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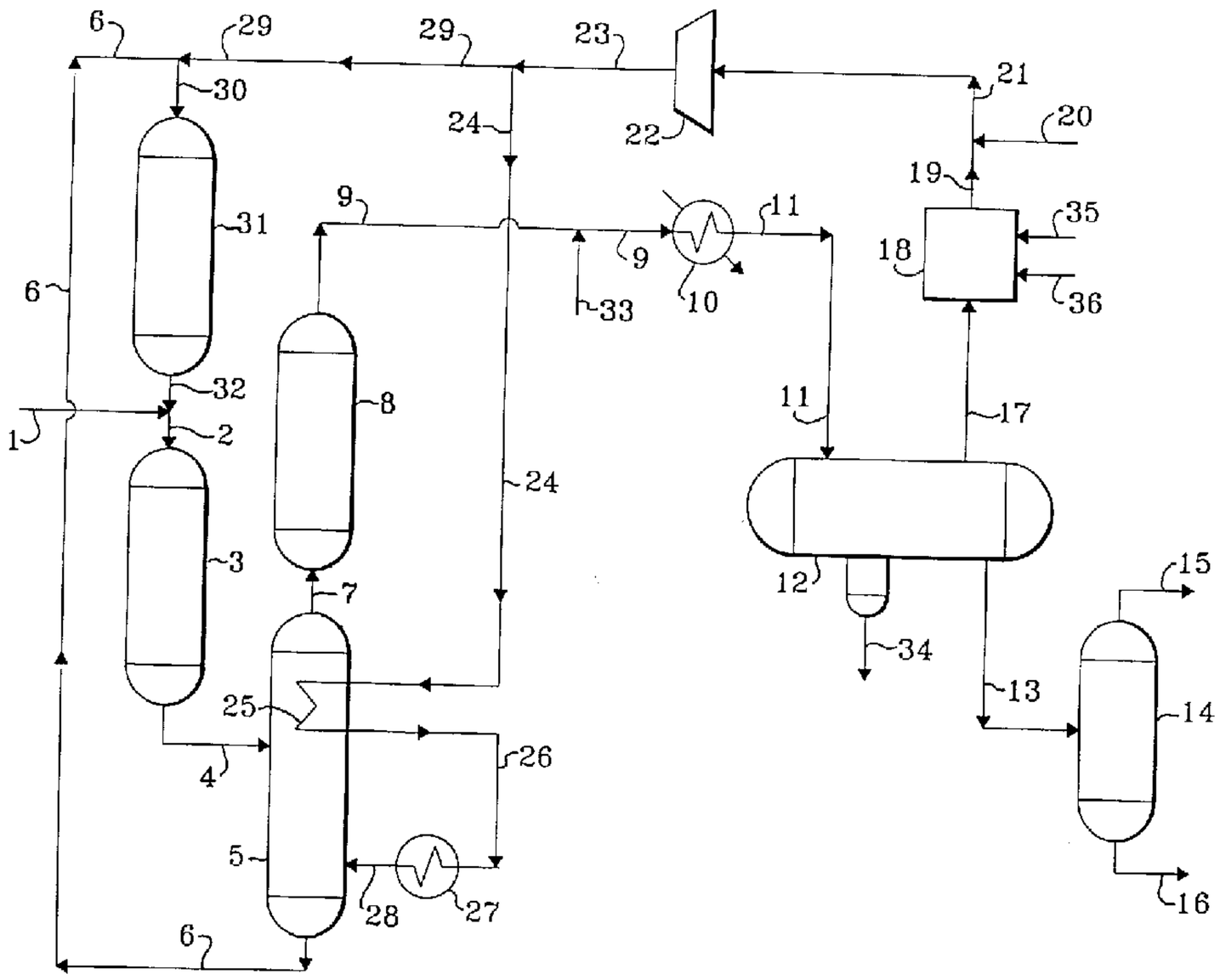
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(54) **PROCEDE INTEGRE D'HYDROTRAITEMENT ET D'HYDROCRAQUAGE**

(54) **INTEGRATED HYDROTREATING AND HYDROCRACKING PROCESS**



(57) A hydrocarbonaceous feedstock is converted in an integrated hydrotreating and hydrocracking process by combining with a hot hydrocracking zone effluent recycle stream containing hydrogen, and the mixture is passed to a denitrification and desulfurization reaction zone to produce hydrogen sulfide and ammonia to thereby clean up the fresh feedstock. The resulting hot, uncooled effluent from the denitrification and desulfurization zone is hydrogen stripped in a stripping zone maintained at essentially the same pressure as the hydrotreating zone with a hydrogen-rich gaseous stream to produce a vapor stream comprising hydrogen, hydrocarbonaceous compounds boiling at a temperature below the boiling range of the fresh feedstock, hydrogen sulfide and ammonia, and a liquid hydrocarbonaceous stream wherein at least a portion of this liquid stream is then passed to a hydrocracking step to produce the hot hydrocracking effluent recycle stream which is passed back to the hydrotreating zone.

" INTEGRATED HYDROTREATING AND HYDROCRACKING PROCESS "

ABSTRACT

A hydrocarbonaceous feedstock is converted in an integrated hydrotreating and hydrocracking process by combining with a hot hydrocracking zone effluent recycle stream containing hydrogen, and the mixture is passed to a denitrification and desulfurization reaction zone to produce hydrogen sulfide and ammonia to thereby clean up the fresh feedstock. The resulting hot, uncooled effluent from the denitrification and desulfurization zone is hydrogen stripped in a stripping zone maintained at essentially the same pressure as the hydrotreating zone with a hydrogen-rich gaseous stream to produce a vapor stream comprising hydrogen, hydrocarbonaceous compounds boiling at a temperature below the boiling range of the fresh feedstock, hydrogen sulfide and ammonia, and a liquid hydrocarbonaceous stream wherein at least a portion of this liquid stream is then passed to a hydrocracking step to produce the hot hydrocracking effluent recycle stream which is passed back to the hydrotreating zone.

" INTEGRATED HYDROTREATING AND HYDROCRACKING PROCESS "

BACKGROUND

Petroleum refiners often produce desirable products such as turbine fuel, diesel fuel and other products known as middle distillates as well as lower boiling hydrocarbonaceous liquids such as naphtha and gasoline by hydrocracking a hydrocarbon feedstock derived from crude oil, for example. Feedstocks most often subjected to hydrocracking are gas oils and heavy gas oils recovered from crude oil by distillation. A typical heavy gas oil comprises a substantial portion of hydrocarbon components boiling above 371°C (700°F), usually at least about 50 percent by weight. A typical vacuum gas oil normally has a boiling point range between 315 to 565°C (600 to 1050°F).

Hydrocracking is generally accomplished by contacting in a hydrocracking reaction vessel or zone the gas oil or other feedstock to be treated with a suitable hydrocracking catalyst under conditions of elevated temperature and pressure in the presence of hydrogen so as to yield a product containing a distribution of hydrocarbon products desired by the refiner. The operating conditions and the hydrocracking catalysts within a hydrocracking reactor influence the yield of the hydrocracked products.

Although a wide variety of process flow schemes, operating conditions and catalysts have been used in commercial activities, there is

always a demand for new hydrocracking methods which provide lower costs and higher liquid product yields. It is generally known that enhanced product selectivity can be achieved at lower conversion per pass (60% to 90% conversion of fresh feed) through the catalytic hydrocracking zone. However, it was previously believed that any advantage of operating at below about 60% conversion per pass was negligible or would only see diminishing returns. Low conversion per pass is generally more expensive, however, the present invention greatly improves the economic benefits of a low conversion per pass process and demonstrates the unexpected advantages.

SUMMARY

The present invention is a catalytic hydrocracking process which provides higher liquid product yields, specifically higher yields of turbine fuel and diesel oil. The process of the present invention provides the yield advantages associated with a low conversion per pass operation without compromising unit economics. Other benefits of a low conversion per pass operation include the elimination of the need for inter-bed hydrogen quench and the minimization of the fresh feed pre-heat since the higher flow rate of recycle liquid will provide additional process heat to initiate the catalytic reaction and an additional heat sink to absorb the heat of reaction. An overall reduction in fuel gas and hydrogen consumption and light ends production may also be obtained. Finally, the low conversion per pass operation requires less catalyst volume.

In one embodiment the present invention relates to a process for hydrocracking a hydrocarbonaceous feedstock which process comprises the steps of: (a) passing a hydrocarbonaceous feedstock and hydrogen to a catalytic denitrification and desulfurization reaction zone at reaction conditions including a temperature from 204 to 482°C (400 to 900°F), a pressure from 3.5 to 17.3 mPa (500 to 2500 psig), a liquid hourly space velocity of the hydrocarbonaceous feedstock from 0.1 to 10 hr^{-1}, with a catalyst; and recovering a denitrification and desulfurization reaction zone effluent therefrom; (b) passing the effluent directly to a hot, high pressure stripper utilizing a hot, hydrogen-rich stripping gas to produce a first vapor stream comprising hydrogen, hydrocarbonaceous compounds boiling at a temperature below the boiling range of the hydrocarbonaceous feedstock, hydrogen sulfide and ammonia, and a first liquid stream comprising hydrocarbonaceous compounds boiling in the range of the hydrocarbonaceous feedstock; (c) passing at least a portion of the first liquid stream to a hydrocracking zone containing a hydrocracking catalyst and operating at a temperature of 204 to 482°C (400 to about 900°F), a pressure from 3.5 to 17.3 mPa (500 to 2500 psig), a liquid hourly space velocity from 0.1 to 15 hr^{-1}; and recovering a hydrocracking zone effluent therefrom; (d) passing the hydrocracking zone effluent to the denitrification and desulfurization reaction zone; (e) condensing at least a portion of the first vapor stream recovered in step (b) to produce a second liquid stream comprising hydrocarbonaceous compounds boiling at a temperature below the boiling range of the hydrocarbonaceous feedstock and a second

vapor stream comprising hydrogen and hydrogen sulfide; and (f) recycling at least a portion of the second vapor stream to the hydrocracking zone.

In a second embodiment, the present invention relates to a process for hydrocracking a hydrocarbonaceous feedstock as described above in the first embodiment wherein at least a second portion of the second vapor stream is embodied into a reflux heat exchange zone located in an upper end of the stripper to produce reflux; and the second portion of the second vapor stream is removed from the reflux heat exchange zone and is introduced into a lower end of the stripper to supply stripping medium.

In a third embodiment the present invention relates to a process for hydrocracking a hydrocarbonaceous feedstock as described in the first embodiment wherein at least a portion of the first vapor stream recovered in step (b) is passed to a post-treat hydrogenation reaction zone to saturate aromatic compounds; and at least a portion of the resulting effluent from the post-treat hydrogenation reaction zone is condensed to produce at least a portion of the second liquid stream comprising hydrocarbonaceous compounds boiling at a temperature below the boiling range of the hydrocarbonaceous feedstock and at least a portion of the second vapor stream comprising hydrogen and hydrogen sulfide.

Further in a fourth embodiment the present invention relates to a process for hydrocracking a hydrocarbonaceous feedstock which process comprises the steps of: (a) passing a hydrocarbonaceous feedstock and

hydrogen to a denitrification and desulfurization catalytic reaction zone at reaction zone conditions including a temperature from 204 to 482°C (400 to 900°F), a pressure from 3.5 to 17.3 mPa (500 to 2500 psig), a liquid hourly space velocity of the hydrocarbonaceous feedstock from 0.1 to 10 hr^{-1}, and recovering a denitrification and desulfurization reaction zone effluent therefrom; (b) passing the effluent directly to a hot, high pressure stripper utilizing a hot, hydrogen-rich stripping gas to produce a first vapor stream comprising hydrogen, hydrocarbonaceous compounds boiling at a temperature below the boiling range of the hydrocarbonaceous feedstock, hydrogen sulfide and ammonia, and a first liquid stream comprising hydrocarbonaceous compounds boiling in the range of the hydrocarbonaceous feedstock; (c) passing at least a portion of the first liquid stream to a hydrocracking zone containing a hydrocracking catalyst and operating at a temperature of 204 to 482°C (400 to 900°F), a pressure from 3.5 to 17.3 mPa (500 to 2500 psig), a liquid hourly space velocity from 0.1 to 15 hr^{-1}; and recovering a hydrocracking zone effluent therefrom; (d) passing the hydrocracking zone effluent to the denitrification and desulfurization reaction zone; (e) passing at least a portion of the first vapor stream recovered in step (b) to a post-treat hydrogenation reaction zone to saturate aromatic compounds; (f) condensing at least a portion of the resulting effluent from the post-treat hydrogenation reaction zone to produce a second liquid stream comprising hydrocarbonaceous compounds boiling at a temperature below the boiling range of the hydrocarbonaceous feedstock and a second vapor stream comprising hydrogen and hydrogen sulfide; (g) recycling at

least a first portion of the second vapor stream to the hydrocracking zone; (h) introducing at least a second portion of the second vapor stream into a reflux heat exchanger located in an upper end of the stripper to produce reflux; and (i) removing and heating the second portion of the second vapor stream from the reflux heat exchange zone and introducing the second portion of the second vapor stream into a lower end of the stripper to supply a stripping medium.

BRIEF DESCRIPTION OF THE DRAWING

The drawing is a simplified process flow diagram of a preferred embodiment of the present invention.

DETAILED DESCRIPTION

It has been discovered that higher liquid product yields and a lower cost of production can be achieved and enjoyed in the above-described integrated hydrotreating and hydrocracking process.

The process of the present invention is particularly useful for hydrocracking a hydrocarbon oil containing hydrocarbons and/or other organic materials to produce a product containing hydrocarbons and/or other organic materials of lower average boiling point and lower average molecular weight. The hydrocarbon feedstocks that may be subjected to hydrocracking by the method of the invention include all mineral oils and synthetic oils (e.g., shale oil, tar sand products, etc.) and fractions thereof. Illustrative hydrocarbon feedstocks include those containing components boiling above 288°C (550°F),

such as atmospheric gas oils, vacuum gas oils, deasphalted, vacuum, and atmospheric residua, hydrotreated or mildly hydrocracked residual oils, coker distillates, straight run distillates, solvent-deasphalted oils, pyrolysis-derived oils, high boiling synthetic oils, cycle oils and cat cracker distilllates. A preferred hydrocracking feedstock is a gas oil or other hydrocarbon fraction having at least 50% by weight, and most usually at least 75% by weight, of its components boiling at temperatures above the end point of the desired product, which end point, in the case of heavy gasoline, is generally in the range from 193 to 216°C (380 to 420°F). One of the most preferred gas oil feedstocks will contain hydrocarbon components which boil above 288°C with best results being achieved with feeds containing at least 25 percent by volume of the components boiling between 316 to 538°C (600 and 1000°F) an especially preferred feedstock boils in the range of 232 to 566°C (450 to 1050°F).

Also included are petroleum distillates wherein at least 90 percent of the components boil in the range from 149 to 427°C (300 to 800°F). The petroleum distillates may be treated to produce both light gasoline fractions (boiling range, for example, from 10 to 85°C (50 to 185°F) and heavy gasoline fractions (boiling range, for example, from 85 to 204°C (185 to 400°F). The present invention is particularly suited for maximizing the yield of liquid products including middle distillate products.

The selected feedstock is first introduced into a catalytic denitrification and desulfurization reaction zone together with a hot

hydrocracking zone effluent at hydrotreating reaction conditions. Preferred denitrification and desulfurization reaction conditions or hydrotreating reaction conditions include a temperature from 204 to 482°C (400 to about 900°F), a pressure from 3.5 to 17.3 mPa (500 to 2500 psig), a liquid hourly space velocity of the fresh hydrocarbonaceous feedstock from 0.1 to 10 hr^{-1} with a hydrotreating catalyst or a combination of hydrotreating catalysts.

The term "hydrotreating" as used herein refers to processes wherein a hydrogen-containing treat gas is used in the presence of suitable catalysts which are primarily active for the removal of heteroatoms, such as sulfur and nitrogen and for some hydrogenation of aromatics. Suitable hydrotreating catalysts for use in the present invention are any known conventional hydrotreating catalysts and include those which are comprised of at least one Group VIII metal, preferably iron, cobalt and nickel, more preferably cobalt and/or nickel and at least one Group VI metal, preferably molybdenum and tungsten, on a high surface area support material, preferably alumina. Other suitable hydrotreating catalysts include zeolitic catalysts, as well as noble metal catalysts where the noble metal is selected from palladium and platinum. It is within the scope of the present invention that more than one type of hydrotreating catalyst be used in the same reaction vessel. The Group VIII metal is typically present in an amount ranging from 2 to 20 wt.%, preferably from 4 to 12 wt.%. The Group VI metal will typically be present in an amount ranging from 1 to 25 wt. %, preferably from 2 to 25 wt.%.

The resulting effluent from the denitrification and desulfurization reaction zone is transferred without intentional heat-exchange (uncooled) and is introduced into a hot, high pressure stripping zone maintained at essentially the same pressure as the denitrification and desulfurization reaction zone where it is countercurrently stripped with a hydrogen-rich gaseous stream to produce a first gaseous hydrocarbonaceous stream containing hydrocarbonaceous compounds boiling at a temperature less than 371°C (700°F), hydrogen sulfide and ammonia, and a first liquid hydrocarbonaceous stream containing hydrocarbonaceous compounds boiling at a temperature greater than 371°C. The stripping zone is preferably maintained at a temperature in the range from 232 to 468°C (450 to about 875°F). The effluent from the denitrification and desulfurization reaction zone is not substantially cooled prior to stripping and would only be lower in temperature due to unavoidable heat loss during transport from the reaction zone to the stripping zone. It is preferred that any cooling of the denitrification and desulfurization reaction zone effluent prior to stripping is less than 56°C (100°F). By maintaining the pressure of the stripping zone at essentially the same pressure as the denitrification and desulfurization reaction zone is meant that any difference in pressure is due to the pressure drop required to flow the effluent stream from the reaction zone to the stripping zone. It is preferred that the pressure drop is less than 690 kPa (100 psig). The hydrogen-rich gaseous stream is preferably supplied to the stripping zone in an amount greater than about 1 wt.% of the hydrocarbonaceous feed to this zone.

In one embodiment, the hydrogen-rich gaseous stream used as the stripping

medium in the stripping zone is first introduced into a reflux heat exchange zone located in an upper end of the stripping zone to produce reflux therefor and then introducing the resulting heated hydrogen-rich gaseous stream into a lower end of the stripping zone to perform the stripping function.

At least a portion of the first liquid hydrocarbonaceous stream containing hydrocarbonaceous compounds boiling at a temperature greater than 371°C (700°F) recovered from the stripping zone is introduced directly into a hydrocracking zone along with added hydrogen. The hydrocracking zone may contain one or more beds of the same or different catalyst. In one embodiment, when the preferred products are middle distillates, the preferred hydrocracking catalysts utilize amorphous bases or low-level zeolite bases combined with one or more Group VIII or Group VIB metal hydrogenating components. In another embodiment, when the preferred products are in the gasoline boiling range, the hydrocracking zone contains a catalyst which comprises, in general, any crystalline zeolite cracking base upon which is deposited a minor proportion of a Group VIII metal hydrogenating component. Additional hydrogenating components may be selected from Group VIB for incorporation with the zeolite base. The zeolite cracking bases are sometimes referred to in the art as molecular sieves and are usually composed of silica, alumina and one or more exchangeable cations such as sodium, magnesium, calcium, rare earth metals, etc. They are further characterized by crystal pores of relatively uniform diameter between 4 and 14 Angstroms (10^{-10} meters). It is preferred to employ

zeolites having a relatively high silica/alumina mole ratio between 3 to 12. Suitable zeolites found in nature include, for example, mordenite, stilbite, heulandite, ferrierite, dachiardite, chabazite, erionite and faujasite. Suitable synthetic zeolites include, for example, the B, X, Y and L crystal types, e.g., synthetic faujasite and mordenite. The preferred zeolites are those having crystal pore diameters between 8-12 Angstroms (10^{-10} meters), wherein the silica/alumina mole ratio is 4 to 6. A prime example of a zeolite falling in the preferred group is synthetic Y molecular sieve.

The natural occurring zeolites are normally found in a sodium form, an alkaline earth metal form, or mixed forms. The synthetic zeolites are nearly always prepared first in the sodium form. In any case, for use as a cracking base it is preferred that most or all of the original zeolitic monovalent metals be ion-exchanged with a polyvalent metal and/or with an ammonium salt followed by heating to decompose the ammonium ions associated with the zeolite, leaving in their place hydrogen ions and/or exchange sites which have actually been decationized by further removal of water. Hydrogen or "decationized" Y zeolites of this nature are more particularly described in US-A-3,130,006.

Mixed polyvalent metal-hydrogen zeolites may be prepared by ion-exchanging first with an ammonium salt, then partially back exchanging with a polyvalent metal salt and then calcining. In some cases, as in the case of synthetic mordenite, the hydrogen forms can be prepared by direct acid

treatment of the alkali metal zeolites. The preferred cracking bases are those which are at least 10%, and preferably at least 20%, metal-cation-deficient, based on the initial ion-exchange capacity. A specifically desirable and stable class of zeolites are those wherein at least about 20% of the ion exchange capacity is satisfied by hydrogen ions.

The active metals employed in the preferred hydrocracking catalysts of the present invention as hydrogenation components are those of Group VIII, i.e., iron, cobalt, nickel, ruthenium, rhodium, palladium, osmium, iridium and platinum. In addition to these metals, other promoters may also be employed in conjunction therewith, including the metals of Group VIB, e.g., molybdenum and tungsten. The amount of hydrogenating metal in the catalyst can vary within wide ranges. Broadly speaking, any amount between 0.05% to 30 wt.% may be used. In the case of the noble metals, it is normally preferred to use 0.05 to 2 wt.%. The preferred method for incorporating the hydrogenating metal is to contact the zeolite base material with an aqueous solution of a suitable compound of the desired metal wherein the metal is present in a cationic form. Following addition of the selected hydrogenating metal or metals, the resulting catalyst powder is then filtered, dried, pelleted with added lubricants, binders or the like if desired, and calcined in air at temperatures of, e.g., 371°-648°C (700-1200°F) in order to activate the catalyst and decompose ammonium ions. Alternatively, the zeolite component may first be pelleted, followed by the addition of the hydrogenating component and activation by calcining. The foregoing catalysts may be employed in undiluted

form, or the powdered zeolite catalyst may be mixed and copelleted with other relatively less active catalysts, diluents or binders such as alumina, silica gel, silica-alumina cogels, activated clays and the like in proportions ranging between 5 to 90 wt.%. These diluents may be employed as such or they may contain a minor proportion of an added hydrogenating metal such as a Group VIB and/or Group VIII metal.

Additional metal promoted hydrocracking catalysts may also be utilized in the process of the present invention which comprises, for example, aluminophosphate molecular sieves, crystalline chromosilicates and other crystalline silicates. Crystalline chromosilicates are more fully described in US-A-4,363,718.

The hydrocracking of the hydrocarbonaceous feedstock in contact with a hydrocracking catalyst is conducted in the presence of hydrogen and preferably at hydrocracking reactor conditions which include a temperature from 232°C (450°F) to 468°C (875°F), a pressure from 3.5 to 20.8 mPa (500 to 3000°F psig), a liquid hourly space velocity (LHSV) from 0.1 to 30 hr^{-1}, and a hydrogen circulation rate from 355 to 4441 (2000 to 25,000 std ft^3/ barrel). In accordance with the present invention, the term "substantial conversion to lower boiling products" is meant to connote the conversion of at least 5 vol. % of the fresh feedstock. In a preferred embodiment, the per pass conversion in the

hydrocracking zone is in the range from 15 to 45%. More preferably the per pass conversion is in the range from 20 to 40%.

The resulting first gaseous hydrocarbonaceous stream containing hydrocarbonaceous compounds boiling at a temperature less than 371°C (700°F), hydrogen, hydrogen sulfide and ammonia from the stripping zone is preferably introduced in an all vapor phase into a post-treat hydrogenation reaction zone to hydrogenate at least a portion of the aromatic compounds in order to improve the quality of the middle distillate, particularly the jet fuel. The post-treat hydrogenation reaction zone may be conducted in a downflow, upflow or radial flow mode of operation and may utilize any known hydrogenation catalyst. The effluent from the post-treat hydrogenation reaction zone is preferably cooled to a temperature in the range from 4 to 60°C (40 to 140°F) and at least partially condensed to produce a second liquid hydrocarbonaceous stream which is recovered and fractionated to produce desired hydrocarbon product streams and to produce a second hydrogen-rich gaseous stream which is bifurcated to provide at least a portion of the added hydrogen introduced into the hydrocracking zone as hereinabove described and at least a portion of the first hydrogen-rich gaseous stream introduced in the stripping zone. Fresh make-up hydrogen may be introduced into the process at any suitable and convenient location but is preferably introduced into the stripping zone. Before the second hydrogen-rich gaseous stream is introduced into the hydrocracking zone, it is preferred that at least a significant portion, at least about 90 wt.%, for

example, of the hydrogen sulfide is removed and recovered by means of known, conventional methods. In a preferred embodiment, the hydrogen-rich gaseous stream introduced into the hydrocracking zone contains less than 50 wppm hydrogen sulfide.

DETAILED DESCRIPTION OF THE DRAWING

In the drawing, the process of the present invention is illustrated by means of a simplified schematic flow diagram in which such details as pumps, instrumentation, heat-exchange and heat-recovery circuits, compressors and similar hardware have been deleted as being non-essential to an understanding of the techniques involved.

With reference now to the drawing, a feed stream comprising vacuum gas oil and heavy coker gas oil is introduced into the process via line 1 and admixed with a hereinafter-described effluent from hydrocracking zone 31 transported via line 32. The resulting admixture is transported via line 2 into hydrotreating zone 3. The resulting effluent from hydrotreating zone 3 is transported via line 4 and introduced into stripping zone 5. A vaporous stream containing hydrocarbons and hydrogen passes upward in stripping zone 5 and contacts heat-exchanger 25 and at least a portion thereof is removed from stripping zone 5 via line 7 and introduced into post-treat hydrotreating zone 8. A liquid hydrocarbonaceous stream is removed from stripping zone 5 via line 6 and is introduced into hydrocracking zone 31 via line 6 and line 30. A gaseous

effluent stream is removed from post-treat hydrotreating zone 8 via line 9 and is introduced into heat-exchanger 10. The resulting cooled effluent from heat-exchanger 10 is transported via line 11 and introduced into vapor-liquid separator 12. A hydrogen-rich gaseous stream containing acid gas compounds is removed from vapor-liquid separator 12 via line 17 and is introduced into acid gas recovery zone 18. A lean solvent is introduced via line 35 into acid gas recovery zone 18 and contacts the hydrogen-rich gaseous stream in order to dissolve an acid gas. A rich solvent containing acid gas is removed from acid gas recovery zone 18 via line 36 and recovered. A hydrogen-rich gaseous stream containing a reduced concentration of acid gas is removed from acid gas recovery zone 18 via line 19 and is admixed with fresh make-up hydrogen which is introduced via line 20. The resulting admixture is transported via line 21 and is introduced into compressor 22. A resulting compressed hydrogen-rich gaseous stream is transported via line 23 and at least a portion is recycled via line 29 and line 30 to hydrocracking zone 31. Another portion of the hydrogen-rich gaseous stream is transported via line 24 and is introduced into heat-exchanger 25. The resulting heated hydrogen-rich gaseous stream is removed from heat-exchanger 25 via line 26 and is introduced into heat-exchanger 27. The resulting heated hydrogen-rich gaseous stream is removed from heat-exchanger 27 and transported via line 28 and introduced into stripping zone 5. An aqueous stream is introduced via line 33 and contacts the flowing stream in line 9 and is subsequently introduced into vapor-liquid separator 12 as hereinabove described. An aqueous stream containing water-soluble salts is

removed from vapor-liquid separator 12 via line 34 and recovered. A liquid stream containing hydrocarbonaceous compounds is removed from vapor-liquid separator 12 via line 13, reduced in pressure and introduced into separation zone 14. A gaseous stream containing hydrogen and normally gaseous hydrocarbons is removed from separation zone 14 via line 15. A liquid stream containing hydrocarbons is removed from separation zone 14 via line 16 and recovered.

ILLUSTRATIVE EMBODIMENT

The process of the present invention is further demonstrated by the following illustrative embodiment. All of the following data were not obtained by the actual performance of the present invention but are considered illustrative of the expected performance of the invention.

A portion of a hydrocracker feedstock having the characteristics presented in Table 1 is hydrocracked in a conventional single stage hydrocracker at operating conditions presented in Table 2 to yield the products described in Table 3. Another portion of the same hydrocracker feedstock is hydrocracked in a hydrocracker of the present invention using the same type of catalyst as the base case at operating conditions presented in Table 2 to yield the products described in Table 3. Yields are calculated based on fresh feed at start of run conditions.

TABLE 1 - HYDROCRACKER FEEDSTOCK ANALYSIS

80/20 Blend Straight Run Vacuum Gas Oil-Coker Gas Oil

Gravity, °API	21 (927 kg/m^3)	
Distillation, Volume Percent	°F	°C
IBP	664	(351)
10	716	(379)
30	767	(408)
50	817	(436)
70	880	(471)
90	965	(518)
FBP	1050	(565)
Sulfur, wt. %	3.01	
Nitrogen, PPM	1256	
Bromine Number	7.5	
Heptane Insolubles, wt. %	<0.05	
Conradson Carbon, wt. %	0.36	
Nickel and Vanadium, PPM	0.4	

TABLE 2 – SUMMARY OF OPERATING CONDITIONS

Flowscheme	Base Case	Low Conversion Per Pass with Improved Yields
Reactor Operating Conditions		
High Pressure Separator Pressure	15.96 mPa (2300 psig)	11.82 mPa (1700 psig)
Liquid Hourly Space Velocity, hr^{-1}		
Hydrotreating Zone	2.18	1.13
Hydrocracking Zone	0.93	3.0
Overall	0.65	0.82
Combined Feed Ratio	**1.5	***3.0
H_2/Fresh Feed	1954 std m^3/m^3 (11,000 SCFB)	1954 std m^3/m^3 (11,000 SCFB)
Conversion, Per Pass*, %	60	30
Total (Gross) Conversion, %*	100	100
Number of Gas Quench Points	3	0
Maximum Reactor ΔT HT/HC	27.8/16.7°C (50/30°F)	33.3/27.8 °C (60/50°F)

* Conversion to 382°C (720°F) End Point Distillate and Lighter
** Recycle Liquid to HT first then to HC
*** Recycle Liquid to HC first then to HT

TABLE 3 - PRODUCT YIELDS

	Base Case		Invention	
	Wt. %	Vol. %	Wt. %	Vol. %
NH_3	0.15		0.15	
H_2S	3.20		3.20	
C_1-C_4	3.68		2.97	
Light Naphtha (C_5-C_6)	6.32	8.76	5.08	7.04
Heavy Naphtha C_7-127°C (260°F)	10.38	12.87	7.68	9.52
Kerosine 127-288°C (260-550°F)	50.16	58.15	48.34	55.92
Diesel 288-382°C (550-720°F)	28.72	31.98	35.11	39.09
Total Middle Distillate	78.88	90.13	83.45	95.01
C_{5+} Total	95.58	111.76	96.21	111.57
C_{4+} Total	98.20	116.01	98.32	115.00
Chemical H_2 Consumption (SCFB)	2.61	1600	2.53	1550

From the above tables it is apparent that the present invention is able to operate at a pressure of 11.8 mPa (1700 psig) or approximately one fourth less than the base case, utilizes a hydrocracking reactor having about 30% less internal volume as well as about 20% less catalyst inventory. Because of the lower hydrocracking reactor zone operating severity in the

present invention, the conversion per pass is reduced from 60% to 30%. These enumerated changes used in the present invention provide a lower cost hydrocracking process as well as providing an increased yield of total middle distillate product. The present invention also has a 8.89 std m^3/m^3 (50 SCFB) lower chemical hydrogen consumption and a 50% less hydrogen loss to fuel gas.

CLAIMS:

1. A process for hydrocracking a hydrocarbonaceous feedstock which process comprises the steps of:

(a) passing the hydrocarbonaceous feedstock (1) and hydrogen to a denitrification and desulfurization catalytic reaction zone (3) at reaction conditions including a temperature from 204 to 482°C (400 to 900°F), a pressure from 3.5 to 17.3 mPa (500 to 2500 psig), a liquid hourly space velocity of said hydrocarbonaceous feedstock from 0.1 to 10 hr^{-1} and recovering a denitrification and desulfurization reaction zone effluent (4) therefrom;

(b) passing the effluent (4) directly to a hot, high pressure stripper (5) utilizing a hot, hydrogen-rich stripping gas (28) to produce a first vapor stream (7) comprising hydrogen; hydrocarbonaceous compounds boiling at a temperature below the boiling range of said hydrocarbonaceous feedstock; hydrogen sulfide and ammonia; and a first liquid stream (6) comprising hydrocarbonaceous compounds boiling in the range of said hydrocarbonaceous feedstock;

(c) passing at least a portion of said first liquid stream (6) to a hydrocracking zone (31) containing a hydrocracking catalyst and operating at a temperature of 204 to 482°C (400 to 900°F), a pressure from 3.5 to 17.3 mPa (500 to 2500 psig), a liquid hourly

space velocity from 0.1 to 15 hr^{-1}; and recovering a hydrocracking zone effluent (32) therefrom;

(d) passing said hydrocracking zone effluent (32) to said denitrification and desulfurization reaction zone (3);

(e) condensing (10) at least a portion of said first vapor stream (7) recovered in step (b) to produce a second liquid stream (34) comprising hydrocarbonaceous compounds boiling at a temperature below the boiling range of said hydrocarbonaceous feedstock and a second vapor stream (17) comprising hydrogen and hydrogen sulfide; and

(f) recycling at least a portion of said second vapor stream (17) to said hydrocracking zone (31).

2. The process of Claim 1 wherein said second vapor stream (17) comprising hydrogen and hydrogen sulfide is treated (18) to remove at least a portion of said hydrogen sulfide.

3. The process of Claim 2 wherein the resulting hydrogen-rich gaseous stream contains less than 50 wppm hydrogen sulfide.

4. The process of Claim 1 wherein said hydrocarbonaceous feedstock boils in the range from 232 to 566°C (450 to 1050°F.)

5. The process of Claim 1 wherein said hot, high pressure stripper (5) is operated at a temperature and pressure which is essentially equal to that of said denitrification and desulfurization reaction zone effluent (4).

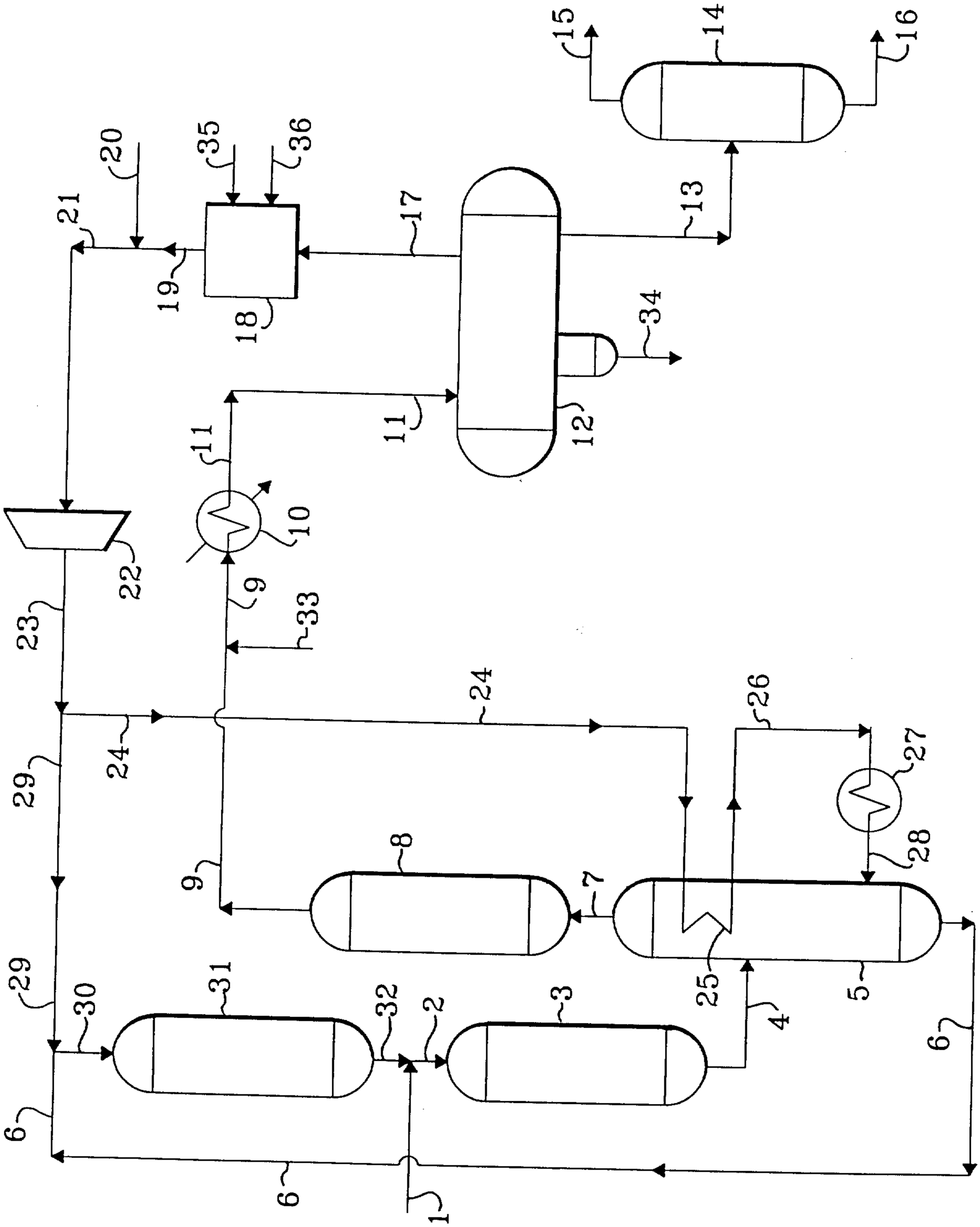
6. The process of Claim 1 wherein at least a portion of said second vapor stream (17) comprising hydrogen and hydrogen sulfide recovered in step (e) is utilized as stripping gas (28) in said hot, high pressure stripper (5).

7. The process of Claim 1 wherein said hydrocracking zone is operated without intermediate hydrogen gas quench points.

8. The process of Claim 1 wherein at least a second portion of said second vapor stream (17) is introduced into a reflux heat exchange zone (25) located in an upper end of said stripper (5) to produce reflux; and said second portion of said second vapor stream is removed from said reflux heat exchange zone (5), heated (27) and introduced (28) into a lower end of said stripper (5) to supply stripping medium.

9. The process of Claim 1 wherein at least a portion of said first vapor stream (7) recovered in step (b) is passed to a post-treat hydrogenation reaction zone (8) to saturate aromatic compounds; and at least a portion of the resulting effluent (9) from said post-treat hydrogenation reaction zone (8) is condensed to produce at least a portion of the second liquid stream comprising hydrocarbonaceous compounds boiling at a temperature below the boiling range of said hydrocarbonaceous feedstock and at least a portion of the second vapor stream comprising hydrogen and hydrogen sulfide;

10. The process of Claim 9 wherein at least a portion of said vapor stream is utilized as stripping gas in said hot, high pressure stripper.


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