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COMBINATION PROCESS FOR CATALYTIC HYDRODESULFURIZATION
AND REFORMING OF HIGH SULFUR HYDROCARBON MIXTURES

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2 Sheets-Sheet 1

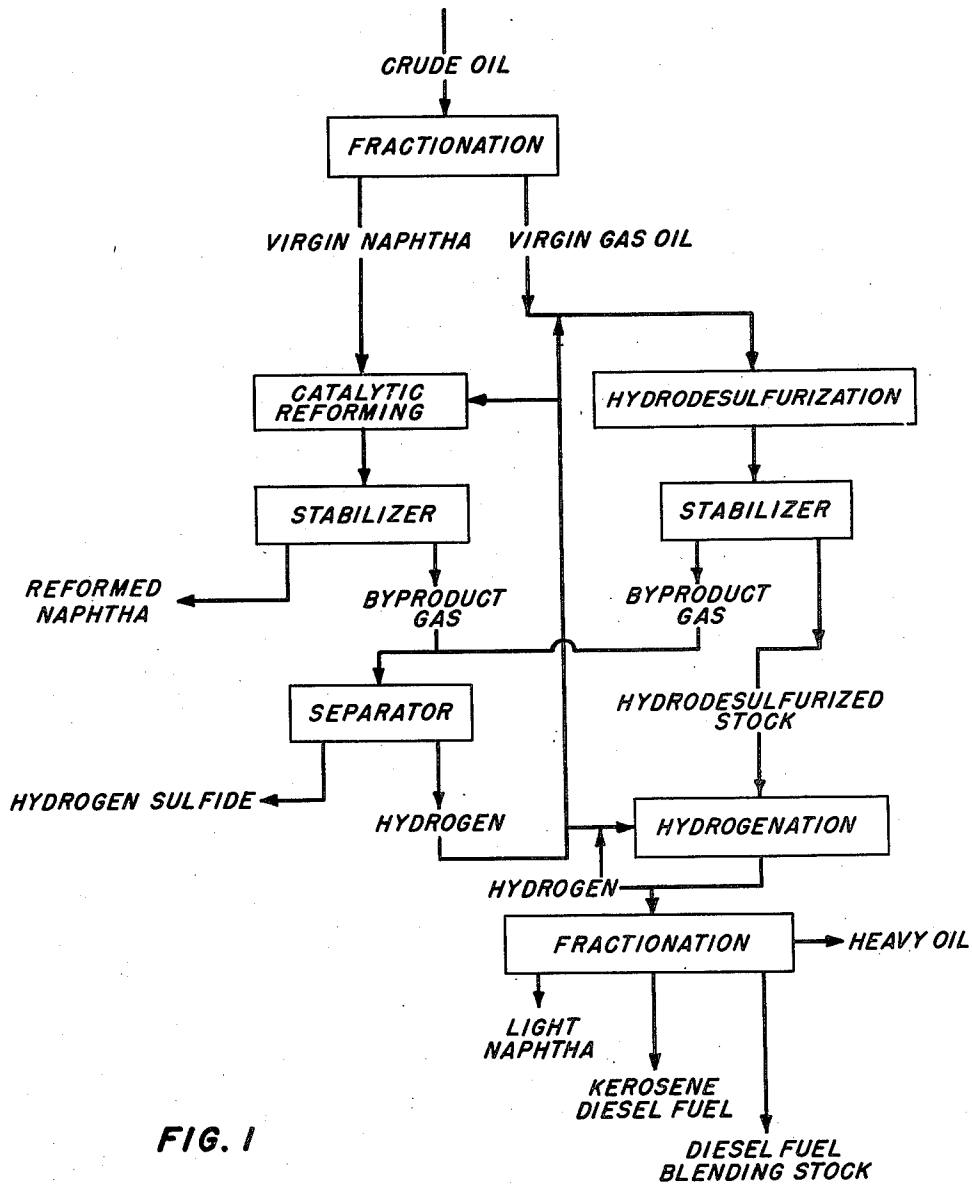


FIG. 1

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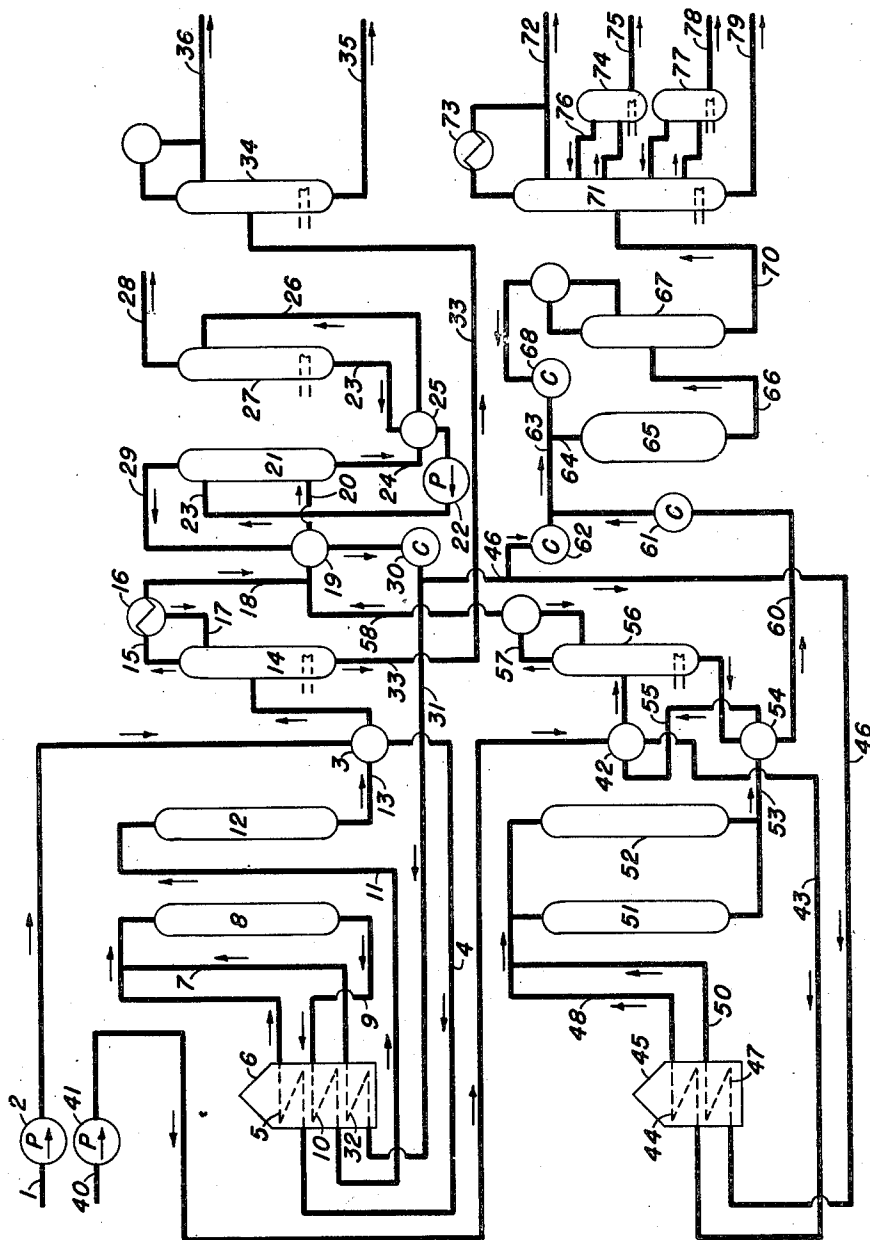


FIG. 2

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1

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COMBINATION PROCESS FOR CATALYTIC HYDRODESULFURIZATION AND REFORMING OF HIGH SULFUR HYDROCARBON MIXTURES

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9 Claims. (Cl. 196—24)

The present invention relates to a combination process of catalytic hydrodesulfurization and reforming of high sulfur hydrocarbon mixtures and, more particularly, a method of treating high sulfur and high aromatic stocks to produce reformed naphtha blending stock, low-aromatic content kerosene and low-sulfur, high cetane number diesel fuel. The results herein are obtained by employing an integrated process of reforming, hydrodesulfurization, and hydrogenation in combination.

The primary object of the invention is to provide a process of producing good quality reformed naphtha, low aromatic kerosene and low sulfur, high cetane number diesel fuel from high sulfur, highly aromatic hydrocarbon mixtures, and crude oils.

Another object is to provide a combination process of reforming, hydrodesulfurization, and hydrogenation to produce good quality products from high sulfur hydrocarbons.

Another object of the invention is to provide an integrated reforming, hydrodesulfurization, and hydrogenation process to accomplish results not attainable by either process alone.

Still another object of the invention is to provide a combination of related catalytic processes which makes possible the adjustment of the severity of one process to the advantage and completeness of the others.

And another object of the invention is to provide a combination process involving the substantial over-all production of hydrogen to accomplish more complete desulfurization and dearomatization to the enhancement of the individual processes and products therefrom.

These and other objects will become apparent as the description thereof proceeds.

In the drawings, Figure 1 is a schematic flow diagram illustrating the generic embodiments of the invention, the relationship of the processes to each other, and to the products produced therefrom.

Figure 2 is a flow diagram illustrating a particular method and apparatus for carrying out the present invention.

In the prior art methods of treating high sulfur crudes to produce desirable products, it has generally been the practice to fractionate the crude to produce naphtha fractions, kerosene fractions, diesel fuel fractions, and bottoms. The naphtha fractions are subjected to thermal or catalytic reforming to produce gasoline fractions. The kerosene fractions from such crudes, because of their high sulfur content, must be subjected to severe desulfurization processes as, for example, extraction with sulfur dioxide. These fractions also contain a high content of aromatic type hydrocarbons which are lost in the extraction process. This loss in aromatics is reflected in a reduction in total yield of kerosene produced from a given amount of crude. Furthermore, the diesel fuel fractions are subjected to separate treatment to increase their cetane number, which may include extraction with sulfur dioxide or the equivalent and thereby further reduce the hydrocarbons available

2

as end products. The present integrated three-step process provides a means for overcoming this loss in yield, and allowing the use of wider fractions of the crude to produce the gasoline blending stocks, kerosenes, and diesel fuel fractions, all having the desirable motor fuel properties, cetane number requirements, and low sulfur content along with decreased aromaticity in the kerosene fractions, without the necessity of further treatment of the individual products. By taking rather wide boiling range fractions of the crude, reforming the lighter fraction rather severely for octane number enhancement, and to form hydrogen for recycling to the hydrodesulfurization and hydrogenation reactions without purification other than removal of hydrogen sulfide, these advantages are further realized and the cost, time, and material savings are increased. Thus, the sulfur and aromatic hydrocarbons that are normally lost by the extraction methods currently in use are converted to new products of increased cetane number, low sulfur content, and improved motor fuel characteristics.

In carrying out the present invention, the hydrocarbon mixture is first fractionated into a naphtha charge stock having a boiling range from about 200° to 400° F. and a heavy fraction having a boiling range from about 350° to 750° F. The light and heavy fractions may come from one crude source or from different crude sources. The naphtha charge stock is preheated by indirect contact with products from the reforming operation and subjected to reforming conditions by first vaporizing under pressure, mixing with hydrogen which has been preheated, and then passing the mixture into the catalytic reforming reactor. The hydrogen is preferably preheated to a higher temperature than the naphtha so that the temperature of the mixture is between about 900° and 1100° F. and preferably about 1000° F. This is accomplished by conducting the hydrogen through the hotter portions of the preheating furnace and the naphtha through the cooler portions of the furnace. The conditions of reforming are adjusted to produce at least the amount of hydrogen that will be needed for the subsequent hydrodesulfurization and hydrogenation reactions. The conditions used during reforming will consequently depend on the naphthenicity and aromaticity of the crudes or hydrocarbon mixtures to be treated or from which the charge stocks are derived.

The reforming reaction is carried out in accordance with known procedures. By "reforming" in accordance with this invention is meant a process of treating a stock at any elevated temperature in the presence of a catalyst capable of promoting dehydrogenation and/or aromatization reactions whereby the straight chain hydrocarbons are aromatized and the naphthenes are dehydrogenated. The reforming process may or may not be carried on in the presence of added hydrogen. There should be a net over-all production of hydrogen during the process which will depend on the conditions used and the type of feed hydrocarbons being treated.

For purposes of this invention, any reforming catalyst may be used. For example, catalysts of the metal oxide or gel type may be used, as an oxide of a metal of the fifth and sixth periodic groups along with refractory metal oxides. A gel type catalyst which is suitable for the reaction comprises 18 to 30 mol percent of chromium oxide and from 82 to 70 mol percent of aluminum oxide, calculated as Cr₂O₃ and Al₂O₃, respectively. Another suitable catalyst is one containing 10 percent by weight of MoO₃ and 90 percent activated alumina or alumina gel. From 5 to 50 mol percent of a metal oxide aromatizing catalyst may be used with the balance being a refractory oxide, such as alumina or the equivalent. The pressure during the reforming reaction may range from 25 to 500 pounds per square inch gauge or higher

3

and the flow rate of reactants through the catalyst bed may be 0.1 to 10 liquid volumes of hydrocarbon per volume of catalyst per hour. The temperature conditions may vary from 800° to 1200° F. as long as the hydrocarbons are in the vapor phase under the conditions imposed.

The preheated hydrogen and hydrocarbons are thoroughly mixed before contacting the reforming catalyst. Since the net reactions are endothermic, there is experienced a fall in temperature in the reaction zone, part of which may be compensated for by preheating the reactants to a higher temperature than reaction temperature. Even under optimum conditions, the catalyst gradually becomes fouled making regeneration necessary. This is accomplished by passing an oxidizing atmosphere through the catalyst at elevated temperatures, preferably at or above reaction temperature. In practice, several reactors may be used so that one or more may be undergoing a regeneration cycle while the others are on stream.

The reformed naphtha or reformed products are next stabilized to remove by-product gases including any hydrogen sulfide formed during the reaction. Stabilization is accomplished by cooling the reformed naphtha to ambient temperature whereby the by-product gaseous hydrocarbons and hydrogen sulfide are separated. The hydrogen sulfide may be separated from the by-product gases by alkali wash or treatment with an amine extractant as in a Girbitol unit, followed by desorption or stripping of the hydrogen sulfide from the amine solution. The recovered hydrogen sulfide is ready for use in preparing sulfur compounds or conversion to free sulfur. The purified by-product gases may be used as such or they may be contacted with absorber oil in a suitable tower to concentrate the hydrogen. In either case, the hydrogen-containing gas is next compressed to reaction pressure and is ready for recycle back to the reforming operation or to subsequent processing operations.

The reformed naphtha is passed to a distillation unit where a small amount of high boiling aromatic material, polymers, etc., are removed under atmospheric pressure. The product is a reformed naphtha gasoline blending stock.

The virgin gas oil having a boiling range of about 350° to 750° F., separated from the hydrocarbon mixture or from a separate source is subjected to hydrodesulfurization. The hydrodesulfurization reaction is carried out in the presence of a hydrodesulfurizing catalyst and added hydrogen at elevated temperatures. In this process, the sulfur compounds in the charge stock are converted into hydrogen sulfide by the action of hydrogen and desulfurization catalysts, such as molybdates, sulfides, and oxides of iron group metals and mixtures including, for example, cobalt molybdate, chromic oxide, vanadium oxide with molybdena and alumina, and sulfides of tungsten, chromium, or uranium. The hydrogen sulfide thus formed is adsorbed by the catalyst and partially or completely reacts with the catalyst to form metallic sulfides. Depending upon the sulfur content of the charge stock, the complete adsorption of hydrogen sulfide by the catalyst continues for one to six hours or for even longer periods of time. The process may be discontinued for regeneration of the catalyst as soon as hydrogen sulfide is evidenced in the product stream. This procedure permits direct use of the hydrogen sulfide-free product gases for recycle purposes. The preferred method of desulfurization involves continuing the operation even after hydrogen sulfide appears in the product gases. In this case, the hydrogen sulfide is removed from the product gases by absorption in ethanalamine or in other suitable solvents. The hydrodesulfurization may be conducted in this manner for many weeks or even months before it is necessary to regenerate the catalyst by oxidation of the carbonaceous deposit. The latter method is preferred for the present purpose; however, it will be understood from the following description that either method may be used.

4

Although any hydrodesulfurizing catalyst may be used, it is preferred to use a cobalt oxide-molybdena-alumina catalyst or a copper-molybdena-alumina catalyst. The process may be carried out in either the liquid or gaseous phase, or mixtures thereof. The temperatures may range from 500° to 800° F. and the pressure from 20 to 1000 pounds per square inch. The gas oil stock or heavy naphtha fraction submitted to this step of the present combination process may contain from 1 to about 7 percent by weight of sulfur existing in the form of various organic sulfur compounds. The charge may be introduced to the catalyst at from 0.5 to 10 liquid volumes per bulk volume of catalyst per hour.

The third step of the present process includes a hydrogenation step and is applied to the hydrodesulfurized product only. The hydrogenation is designed to reduce the aromaticity and unsaturation. The hydrogenation reaction is conducted under hydrogenation conditions with the known hydrogenation catalysts. Desulfurization may be accomplished during this last step to some extent but the principal reactions are hydrogenation of the aromatics and polynuclear ring or straight chain hydrocarbons that are present. The degree of hydrogenation will depend on the conditions employed. The temperatures used during the hydrogenation step may range from 300° to 700° F. with pressures from 500 to 3500 pounds per square inch. The hydrogen flow rate may vary from 500 to 6000 cubic feet of hydrogen per barrel of hydrodesulfurized naphtha. Sufficient residence time should be provided to allow almost complete hydrogenation of the heavy naphtha for best results.

A common recycle system for the hydrogen from all three steps is maintained in accordance with the present invention as shown in Figure 1. The excess hydrogen from the hydrogenation reaction is separated by stabilization and may be recycled back to the hydrogenation reaction or to the reforming and hydrodesulfurization reactions. Also, the hydrogen separated from the reformed products and the hydrodesulfurized products is recycled to one or the other of these reactions or is available for the hydrogenation reaction. Hydrogen separated from the hydrogenated products is recycled back to the hydrogenation reaction without purification.

The process may be illustrated by reference to Figure 2 showing in more detail the flow relationship of the reactants and products. The naphtha charge from the fractionation unit enters at line 1 propelled by pump 2, passes through heat exchanger 3 via line 4 into coil 5 of furnace 6. The preheated naphtha is mixed with recirculated hydrogen (the source of which will be described) from line 7 and passes to head reformer 8. Products from the initial reforming are conducted through line 9 into coil 10 of furnace 6 for reheating in order to complete the reforming reactions because the reactions taking place in reformer 8 are endothermic and the temperature declines. The reheated partially reformed products pass from coil 10 into line 11 and thence to tail reformer 12. In practice, four reformer reactors may be used so that a regeneration cycle may be utilized and the entire reaction made continuous. Two reforming reactors may be on stream while the other two are undergoing regeneration. Coil 32 of furnace 6 is used to preheat the recirculated hydrogen.

It has been found that the conditions of reforming are fixed by the amount of hydrogen that must be produced for the balance of the processes. Also, it is advantageous to preheat the recirculated hydrogen to a temperature which is much higher than the naphtha charge entering the reformer. The arrangement of coils 5, 10, and 32 in furnace 6 within progressively hotter positions in the furnace makes this possible. The hydrogen being in the lower and hotter coil 32 and the partially reformed products being in coil 10 reduce the heat imparted to coil 5 to the point that the control of the preheating of the naphtha charge therein to avoid substantial cracking is

facilitated. The temperature of the mixture of charge naphtha and hydrogen entering reformer 8 is about 1000° F.

The reformed products pass from tail reformer 12 via line 13 into heat exchanger 3 to preheat the incoming charge naphtha and pass into stabilizer 14 at about 300° F. under about 250 p. s. i. g. In stabilizer 14 the pressure is reduced to release the hydrogen and hydrogen sulfide along with any fixed gases which pass via line 15 into condenser 16. Any liquefiable products are returned by line 17. To separate the hydrogen from the hydrogen sulfide and any fixed gas, the vapors at about 100° F. and 240 p. s. i. g. issuing from condenser 16 are passed via line 18, heat exchanger 19, line 20 into the bottom of the first stage of a Girbitol unit.

In the first stage absorber 21 the gases are passed in countercurrent contact with an amine solution or the equivalent which is recirculated by pump 22 via line 23 to the top of the unit. The amine solution washes out the hydrogen sulfide and the spent solution passes via line 24, heat exchanger 25, and line 26 to second stage desorber 27 of the Girbitol unit. In desorber 27 the pressure is reduced to atmospheric and the hydrogen sulfide released from the amine solution by the application of heat. The released hydrogen sulfide is drawn off at line 28 and may be converted to free sulfur.

The purified hydrogen passes overhead from absorber 21 via line 29 through heat exchanger 19, where it cools the vapors from stabilizer 14, and then it is compressed in compressor 30 to about 250 p. s. i. g., pumped through line 31 for preheating in coil 32, as previously described, to complete the hydrogen recycle for the reforming step. The reformed products from stabilizer 14 pass through line 33 to fractionator 34 where a small amount of high boiling aromatic material, polymers, etc., are separated at 35 and a reformed naphtha gasoline is taken off as overhead at 36.

The second stage of the process, comprising the hydrodesulfurization and hydrogenation reactions, is conducted by introducing the heavy naphtha at line 40 propelled by pump 41 through heat exchanger 42, line 43, into coil 44 of furnace 45. In the arrangement shown, one Girbitol unit is used to separate and purify the hydrogen present in the reformed products and in the hydrodesulfurization products. Separate Girbitol units may be employed if necessary. Purified hydrogen from the combined output of the Girbitol unit may be used for treating the heavy naphtha. For this purpose, a part of the hydrogen from compressor 30 is diverted over line 46 to preheating coil 47 of furnace 45. Preheated heavy naphtha at about 375° to 700° F. passes from coil 44 into line 48 to meet the preheated hydrogen at 700° F. from coil 47 and line 50 and the mixture passes at about 250 p. s. i. g. into either reactor 51 or 52 for hydrodesulfurization. One of the hydrodesulfurization reactors may be on stream while the other is undergoing regeneration. The desulfurized products pass by line 53, heat exchanger 54, line 55, heat exchanger 42, into stabilizer 56. In heat exchanger 42, the incoming heavy naphtha is preheated by the hot desulfurized products. In stabilizer 56, which functions like stabilizer 14, the hydrogen and hydrogen sulfide are separated, and sent by lines 57 and 58 through heat exchanger 19 to the Girbitol unit 21. The products from stabilizer 56 pass through heat exchanger 54, line 60, compressor 61, and mix with recycle hydrogen from line 46 after it has been compressed by compressor 62. The mixture passes via lines 63 and 64 to hydrogenation unit 65. The recycle hydrogen is used in the hydrogenation without further purification. The mixture entering the hydrogenation unit 65 is maintained at about 425° F. under 3000 p. s. i. g. with about 5000 cubic feet of hydrogen, as measured under standard conditions, per barrel of desulfurized naphtha. Sufficient residence time in the hydrogenation unit is maintained to provide for substantially

complete hydrogenation of the heavy naphtha and light gas oil.

The hydrogenated product from unit 65 passes through line 66 to stabilizer 67. The separated hydrogen is recompressed by compressor 68 and returned to the reaction, again without further purification. The resultant heavy naphtha product from stabilizer 67 passes through line 70 to fractionator 71, operated under atmospheric pressure, wherein a light naphtha fraction is removed at 72, through the use of reflux condenser 73. Similarly, a kerosene and No. 1 diesel fuel are removed by stripping tower 74 and collected at 75 while light ends are returned via line 76; and a diesel fuel blending stock is taken off by means of a second stripping tower 77 and collected at 78. The bottoms fraction is removed at 79.

The flow system just described necessarily omits mention of the appropriate accessory equipment and control equipment commonly known in the art, for the sake of brevity. In addition the necessary equipment to accomplish the regeneration steps has been omitted, such means being well known and not a part of this invention.

In order to further demonstrate the invention, the following specific example is given. A high sulfur crude is fractionated to give a light virgin naphtha and a heavy naphtha, a gas oil and bottoms fractions. The properties of the crude and the distillation products therefrom are shown in Table I. The crude oil used was a Worland crude from the State of Wyoming.

TABLE I
Charge properties

Fraction	Yield Percent Vol.	S, wt. Percent	API Gravity	Aromatics, Vol. Percent	Naphthenes, Vol. Percent	Paraffins, Vol. Percent
Total crude	100.0	1.70	37.9	-----	-----	-----
IBP-400° F. ¹	39.0	.30	53.2	15	38	47
400°-500° F. ²	24.7	1.01	33.1	26	-----	-----
Gas Oil	23.5	2.32	-----	-----	-----	-----
550° F. Flash bottoms	11.0	3.76	-----	-----	-----	-----

¹ ASTM Motor Method Octane Number, 51.7.

² Splits into 12.4% kerosene (S-0.81%), 12.4% Diesel Fuel (S-1.21%, cetane number, 40).

Considering a crude capacity of 10,000 bbls./day, the gasoline or light naphtha cut will comprise 3,900 bbls./day to be sent to the reformers 8 and 12 (Figure 2) and the heavy naphtha fraction (400-600° F.) will comprise about 2,470 bbls./day which will yield approximately 1,220 bbls./day each of desulfurized and hydrogenated kerosene and diesel fuel. The reforming reaction was carried out at about 960° F. During the reforming operation, the naphthenes and some of the paraffin hydrocarbons are converted to aromatics and the sulfur compounds are reduced to the corresponding hydrocarbons and hydrogen sulfide. The products and material balance from this operation are shown in Table II.

TABLE II
Reforming operation

Fraction	Wt. Percent S	Percent Aromatics	Motor Octane No.	Bbls./ day	Hydrogen Production, 1,000 of S. C. F. D.	Lbs./ day as S
Charge	.30	15	51.7	3,900	1-36	3,090
Gasoline	.005	55	80.0	3,650	4,250	50
H ₂ S	-----	-----	-----	-----	1-36	3,040
Net H ₂ Produced	-----	-----	-----	-----	4,178	-----
Polymers	-----	90	-----	2	-----	-----

¹ H₂ to hydrogenate olefins produced by cracking S compounds and to produce H₂S.

The heavy naphtha charge enters the process through line 40 and is subjected to hydrodesulfurization in reactors 51 and 52 maintained at about 800° F. in the presence

of all or a portion of the hydrogen from the reforming step. During this reaction little or no dehydrogenation occurs and the predominant portion of the olefins produced resulting from cracking of the sulfur compounds are hydrogenated during hydrodesulfurization. The hydrogen separated from the products is purified and may be used to supplement the reforming operation, or for recirculation to the hydrodesulfurization reaction or for use as make-up hydrogen for the subsequent hydrogenation reaction.

The hydrodesulfurized products are pumped under high pressure of about 1,000 p. s. i. g. and under relatively low temperature up to about 550° F. to the hydrogenation reactor 65. In this reaction, any well known hydrogenation catalyst may be used as a nickel or nickel oxide catalyst. The hydrogenation proceeds under conditions such that the content of aromatics in the feed is converted to the corresponding saturated compounds, thus reducing the aromaticity of the kerosene fraction therein while at the same time effecting little or no desulfurization or transformation of the other hydrocarbons present. The recycle of hydrogen from this operation without subsequent purification has been found to facilitate this result to give kerosene fractions of good burning quality and diesel fuel blending stocks having high cetane numbers. The hydrogenated products are distilled for separation of light and heavy fractions and blending stocks for kerosenes and diesel fuels. Table III presents the properties and material balance of the products of this combined hydrodesulfurization-hydrogenation process.

TABLE III
Hydrodesulfurization-hydrogenation operation

Fraction	Percent Aromatics	Percent Sulfur	Bbls./day	Consumption of H ₂		Cetane No.
				1,000 cf S.c.e.d.	Lbs./day as S	
Charge (400-600° F.)	26	1.01	2,470	² 80	7,260	(40)
Kerosene	3	.03	2,490	2,280	220	
Diesel Fuel	2	.02	1,220	910	65	
Light Naphtha ¹	4	.04	1,220	1,280	145	52
Heavy Oil ¹	1	.01	25	30	2	
H ₂ S	5	.05	25	60	8	
Excess H ₂				² 80	7,670	
				1,738		

¹ Tails from kerosene and diesel fuel production. Not essential to operation.

² H₂ to produce H₂S and hydrogenate olefins.

Since the distillation ranges of kerosene and diesel fuel overlap blending must be used to allow continuous production of both of these products. It is particularly pointed out that the production of kerosene and diesel fuel nearly equals the amount of material charged to the process. This represents a material improvement over presently known methods of producing these products. For example, if extraction methods were employed only 1,790 bbls./day of kerosene plus diesel fuel would be produced, assuming optimum conditions under which extraction of 90 percent purity aromatics can be attained at best. The cetane number of the diesel fuel would be raised, by the present invention from 40 to 52, the kerosene is water white, low in sulfur content, of good color, and has excellent burning qualities.

The hydrocarbon mixtures to be treated in accordance with this invention comprise any mixture of hydrocarbons regardless of source or chemical constitution which have a high content of sulfur or sulfur compounds. Such hydrocarbon mixtures generally have correspondingly high aromatic hydrocarbon contents and as such are extremely difficult to treat for the purpose of producing good fuels, naphthas, and kerosenes. The sulfur may be present as elemental sulfur but is generally present as sulfur compounds including hydrogen sulfide, organic sulfides, and disulfides of aromatic, naphthenic, and polycyclic origin.

By high sulfur content is meant those mixtures including crude oils having from 1.0 to 3.0 weight percent of sulfur present. Also included are those mixtures having as high as 5.0 percent of sulfur. Such crude oils or mixtures may have an API gravity ranging from about 20 to 40.

The lighter fractions separated for treatment in accordance with this invention will naturally contain lesser amounts of total sulfur as a result of fractionation. The light fraction which will boil above about 200° F. will have a sulfur content of about 0.1 to 0.4 weight percent although this amount may be above or below this figure. Likewise, the heavy fraction boiling above about 350° F. will have 1.0 percent or more of sulfur. The volume percent of aromatics, naphthenes, and paraffins present in these fractions may vary somewhat from that described in Table I, depending on the source of the crude oil.

An advantage of the present invention is that, by the employment of a common recycle system between the three integrated reactions, the conditions of reforming may be adjusted and maintained to continuously produce good blending naphtha while at the same time, because of the severity of reforming, produce sufficient recycle hydrogen to be used in both the hydrodesulfurization reaction and the hydrogenation reaction.

What is claimed is:

1. The process for improving the motor fuel characteristics and decreasing the sulfur content and aromaticity of fractions obtained from high sulfur content hydrocarbon mixtures which comprises separating said hydrocarbon mixtures into a light fraction having an initial boiling point of about 200° F. and a heavy fraction having an initial boiling point of about 350° F., subjecting said light fraction to catalytic reforming and separating a reformed naphtha blending stock, subjecting said heavy fraction to hydrodesulfurization to produce a hydrodesulfurized product, subjecting said hydrodesulfurized product to hydrogenation at about 300° to 700° F. under conditions to dearomatize same and fractionating the hydrogenated product so produced to yield fractions of improved cetane number and low aromatic content.

2. The process in accordance with claim 1 in which said hydrogenation is carried out at about 300° to 700° F. in the presence of a catalyst selected from the group consisting of nickel and nickel oxide.

3. Process for improving the motor fuel characteristics and decreasing the sulfur content and aromaticity of fractions obtained from high sulfur content hydrocarbon mixtures which comprises separating said hydrocarbon mixtures into a light fraction boiling between about 200° to 400° F. and a heavy fraction boiling between about 350° to 750° F., subjecting said light fraction to catalytic reforming and separating a reformed naphtha blending stock and a hydrogen-rich fraction therefrom, subjecting said heavy fraction to hydrodesulfurization in the presence of said hydrogen-rich fraction and separating a hydrodesulfurized product and a second hydrogen-rich fraction therefrom, subjecting said hydrodesulfurized product to hydrogenation at about 300° to 700° F. under conditions to dearomatize same in the presence of at least a portion of said hydrogen-rich fractions and fractionating the hydrogenated products so produced to yield a light naphtha, a diesel fuel fraction having good engine fuel characteristics and improved cetane number, and a kerosene fraction of low aromatic content.

4. The process in accordance with claim 3 in which the hydrocarbon mixture being treated comprises a crude oil having a total sulfur content of at least about 1.00 weight percent.

5. The process in accordance with claim 4 in which the hydrocarbon mixture is a Worland crude having a total sulfur content of about 1.70 weight percent.

6. The process in accordance with claim 3 in which the hydrocarbon mixture is a high sulfur crude having an API gravity of about 37.9.

7. The process in accordance with claim 1 in which

said hydrogenation is conducted at a temperature of about 550° F. in the presence of a catalyst selected from the group consisting of nickel and nickel oxide.

8. A process for improving the diesel fuel characteristics and decreasing the sulfur content and aromaticity of fractions obtained from a crude oil containing at least about 1.0 percent by weight of sulphur comprising fractionating said crude oil to obtain a light fraction boiling between about 200° to 400° F. and a heavy fraction boiling between about 350° to 750° F., subjecting said light fraction to reforming under severe conditions at a temperature of about 900° to 1100° F. in the presence of hydrogen and separating a reformed naphtha blending stock, subjecting said heavy naphtha to hydrodesulfurization at a temperature from about 500° to 800° F. and separating a hydrodesulfurized product, subjecting said hydrodesulfurized product to hydrogenation at a temperature of about 300° to 700° F. under dearomatizing conditions, and fractionating the products produced from said hydrogenation to produce fractions of improved characteristics.

9. The process of improving the motor fuel characteristics of reformed naphtha, decreasing the sulfur content and aromaticity of light naphtha and kerosene fractions, and increasing the cetane number of diesel fuel fractions obtained from crude hydrocarbon mixtures hav-

ing a total sulfur content of about 1.7 weight percent and an API gravity of about 37.9 comprising separating said mixture into a light fraction having a boiling range between about 200° and 400° F. and a heavy fraction having a boiling range between about 350° and 600° F., subjecting said light fraction to reforming at about 960° F. to produce a reformed naphtha having a sulfur content of about 0.005 weight percent, subjecting said heavy fraction to hydrodesulfurization at about 800° F. to produce a product having a sulfur content of about 0.15 weight percent and subjecting said hydrodesulfurized product to hydrogenation at about 300° to 700° F. under conditions to dearomatize same and fractionating the hydrogenated product to produce a kerosene fraction having about 0.02 percent sulfur by weight and a diesel fuel having a centane number of about 52.

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