

[54] **SIMULTANEOUS HYDROFINING OF
COKER GAS OIL, VACUUM GAS OILS
AND VIRGIN KEROSENE**

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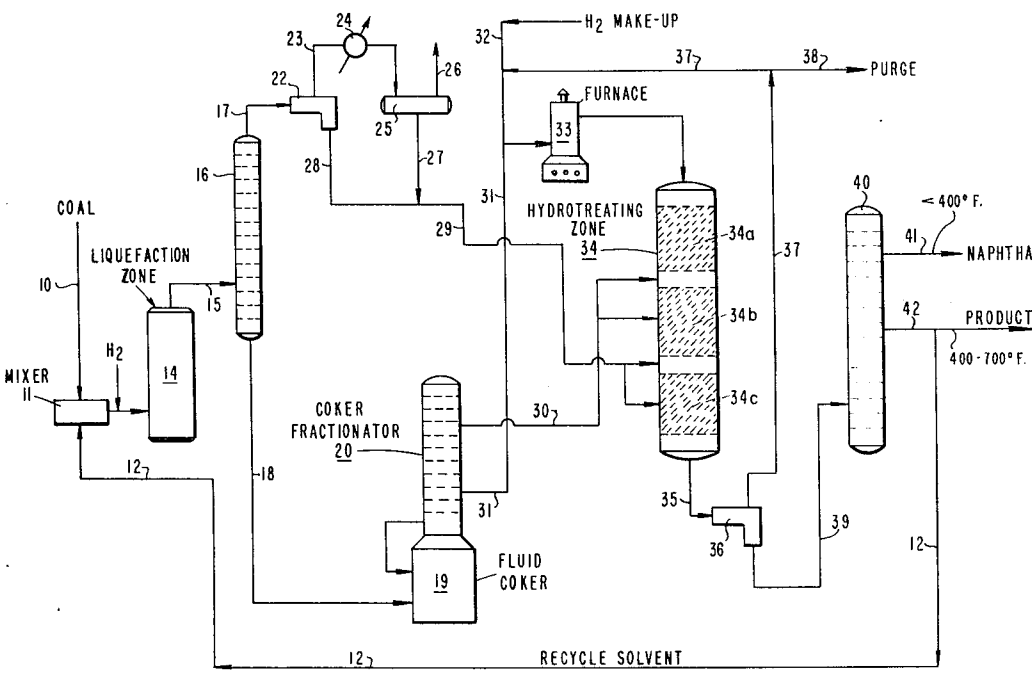
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[57] **ABSTRACT**

A plurality of feedstreams are simultaneously hydrofined in a single reactor which provides four hydrofining zones. A coker gas oil is passed into the first hydrofining zone; vacuum gas oil quenches the hydrofined effluent from the first and second hydrofining zones, and virgin kerosene quenches the hydrofined effluent from the third hydrofining zone. Desulfurized and denitrogenated light ends, jet fuel and catalytic cracker feed are recovered.

3 Claims, 1 Drawing Figure



SIMULTANEOUS HYDROFINING OF COKER GAS OIL, VACUUM GAS OILS AND VIRGIN KEROSENE

BACKGROUND OF THE INVENTION

1. Field of the Invention

This invention is directed to the catalytic hydrotreatment of hydrocarbonaceous feedstreams suitably derived from petroleum crude oils, shale oils, tar sand oils, liquefied coal products and the like. More particularly, it involves a simultaneous hydrogenation of hydrofining hydrotreatment in a single reactor of a plurality of hydrocarbonaceous feedstreams according to a pattern of introduction which selectively hydrogenates of hydrofines each feedstream and which at least partially controls the exothermic heat release created by the hydrogenation or hydrofining reactions.

2. Prior Processes

In producing fuel products either from petroleum, coal, shale, or tar sand sources, various hydrocarbonaceous streams differing in hydrogen-deficiencies, boiling points, and sulfur and nitrogen contents are developed at diverse stages in processing. These are streams which must be hydrotreated to upgrade hydrogen content and/or to hydrodenitrogenate and hydrodesulfurize them, either to produce satisfactory product, on the one hand, or streams suitable for further processing or other uses, on the other. Heretofore hydrotreating of different hydrocarbonaceous streams has been accomplished by combining one or more streams into a single stream and nonselectively hydrotreating that stream in a single reactor, or else by individually hydrotreating uncombined streams in reactors either dedicated to them or blocked to process only one stream at a time. Hydrotreating reactions are, of course, exothermic and necessitate some means of temperature control, especially where large hydrogen addition is required in order to produce suitable effluent streams. Various means have been suggested for this temperature control, including injection of cold recycle treat gas at high rates, a common but very expensive operation.

SUMMARY OF THE INVENTION

We have now found that the inefficiencies, duplications and consequent high cost of refining processes heretofore used to hydrotreat different hydrocarbonaceous feedstocks are overcome by our discovery of a new process for simultaneously but selectively hydrotreating at least two hydrocarbonaceous feedstocks of different compositions in a single hydrotreating unit. Our process involves the injection of differently constituted feedstocks at various points in the single reactor according to a predetermined pattern of sequence and spacing whereby a desired rate of hydrotreatment is effected for each feedstock with a minimum use of catalyst, simultaneously making use of the multiple point injection technique as a method for removing exothermic heat produced in the reactor. This process is applicable both to hydrogenation and hydrofining hydrotreatments.

More particularly, the present process comprises the steps of introducing a first hydrocarbonaceous feedstock and an excess of hydrogen into the inlet end of a first hydrotreating zone in a reactor unit at predetermined rates and passing the feedstock and hydrogen in admixture through a first hydrotreating zone in contact

with a suitable hydrotreating catalyst under suitable hydrotreating conditions to obtain a first zone effluent. The temperature of the admixture rises as it travels through the first zone absorbing heat produced by the exothermic hydrotreating reaction. A second hydrocarbonaceous feedstock of different composition than said first feedstock is introduced into said reactor unit between said first hydrotreating zone and a second hydrotreating zone and admixed with the total effluent from the first zone. The rate of feed and temperature of the second feedstock are sufficient to reduce the temperature of the first zone effluent to within a predetermined range, and preferably are sufficient to quench an amount of heat in the first zone effluent substantially equal to the heat generated by the hydrotreating reactions in the first zone. The admixture of first zone effluent and second feedstock is then passed through a second hydrotreating zone in contact with a suitable hydrotreating catalyst under suitable hydrotreating conditions. The feed rates of said first and second feedstocks are correlated with the hydrotreating conditions in the first and second hydrotreating zones to produce a predetermined hydrogen uptake by the first feedstock in the first and second hydrotreating zones and by the second feedstock in the second hydrotreating zone. Thereafter the total effluent from the hydrotreating reactor unit is removed, and hydrotreated fractions are recovered from the effluent. As a result of this hydrotreatment process, the hydrogen uptake of a fraction corresponding to the boiling range of the first feedstock is greater, and that of the fraction corresponding to the boiling range of the second feedstock is lesser, than if both the first and second feedstocks had been introduced together into the inlet end of the first hydrotreating zone and passed in admixture through both the first and second hydrotreating zones.

By this method, one can thus impart the optimum desired level of hydrogen uptake to each different feedstock introduced into the reactor. In addition, the heat quench provided by the subsequently injected feedstock or feedstocks reduces or, in preferred cases, eliminates the need for auxiliary cooling measures to maintain the second or downstream zone or zones within desired temperature ranges.

The first aspect of this process, i.e., the patterned introduction of feedstocks to the hydrotreating reactor so as to adjust as desired the level of hydrogen uptake to each stream, will best be understood from the following cases involving hydrogenation of a liquefied coal stream.

CONVENTIONAL CASE: SINGLE PASS HYDROGENATION

A coal liquid boiling over the range from about 400° F. to about 700° F. and a hydrogen content of 8.84 weight percent is fed at the rate of 100 pounds per hour to the inlet of a hydrogenation reactor where it is hydrogenated in the presence of a cobalt molybdate catalyst (1-10 weight percent cobalt oxide and 5-40 weight percent molybdenum oxide) on an alumina support under hydrogenation conditions including a temperature of about 700° F., a pressure of about 1,300 psig, a hydrogen treat rate of about 5,000 SCF/B of total feed and a liquid hourly space velocity of about 1.0 w/hr/w. Total hydrogen uptake is 800 SCF/B of

total feed. Exothermic heat generated by the hydrogenation reaction is controlled by cold treat gas recycle. The effluent recovered from the hydrogenation reactor has a hydrogen content of 9.96 weight percent. On fractionation with a cut point of 530° F., the 400°–530° F. fraction has 9.37 weight percent hydrogen and 530°–700° F. fraction has a hydrogen content of 10.68 weight percent.

SPLIT FEED QUENCH CASES

In one case a heavy 530°–700° F. coal liquid cut having a hydrogen content of 8.29 weight percent is fed to the reactor inlet (top). The hydrogenation reactor is operated under conditions like those in the single pass hydrogenation case, and a light 400°–530° F. coal liquid fraction having a hydrogen content of 9.45 weight percent is fed to the reactor as quench after 50 percent of the total required hydrogen uptake of 800 SCF/B has occurred. In a variation of Case One, the light cut is fed to the top of the reactor and the heavy cut used as quench.

In a second case, the heavy and light coal liquid fractions are fed to the reactor as in Case One and its variations, except that the feed used as quench is fed to the reactor at the point where 25 percent of the required 800 SCF H₂ uptake/B of total feed occurs.

Based on experimental rate equations for the total feed, the light cut alone and the heavy cut alone, the following results occur:

SELECTIVE DISTRIBUTION OF HYDROGEN UPTAKE IN LIQUID PRODUCT

	Conventional Case Total Feed at Top	Case 1 Heavy at Top	Case 2 Light at Top	Case 2 Heavy at Top	Light at Top
Heavy Cut, Wt. % H ₂	9.37	9.63	9.13	9.48	9.27
Light Cut, Wt. % H ₂	10.68	10.33	10.90	10.53	10.76

The foregoing cases are quite obviously simplifications; in an actual design, the total feedstock is not fed to the reactor at a single point, but at various points to the reactor consistent with heat balance requirements. However, the examples are directionally indicative of the type of results which can be expected to occur, and they do illustrate the first aspect of the invention, that, for a given level of hydrogen consumption, distribution of the hydrogen in the liquid product recovered from the reactor can be changed significantly by adjusting the sequence and spacing pattern in which the feeds are introduced at different points along the reactor. To give an example, where the 400/700° F. coal liquid used in these cases is to be employed as a hydrogen-donor coal solvent, it is preferable to minimize hydrogen uptake in the heavy cut without over-hydrogenating the light cut (which would reduce the preferred tetrahydronaphthalene molecules to decahydronaphthalene molecules) in order to maximize the total hydrogen donor activity of the solvent. This requires the first stream, which is fed to the top of the reactor, to be the heavy cut, and the second stream, which is injected as quench, to be the light cut. However, if, unlike the coal liquid used in these cases, the initial hydrogen content were very low (say 7.5 weight percent), it would be preferred to maximize

tetrahydronaphthalene production for the total hydrogen uptake; accordingly, the light cut would be fed to the top of the reactor, and the heavy cut would be injected for quench. Similarly, when hydrodesulfurization and/or hydrodenitrogenation is to be effected, the feedstock having the greater reducible sulfur content and/or the greater organic nitrogen content will normally be introduced to the top of the reactor and a feedstock of lesser such content will be used as quench.

The first and second aspects of the case are illustrated by the following example, in which a coal liquid similar to that of the foregoing cases is hydrogenated without requiring supplemental quench: For a total hydrogen consumption of about 850 SCF/B of total feed, with hydrogenation conditions including a temperature of about 700° F. and a pressure of about 1,500 psig for a cobalt molybdate catalyst volume of 30,000 cubic feed equally distributed in four beds, a heavy coal liquid fraction (560/700° F.) having a hydrogen content of 8.61 percent is introduced in admixture with an excess of hydrogen into the top of the reactor at the rate of 72.2 thousand barrels per day. A light coal liquid fraction (400/560° F.) having a hydrogen content of 9.2 weight percent is fed into the reactor at a temperature within the range from 400° F. to 450° F. at the rate of 43.9 thousand barrels per day at points in the reactor where respectively about 25 percent, 50 percent and 75 percent of the total hydrogen consumption will have occurred, admixing at each such point with the effluent respectively from the first, second and third beds, absorbing sufficient heat from these respective effluents to maintain the temperature in those respective beds within a predetermined range around 700° F. so that no supplemental quench gas or other coolant is required for that purpose. The 400/700° F. liquid product recovered from the reactor contains 9.5 weight percent hydrogen. Liquid fractions boiling up to 400° F. contain 11.9 weight percent hydrogen.

For both aspects of the invention, the sequence and spacing pattern according to which different hydrocarbonaceous feedstocks are introduced into a hydrotreating reactor, the temperatures and rates at which these streams are introduced, the catalyst utilization of each of the streams, and other variables of the invention depend, of course, upon the kinetics of the particular hydrotreating reactions involved. Accordingly, the invention is advantageously illustrated by reference to a specific preferred embodiment and by further examples.

DESCRIPTION OF THE DRAWING

The drawing is a schematic flow diagram of a coal liquefaction system in which a coal liquid is hydrogenated in accordance with this invention, the hydrogenated product being recycled for use as a coal solvent.

DESCRIPTION OF THE PREFERRED EMBODIMENT

Referring to FIG. 1, particulate coal is introduced by way of line 10 to a mixing zone 11 in which it is combined to form a slurry with a recycle stream of oil solvent introduced into the mixing zone by way of line 12. The solvent/coal ratio in the mixing zone may suitably

be about 1.2 parts by weight of solvent per part of coal. The slurry from the mixing zone 11 is conducted by way of line 13 into a liquefaction zone 14 which is maintained under liquefaction conditions including a temperature within the range from about 700° F. to about 1,000° F., and a pressure from about 350 psig to about 3,000 psig. Hydrogen may also be introduced in the gaseous form by way of line 13 if desired, although it is not strictly necessary. Within the liquefaction zone 14, hydrogen is transferred from the hydrogen-donor solvent to the coal, causing a depolymerization of the coal along with a solvation effect of the solvent. As a result, there is discharged from the liquefaction zone 14 a mixture of undepleted hydrogen-donor solvent, depleted hydrogen-donor solvent, dissolved coal, undissolved coal and mineral matter. This mixture is removed from liquefaction zone 14 by line 15 and conducted to a distillation tower 16 where it is fractionated. Portions boiling below a selected boiling point, for example, below about 500° F. are recovered overhead by way of line 17, while portions boiling above that boiling point are removed from the fractionator by way of line 18. The higher boiling mixture recovered by way of line 18 is transported to a fluid coker 19 which is preferably operated with a dense phase bed of coke particles maintained in fluidized state by steam and by evolution of vapor volatilization and cracking of the feedstream, as known in the art. Within the fluid coker 19, the liquid hydrocarbons undergo thermal cracking to produce vaporous products which pass upwardly usually through a cyclone separator (not illustrated), which returns coke particles back to the fluidized bed and permits the vaporous particles to ascend upwardly into a coker fractionator 20 which is suitably mounted atop the fluid coker. In coker fractionator 20, the vaporous products are liquefied and distilled according to boiling point. A bottoms product such as hydrocarbons boiling above about 1,000° F. is suitably withdrawn by way of line 21 and recycled to the fluid coker 19 for further conversion.

From liquefaction reaction fractionator 16, the lower boiling portion recovered overhead by way of line 17 is conducted to a hot separator 22 where vaporous products are recovered overhead and carried by line 23 through a heat exchanger 24 where they are cooled to the liquid phase. The liquid product from heat exchanger 24 is then introduced into a cool separator 25, where hydrocarbons still in the vaporous form are recovered by gas line 26. The liquid hydrocarbons in cool separator 25 are withdrawn by way of line 27 and combined with the liquid hydrocarbons withdrawn by way of line 28 from hot separator 22. The combined liquid hydrocarbons from lines 27 and 28 boil over the range from about 100° F. to about 500° F. and have a hydrogen content on the order of 8.70 weight percent.

Coker fractionator 20 is operated to recover a number of streams of different boiling points. For the sake of illustration, only two recovery streams are illustrated, a medium cut stream recovered by way of line 30 which boils over the range from about 350° F. to about 560° F., having a hydrogen content of 8.37 weight percent, and a heavy cut stream, recovered by way of line 31, which has a boiling range of from about 560° F. to about 700° F. and a hydrogen content of 7.78 weight percent. Because a 400/700° F. cut from

streams 29, 30 and 31 is needed for solvent recycle stream 12, and because the hydrogen content of these streams is lower than the 9-10 weight percent hydrogen desired for a solvent recycle stream, the hydrogen content of the streams is upgraded. For a given hydrogen consumption rate specification, in this case about 850 SCF/B of total feed, it is desirable to add the hydrogen to the most hydrogen deficient of the streams 29, 30 and 31 in the desired boiling range, here, the heavy coker fractionator stream in line 31. Accordingly, pursuant to this invention, the stream in line 31 is suitably admixed with an excess of hydrogen introduced by way of line 32, heated in a furnace 33, and then introduced at the rate of 19,090 barrels per day into the inlet end of a mixed gas and liquid phase hydrogenation reactor unit 34 for sequential passage through hydrogenation zones 34a, 34b, and 34c. The hydrogen stream may be separately introduced if desired. In this case, however, the combined streams are introduced at a temperature of about 517° F. and at a rate of 19,090 barrels per day. Streams 30 and 29 are injected in a patterned sequence and spacing into reactor unit 34 and at temperatures and feed rates which provide a desired level of heat quench and hydrogenation in the reactor.

Hydrogenation reactor unit 34 may consist of one or more vessels. For simplicity, reactor unit 34 is illustrated as one large vessel, but in practice it is preferred to utilize a plurality of parallel vessels each of which receives an identical pattern of feed streams. For example, a suitable reactor unit comprises five parallel reactors each about 11.5 feet in diameter and 52 feet in length. The reactor unit contains a suitable hydrogenation catalyst in a series of beds. A hydrogenation zone in a reactor unit may comprise one or more beds. Conventional contacting apparatus is arranged between beds to distribute fluids introduced therein for admixing with effluent fluids from an upstream bed. A suitable arrangement in the aforesaid five-reactor unit comprises five beds, the first two being 6 feet in length and the next three 10 feet in length, a 2-foot long distributing area being between the first and second, second and third, and third and fourth beds. In the drawing, the first bed constitutes the first hydrogenation zone 34a, the second and third beds constitute the second zone 34b, and the fourth and fifth beds make up the third zone 34c. For simplicity, a reactor unit of one vessel with catalyst volume equivalent to the five vessel arrangement is illustrated. Interbed distributing areas are not depicted between the second and third beds and between the fourth and fifth beds, although these will be understood to separate those beds.

Hydrogenation conditions maintained in the reactor include a nondestructive (i.e., noncracking) temperature within the range from about 650° F. to about 850° F., preferably about 700° F., and pressures suitably within the range from about 650 psig to about 2,000 psig, preferably about 1,300 psig. Hydrogen is admixed with stream 31 in sufficient excess to provide a total hydrogen treat rate in reactor 34 within the range from about 1,000 to about 10,000, preferably up to about 5,000, standard cubic feet of hydrogen per barrel of total feed.

After introduction into hydrogenation reactor 34, the feedstock from line 31 passes through the first

hydrogenation zone 34a with hydrogen in contact with the hydrogenation catalyst under the above hydrogenation conditions. The total effluent from first hydrogenation zone 34a then passes from the first hydrogenation zone and admixes with the light coker fractionator stream from line 30 which is introduced into reactor 34 in the area between zones 34a and 34b. The combined streams are then passed through zone 34b in contact with the hydrogenation catalyst under the above-indicated hydrogenation conditions. The light coker fractionator stream from line 30 is preferably introduced into second hydrogenation zone 34b at a plurality of points, for example, between the first and second beds (about 8 feet from the inlet to zone 34a in the parallel reactor arrangement) and between the second and third beds (about 16 feet from such inlet in the parallel arrangement) at feed rates and at a temperature sufficient to reduce the temperature of the effluent from beds one and two to within a predetermined range in zone 34b. Advantageously, the light coker fraction from line 30 is introduced in amounts sufficient without additional heat quench to offset the heat of reaction generated by the exothermic hydrogenation reaction in the hydrogenation zone upstream of the point of introduction. For example, the stream from line 30 is suitably introduced into the reactor at a temperature of about 418° F. at points 8 and 16 feet from the inlet to zone 34a in the parallel reactor configuration in a volume of about 30,750 barrels per day. This rate and temperature is effective to reduce the temperature in the hydrogenation zone at the points of introduction to about 650° F.

The total effluent from hydrogenation zone 34b is then passed from zone 34b and admixed with the raw liquefaction product stream from line 29 which is introduced into reactor 34 between zones 34b and 34c. The raw liquefaction stream is introduced at a temperature of about 500° F. and a volume of 52,100 barrels per day, at points 28 feet and 40 feet from the inlet to zone 34a in the parallel reactor configuration, offsetting the exothermic heat generated in the third bed of the second zone and reducing the temperature in the reactor, where stream 29 and the downflowing effluent mix, to about 650° F. after passing through zone 34c. The liquid hourly space velocity (LHSV) for the line 31 feed stream suitably ranges from about 0.2 wt/hr/wt to about 1.5 wt/hr/wt, preferably about 0.5 wt/hr/wt. The LHSV for the feed stream from line 30 exceeds that of the line 31 feed stream, whereas the LHSV of the line 29 feed stream exceeds that of the line 30 feed stream.

Effluent from the hydrogenation reactor 34 is removed by line 35 and passed to a high pressure gas-liquid separator 36 which exhausts hydrogen gas overhead to line 37 for purge at line 38, makeup with fresh hydrogen, and recycle as treat gas at line 32. The hydrogenated liquid product is carried from the separator 36 by line 39 to fractionator 40 where the naphtha liquids boiling below about 400° F. are split from the higher boiling liquids boiling from about 400° F. to about 700° F. The naphtha liquids, which now have a hydrogen content of 11.96 weight percent, are recovered from fractionator 40 by line 41 for use or for further processing, as desired. The 400/700° F. coal liquid stream, upgraded in hydrogen content to 9.51 weight percent, is recovered from fractionator 40 by

line 42 for further processing, as desired, a sidestream 12 being recycled from line 42 for use as a hydrogen donor solvent for the coal liquefaction process.

In the foregoing and in the succeeding examples, the hydrotreating catalysts employed are of conventional nature. Without being limited to any particular catalyst, these catalysts will typically comprise an alumina or silica-alumina support carrying one or more iron group metals and one or more metals of Group VI-B of the Periodic Table in the form of the oxides or sulfides. In particular, combination of one or more Group VI-B metal oxides of sulfides with one or more Group VIII metal oxides or sulfides are preferred. For example, typical catalyst metal combinations contemplated are oxides and/or sulfides of cobalt-molybdenum, nickel-tungsten, nickel-molybdenum-tungsten, cobalt-nickel-molybdenum, nickel-molybdenum, etc. As a typical example, one catalyst will comprise a high metal-content sulfided cobalt-molybdenum-alumina catalyst containing about 1 to 10 weight percent cobalt oxide and about 5 to 40 weight percent molybdenum oxide, especially about 2 to 5 weight percent cobalt and about 10 to 30 weight percent molybdenum. It will be understood that other oxides and sulfides will be useful, such as those of iron, nickel, chromium, tungsten, etc. The preparation of these catalysts is now well known in the art. The active metals can be added to the relatively inert carrier by impregnation from aqueous solutions followed by drying and calcining to activate the composition. Suitable carriers include, for example, activated alumina, activated alumina-silica, zirconia, titania, etc., and mixtures thereof. Activated clays, such as bauxite, bentonite and montmorillonite, may also be employed.

The embodiment presented above illustrates in detail a balanced process in which the pattern of injection sequence and spacing as well as the temperature and rate of injection are coordinated to selectively hydrogenate lower hydrogen content feed streams while eliminating the need for auxiliary quench gas or other coolant in the hydrogenation reactor. The following examples shown the application of the invention in a case where quench oil injection is sufficient to reduce but not offset the exothermic heat generated in a hydrogenation reactor (Example 1) and wherein various feed streams are combined for hydrodesulfurization and hydrodenitrogenation hydrofining according to this invention (Example 2).

EXAMPLE 1

A plurality of coal liquefaction streams from a coker scrubber, a coker fractionator and a cat cracker fractionator are hydrogenated before being fed to a cat cracker reaction zone, in order to minimize the formation of coke in the cat cracking zone. A 700° F.+ stream from a coker scrubber which has a hydrogen content of 6.66 weight percent is combined with about 7,000 SCF/B of hydrogen treat gas and heated in a furnace to 653° F., after which it is introduced at the rate of 66,100 barrels per day into the inlet of a hydrogenation reactor comprising four hydrogenation zones each containing a bed of cobalt molybdate catalyst. The beds can vary in size, but suitably the first three zones contain beds of about 13,000 cubic feet, while the last zone has a 36,000 cubic feet bed. The hydrogenation

reactor is operated at a temperature of 700° F. and a pressure of 1,800 psig. The cat cracker recycle bottoms stream boiling above 650° F. having a hydrogen content of 6.40 weight percent is introduced into the hydrogenation unit between the first and second hydrogenation zones, at a temperature of about 512° F. and at a daily rate of 31,400 barrels. The coker fractionator stream, boiling over the range from 550° to 700° F. and having a hydrogen content of 7.76 weight percent is introduced between the second and third hydrogenation zones at a rate of 41,950 barrels per day and a temperature of 517° F. Finally, a cat cracker recycle stream boiling over the range from about 450° F. to about 650° F. and having 8.60 weight percent hydrogen is introduced at a temperature of 449° F. into the hydrogenation unit between the third and fourth hydrogenation zones at a rate of 79,000 barrels per day. The effluent from each hydrogenation zone flows into the next downstream zone in admixture with the stream introduced between each zone. Recycle treat gas is added to each bed as needed to maintain desired temperature levels. The liquid hourly space velocity of the coker scrubber stream is lowest, about 0.5 w/hr/w, and that of the hydrocracker fractionator side stream is highest, about 2.0, and the coker fractionator stream is intermediate, from about 0.8 to about 1.2 or so. The hydrogen content of the product stream recovered from the hydrogenation reactor is 9.5 weight percent.

EXAMPLE 2

A coker gas oil and a virgin gas oil, normally upgraded to hydrocracker feed in a cat-feed hydrofiner which uses recycle gas as quench to control highly exothermic reactions in it, are instead hydrotreated in a combination hydrofiner with a virgin kerosene which is upgraded to jet fuel. Four catalyst beds are employed. The first and second upstream beds utilize a nickel molybdate catalyst. The terminal downstream bed utilizes a nickel tungsten catalyst. The third or intermediate bed uses either a nickel molybdate or a nickel tungsten catalyst. The nickel molybdate catalysts are more active than nickel tungsten catalysts for hydrodesulfurization and hydrodenitrogenation. The nickel tungsten catalyst is as active as the nickel molybdate for aromatic reduction and also gives a kerosene smoke point increase which is required. The average temperature in the beds is 700° F., the average pressure is 1,500 psig, the overall space velocity is 0.30 v/hr/v, and the gas treat rate is 2,500 SCF/B, with the treat being 80 percent hydrogen.

The coker gas oil boiling from about 600° F. to about 900° F., after heat exchange with the effluent from the combination hydrofiner, is introduced at a temperature of about 675° F. to the first bed of the combination hydrofiner. Virgin gas oil boiling from about 675° F. to about 1,000° F. is introduced between the second and third beds to quench the effluent from the first and second beds. The effluent from the third bed is quenched with virgin kerosene boiling from about 300° F. to about 500° F. The product issued from the combination hydrofiner, after heat exchange with the coker gas oil and separation of hydrogen gas for recycle to the combination hydrofiner, there is fractionated to recover light ends, jet fuel and an upgraded gas oil fraction which is fed to a cat cracker. The desulfurization

of the gas oil fractions is 95 percent and desulfurization of the virgin kerosene is 99 percent. Denitrogenation of the gas oil fractions is 75 percent; for the virgin kerosene, it is 90 percent. Aromatics reduction in the kerosene fraction is 10 percent, and aromatic ring saturation in the gas oil fractions is 30 percent.

From the foregoing, it can be seen that a plurality of hydrocarbonaceous feed streams of different composition, including streams of various boiling ranges and hydrogen deficiencies, can be simultaneously hydrotreated in a single hydrotreating reactor unit, according to our invention, to optimize the hydrogen treatment desired for the different feedstocks, while making use of the different feedstocks to at least partially quench the effluent from preceding beds. While various modifications will now occur to those skilled in the art from this disclosure of our invention, insofar as those modifications are equivalent ways of performing the invention, they are deemed within the scope of the invention as hereinafter claimed.

We claim:

1. A combined hydrofining process comprising:

feeding a coker gas oil stream boiling from about 600° to 900° F. to the inlet of a combined hydrofining reactor unit at a predetermined temperature and rate with an excess of hydrogen, passing the coker gas oil stream in admixture with said hydrogen through a first hydrofining zone in a first hydrofining zone containing therein a nickel molybdate catalyst under hydrofining conditions to obtain a first zone effluent;

feeding a first vacuum gas oil stream boiling from about 675° F. to about 800° F. to said combined hydrofining reactor unit between said first hydrofining zone and a second hydrofining zone to admix with said first zone effluent, said first vacuum gas oil stream being fed at a temperature and rate of feed calculated to quench an amount of heat in said first zone effluent substantially equal to the heat generated by the hydrofining reactions in said first hydrofining zone;

passing said first zone effluent and first vacuum gas oil mixture through said second hydrofining zone containing therein a nickel molybdate catalyst under hydrofining conditions to obtain a second zone effluent;

feeding a second vacuum gas oil stream boiling from about 800° F. to 1,000° F. to said combined hydrofining reactor unit between said second hydrofining zone and a third hydrofining zone to admix with said second zone effluent, said second vacuum gas oil stream being fed at a temperature and rate of feed calculated to quench an amount of heat in said second zone effluent substantially equal to the heat generated by the hydrofining reactions in said second hydrofining zone;

passing said second zone effluent and second vacuum gas oil mixture through said third hydrofining zone containing therein a nickel catalyst selected from nickel molybdate and nickel tungsten, under hydrofining conditions to obtain a third zone effluent;

feeding a virgin kerosene stream boiling from about 300° F. to 500° F. to said combined hydrofining reactor between said third hydrofining zone and a

fourth hydrofining zone to admix with said third zone effluent, said virgin kerosene stream being fed at a temperature and rate of feed calculated to quench an amount of heat in said third zone effluent substantially equal to the heat generated by the hydrofining reactions in said third hydrofining zone;

passing said third zone effluent and virgin kerosene mixture through said fourth hydrofining zone containing therein a nickel tungsten catalyst, under hydrofining conditions;

the feed rates of said coker gas oil, said first and said second gas oil and said virgin kerosene streams and the hydrofining conditions in said first, second, third and fourth hydrofining zones being correlated to effect a predetermined hydrodesulfurization and hydrodenitrogenation of said streams in said reactor unit;

removing the combined hydrofined effluent from

said fourth hydrofining zone, and recovering light ends, jet fuel and an upgraded catalytic cracker feed from said combined hydrofined effluent.

2. The process of claim 1 wherein said hydrofining conditions and said feed rates are adjusted to provide about 95 percent desulfurization of the vacuum gas oil streams, about 99 percent desulfurization of the virgin kerosene stream, about 75 percent denitrogenation of the vacuum gas oil streams, and about 90 percent denitrogenation of the kerosene stream.

3. The process of claim 1 wherein the hydrofining conditions in said combined reactor comprise a temperature of about 700° F., a pressure of about 1,500 psig, an overall LHSV of about 0.30 v/hr/v and the equivalent of a gas treat rate of 2,500 SCF/B total feed with a treat gas of 80 percent hydrogen.

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