extract Yield on LVGO, wt. %	53	43	21
Refractive Index 60°C.	1.523	1.508	1.516

vent treat, etc. illustrated in the examples. As shown

TABLE II

Feedstock Example	Light Arabian HVGO		
	4 Extraction Alone	5 Hydrocracking	6 Extraction
Hydrocracking			
Temperature, °F.	—	650	750
H ₂ Consumption, SCF/B	—	251	295
Hydrocrackate Boiling Range, °F.	—	700+	700+
Yield on Hydrocracking, wt. %	100	98	89
DWO Viscosity 100°F, SUS	1738	940	486
VI _E	54	62	78

TABLE II-continued

Feedstock Example Hydrocracking	4 Extraction Alone	Light Arabian HVGO 5 Hydrocracking	6 Extraction
UV Stability	—	Fail	Fail
Extraction			
Water in Phenol, vol. %	3	3	3
Temperature, °F.	170	170	170
Treat, vol. %	160	155	70
Raffinate Yield on Extraction, wt. %	47	57	73
DWO Yield on HVGO, wt. %	38	47	57
Viscosity 100°F.	590	510	410
VI _E	95	95	95
Color, ASTM	5.0	3.5	4.0
UV stability	Pass	Pass	Pass
Extract Yield on HVGO, wt. %	53	42	24
Refractive Index 60°C.	1.535	1.523	1.529

TABLE III

	Aram-A Feed	Hydrocracking	Extraction		
Aqueous Phenol Treat, vol. %	—	—	150	260	400
700–925°F Cut					
Waxy Yield on Aram-A, wt. %	58.3	60.2	41.7	29.6	23.6
Dry Wax on Lube Cut, wt. %	10.2	10.4	13.6	15.6	16.3
Dewaxed Oil: Yield on Aram-A, wt. %	52.3	53.9	36.0	25.0	19.8
Visc. at 100°F, SUS	374	198.4	190	179	175
Visc. at 210°F, SUS	52.0	45.4	45.7	45.3	45.8
VI _E	58	84	98	103	108
Color, ASTM	4.5	L2.0	L1.0	L0.5	L0.5
925–1050°F Cut					
Waxy Yield on Aram-A, wt. %	24.9	18.4	18.4	17.1	14.4
Dry Wax on Lube Cut, wt. %	9.4	15.2	19.0	18.1	18.7
Dewaxed Oil: Yield on Aram-A, wt. %	22.5	15.6	14.9	14.0	11.7
Visc. at 100°F, SUS	1890	797	712	569	517
Visc. at 210°F, SUS	100.6	76.3	72.5	67.5	65.8
VI _E	58	90	91	98	103
Color, ASTM	D8.0	3.5	L2.0	L1.5	L1.5
1050°F+ Cut					
Waxy Yield on Aram-A, wt. %	16.8	11.2	10.4	10.1	10.8
Dry Wax on Lube Cut, wt. %	10.0	14.1	12.4	16.5	16.6
Dewaxed Oil: Yield on Aram-A, wt. %	15.1	9.6	9.1	8.4	9.0
Visc. at 100°F, SUS	10410	3185		2993	2537
Visc. at 210°F, SUS	311	175.4	176.2	170.6	159.4
VI _E	82	95	96	97	101
Color, ASTM	D8.0	D8.0	D8.0	L6.5	7.0

What is claimed is:

1. A process for producing high VI, UV stable lubricating oil products from substantially asphalt-free lube feed stocks that have not been subjected to solvent extraction for the removal of aromatics therefrom, as an extract which comprises the steps of (a) passing said lubricating oil stock and hydrogen into a hydrocracking zone maintained at a pressure below about 1500 psig with the presence of a hydrocracking catalyst, thereby forming a hydrocrackate product containing lube oil components, (b) solvent extracting at least a portion of said hydrocrackate product with a solvent having preferential solubility for aromatics, thereby forming extract and raffinate phases, and (c) recycling a substantial portion of the extract phase, substantially free of solvent, to step (a), wherein it is introduced into the hydrocracking zone either separately or in combination with said lubricating oil stock.

2. The process of claim 1 wherein the solvent is selected from the group consisting of furfural, acetophenone, liquid SO₂, acetonitrile, phenol, n-methyl-2-pyrrolidone, dimethyl formamide and mixtures thereof.

3. The process of claim 2 wherein the solvent is phenol, and wherein the phenol is admixed with a minor amount of water ranging between about 0.1 and 15LV% of the solvent mixture.

4. The process of claim 3 wherein the solvent extraction temperature ranges between about 120° and 200°F.

5. The process of claim 1 wherein the hydrocracking operation is conducted at a temperature ranging between about 500° and 825°F, at a pressure below about 1500 psig and at a liquid hourly space velocity ranging between about 0.2 and 4 V/V/Hour and wherein the catalyst comprises a major amount of an amorphous component and a minor amount of a hydrogenation component selected from the group consisting of the Group VIB and VIII transition metals, their oxides and sulfides and mixtures thereof.

6. The process of claim 5 wherein the catalyst contains, in addition to the amorphous component and the hydrogenation component, less than about 25 wt.%, based on total catalyst, of a crystalline aluminosilicate component.

7. The process of claim 1 wherein the solvent extraction temperature ranges between about 0° and 350°F.

8. The process of claim 1 wherein the solvent/oil liquid volume ratio in step (b) ranges between about 6:1 and 0.25:1.

9. The process of claim 1 wherein the hydrocrackate is separated into a broad cut fraction boiling above about 650°F, said fraction containing substantially all

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of the lube oil components in said hydrocrackate and, thereafter, solvent extracting said fraction in step (b).

10. The process of claim 1 wherein the catalyst comprises a major amount of an amorphous component and minor amounts of (1) a hydrogenation component selected from the group consisting of the Group VIB and VIII transition metals, their oxides and sulfides and mixtures thereof and (2) P_2O_5 .

11. The process of claim 1 wherein the catalyst comprises a major amount of an amorphous component and minor amounts of (1) a hydrogenation component selected from the group consisting of the Group VIB

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and VIII transition metals, their oxides and sulfides and mixtures thereof, (2) a crystalline aluminosilicate component and (3) P_2O_5 .

12. The process of claim 1 wherein said feed stock is a deasphalted petroleum oil exhibiting an initial boiling point in the range of 930° to about 1050°F and a Conradson Carbon Residue number less than about 3.

13. A process of claim 1 wherein said feed stock is a heavy gas oil boiling predominately between about 650° and 1050°F and exhibiting a viscosity ranging from about 35 to 200 SUS at 210°F.

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