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PROCESS FOR MANUFACTURE OF HIGH OCTANE NAPHTHAS

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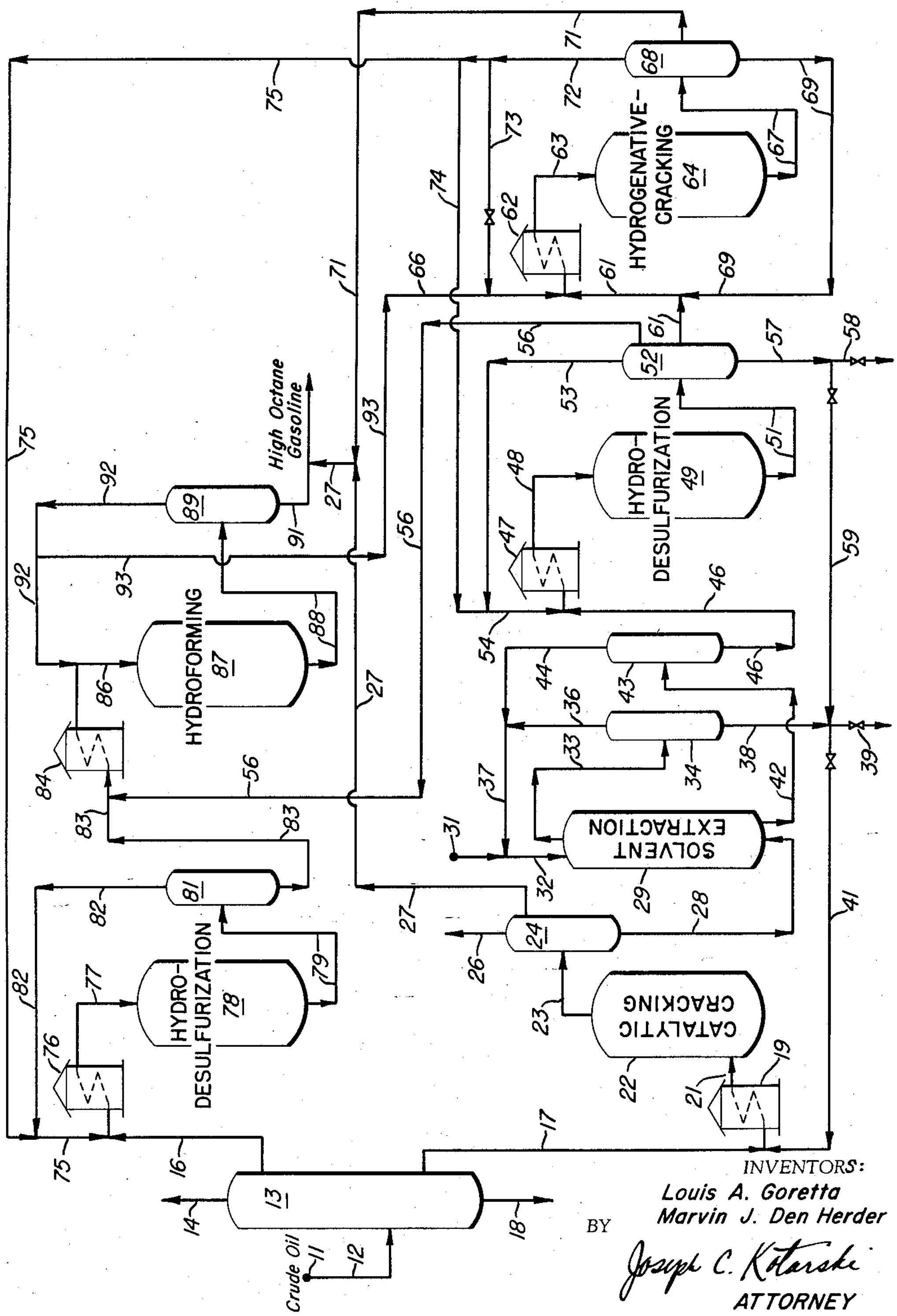

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PROCESS FOR MANUFACTURE OF HIGH OCTANE NAPHTHAS

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This invention relates to a process whereby hydrocarbon oils are catalytically converted to high octane naphthas in high yields.

A problem which continuously confronts petroleum refiners is to produce the highest octane number naphthas for blending into high anti-knock gasoline, and at the same time maximizing the yields of such naphthas. Various processes such as catalytic cracking of gas oils, hydroforming of virgin and cracked naphthas, hydrodesulfurization of oils, and hydrogenation of oils have been proposed or are in actual use for the production of high octane naphtha fractions necessary.

An object of this invention is to provide a combination of refining processes which produces naphthas of maximum octane number in high yields. A further object is to provide a combination of refining processes for converting gas oil to high octane naphtha fractions in substantial yields while reducing the rates of catalyst deactivation and coke deposition thereon that are normally encountered. A further object is to provide an integrated petroleum refining process wherein hydrogen is produced and is used to maximum advantage in reducing the rate of catalyst deactivation and maximizing the octane number of the naphtha fractions produced. Other objects and advantages of the invention will be apparent from the detailed description thereof.

In one aspect of the present invention, a gas oil is catalytically cracked to produce a high octane naphtha and unconverted gas oil. The unconverted gas oil is separated from the high octane naphtha and the former is contacted with a selective solvent which preferentially extracts aromatic hydrocarbons. A hydrocarbon extract phase rich in aromatic hydrocarbons and a hydrocarbon raffinate phase lean in aromatic hydrocarbons is thereby produced. The hydrocarbon extract phase is then hydrodesulfurized by contact with hydrogen and a catalyst at an elevated temperature. Thereafter the hydrodesulfurized extract phase is fractionated into a lower boiling fraction, which usually boils between about 400 and 600° F., and a higher boiling fraction. The lower boiling fraction of the hydrodesulfurized hydrocarbon extract is hydrogenatively cracked by contacting it at a temperature of about 850° to 1000° F. and a pressure of about 1000 to 5000 p.s.i.g. with hydrogen and a dual-functioning catalyst having hydrogenation and cracking properties. A very high octane number naphtha on the order of 100+ (F–1) can thereby be produced. The higher boiling fraction of the hydrodesulfurized hydrocarbon extract makes a superior charge stock to catalytic cracking. A naphtha having an octane number above 90, e.g. 95 F–1 is produced therefrom. By splitting the hydrodesulfurized extract at about the cut point indicated and catalytically cracking the higher boiling portion rather than passing the latter to the hydrogenative cracking step, higher octane products can be produced. Extraction of the unconverted gas oil to produce an aromatics-rich extract phase (a portion of which is subjected to hydrogenative cracking) is essential in obtaining the 100+ (F–1) octane number naphtha. At the same time extraction prior to hydrodesulfurization reduces the hydrogen requirements for hydrodesulfurization. The raffinate from the solvent extraction step may be catalytically cracked to make more high octane naphtha for gasoline blending, or it may be employed as a very superior blending stock for fuel oils.

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Another aspect of the present invention concerns an integrated process wherein crude oil is fractionated and naphtha and gas oil fractions are recovered therefrom. The gas oil fraction is then processed in the manner indicated in the preceding paragraph, i.e. it is catalytically cracked, unconverted gas oil is solvent extracted, the extract phase is hydrodesulfurized and then fractionated to produce a lower boiling fraction (substantially all of which boils between about 400° and 600° F.) and a higher boiling fraction, the higher boiling fraction is catalytically cracked, and the lower boiling fraction is hydrogenatively cracked. The naphtha fraction removed from the crude oil is preferably hydrodesulfurized first and is then catalytically hydroformed to improve its octane number and at the same time produce hydrogen. A hydrogen stream is separated from the products of the catalytic hydroforming step, and the hydrogen stream is employed in the hydrogenative cracking step. A portion of this hydrogen stream from catalytic hydroforming may also be used in the hydrodesulfurization steps. Because the activity of the hydrogenative-cracking catalyst becomes reduced when subjected to large amounts of hydrogen sulfide and nitrogen compounds, it is important in this integrated process to employ the hydrogen produced from the catalytic hydroforming of a hydrodesulfurized naphtha. After the hydrogen passes through the hydrogenative-cracking unit it may advantageously then be employed in the hydrodesulfurization units. The makeup hydrogen to the hydrodesulfurization units (makeup hydrogen is added to compensate for that which is consumed in the hydrodesulfurization reaction) is thus the hydrogen stream which has been recovered from the products of hydrogenative cracking. This latter hydrogen stream has a higher H_2S content than does the hydrogen stream recovered from the products of hydroforming. In this manner the hydrogen sulfide content within the hydrogenative cracking unit is maintained at a low level which is beneficial to the maintenance of high activity of the hydrogenative-cracking catalyst.

Figure 1 shows in diagrammatic form an embodiment of the present invention whereby components of crude oil are converted to high octane naphtha fractions in high yield. Numerous pumps, heaters, and detailed features have been omitted for the purpose of clarity. These omitted features will be apparent to those skilled in the art.

In this embodiment 30,000 barrels/day of crude oil is charged from source **11** by way of line **12** into a crude oil fractitionation system as represented by vessel **13**. From fractionating system **13** fixed gases are removed overhead by way of line **14**; 7,500 barrels/day of a naphtha fraction is removed by way of line **16**; about 12,000 barrels/day of a gas oil fraction is removed by way of line **17**; and a residual oil is removed from the bottom by way of line **18**. The gas oil, which may boil within the range of about 400 or 450° to 900° F. and above and in the embodiment herein boils within the range of 450 to 750° F., is passed by way of line **17** into furnace **19** wherein it is heated to a temperature suitable for catalytic cracking. The heated gas oil is then passed by way of line **21** into a catalytic cracking unit, indicated herein as vessel **22**. The gas oil is catalytically cracked under usual cracking conditions which may comprise a temperature of 850° to 1050° F., a pressure of 5 to 50 p.s.i.g., a catalyst to oil ratio in the range of about 2:1

to 20:1 on a weight basis, a weight space velocity in the range of about 0.2 to 20 pounds of oil per pound of catalyst per hour. A silaceous cracking catalyst such as natural clay, activated natural clay, synthetic catalysts such as silica alumina, silica magnesia, silica alumina zirconia, etc. is used. In the embodiment shown herein, the gas oil is catalytically cracked using a silica alumina catalyst, a temperature of about 950° F. and a pressure of about 25 p.s.i.g., a catalyst to oil weight ratio of about 10, and a space velocity of about 3 pounds of oil per hour per pound of catalyst in the reactor. A conversion of about 50% of the gas oil to lower boiling products, principally high octane naphtha, is obtained. Any of the various types of commercial catalytic cracking processes such as fluidized bed, moving bed, fixed bed, etc. may be employed.

The products from the catalytic cracking step are removed and passed by way of line **23** into fractionating system indicated herein as fractionator **24**. Light gases are removed from fractionator **24** by way of line **26** and passed to a vapor recovery means, not shown herein, for the removal of C_3 and C_4 hydrocarbons. A naphtha fraction is removed from fractionator **24** by way of line **27**. It has an octane number of approximately 95 or greater, the greater octane number being obtained at conversions of gas oil in excess of 50% or thereabouts. Unconverted gas oil, in the amount of about 6,000 barrels/day based upon one-pass operation, is removed from fractionator **24** and passed by way of line **28** to solvent extraction vessel **29**. This unconverted gas oil (commonly called catalytic cycle oil) boils within the range of about 400° to 750 or 800° F. and has an aromatics content in the neighborhood of about 50% and a rather high sulfur content. Its high content of polycyclic aromatic hydrocarbons makes it an undesirable catalytic cracking charge stock since it is difficult to crack, gives low gasoline yields and causes the deposition of large amounts of coke upon the catalyst and leads to more rapid deactivation of the catalyst.

A typical solvent extraction with a solvent which is selective for aromatic hydrocarbons is carried out in extraction vessel **29**. Selective solvents such as liquid SO_2, phenol, cresol, Chlorex, furfural, etc. may be used in amounts of about 25 to 200 volume precent based upon oil in a solvent extraction process employing one or a considerable number of stages. In the embodiment shown herein liquid SO_2 is employed as the selective solvent at an extraction temperature of about −10 to 25° C., e.g. about 15° C. and under sufficient pressure to maintain the SO_2 in the liquid phase. The extraction is carried out in three stages, employing about 50 volume percent of liquid SO_2 based upon oil in each stage. In the schematic diagram shown in Figure 1 liquid SO_2 from source **31** is passed by way of line **32** into the top of extraction vessel **29**. The descending stream of liquid SO_2 passes downwardly through the ascending stream of lighter oil in extraction unit **29** and extracts the aromatic hydrocarbons (as well as considerable amounts of sulfur compounds) from the cycle oil. The raffinate phase which consists primarily of cycle oil, which is now lean in aromatics, together with some occluded SO_2 is removed from the top of extraction vessel **29**. It is then passed by way of line **33** into flash drum **34** wherein SO_2 is vented from the oil. The SO_2 is removed from the top of flash drum **34** and passed by way of line **36** into line **37** by which it is passed to line **32** for recycle to extraction vessel **29**. The hydrocarbon raffinate oil, which may amount to 3,000 barrels/day based upon the first pass of oil through the process, is removed from the bottom of flash drum **34** by way of line **38** and freed of residual SO_2 by equipment not shown. A portion of it may be withdrawn by way of valved line **39** and used for blending in the manufacture of high quality fuel oil. The remainder is passed by way of line **41** to line **17** as a portion of the charge to the catalytic cracking step. This raffinate oil is an excellent charge stock to catalytic cracking. An extract phase consisting of liquid SO_2 containing dissolved cycle oil which is enriched in aromatic hydrocarbons, is removed from the bottom of extraction vessel **29** and passed by way of line **42** into flash drum **43**. SO_2, which is flashed from the extract phase in flash drum **43**, is taken overhead and passed by way of line **44** into line **37** and then is recycled by way of line **32** to extraction vessel **29**. The aromatics-rich hydrocarbon extract phase, which boils between about 400° and 750 to 800° F. and now has an aromatics content of about 80%, is removed from the bottom of flash drum **43**, freed from residual SO_2 and passed by way of line **46** into furnace **47**. The cycle oil extract may amount to about 3,000 barrels/day based upon the initial pass of crude oil on a once-through basis in this embodiment (all figures presented herein concerning amounts of oil charged to various units and hydrogen consumed therein are based upon the initial charge of 30,000 barrels/day of crude oil without compensation for recycling of streams to various units).

The aromatic-rich hydrocarbon extract phase is heated in furnace **47** to about the hydrodesulfurization reaction temperature, and is then passed by way of line **48** into hydrodesulfurization unit **49**. In the hydrodesulfurization unit, the hydrocarbon extract is contacted with a hydrodesulfurization catalyst at a temperature between about 550° to 850° F. with hydrogen in an amount between about 1000 and 5000 s.c.f. per barrel of oil at a pressure between about 500 and 2000 p.s.i.g., e.g. about 750 to 1500 p.s.i.g., and at a space velocity of about 0.5 to 20 volumes of oil per volume of catalyst per hour. Any of the various hydrodesulfurization catalysts such as the mixed oxides of cobalt and molybdenum supported on an alumina carrier, molybdena on alumina, nickle tungsten sulfide, and in general the oxides and/or sulfides of groups 6 and/or 8 metals of the periodic table supported upon an alumina-type carrier may be employed. In the embodiment illustrated herein, the aromatics-rich extract is contacted with approximately 3000 s.c.f. of hydrogen per barrel of oil at a pressure of about 1000 p.s.i.g. and a temperature of about 750° F. while employing a space velocity of about 5 volumes of oil per hour per volume of catalyst. A cobalt oxide-molybdenum oxide-alumina containing about 3% cobalt oxide and 9% molybdenum oxide is used. Approximately 90% hydrodesulfurization of the cycle oil is obtained. Hydrogen consumption is about 1000 s.c.f./barrel of charge to hydrodesulfurization unit **49**. Based upon the 3,000 barrels per day of cycle oil extract obtained in the initial pass of crude oil through the process, hydrogen consumption amounts to about three million s.c.f. of hydrogen per day. Additional hydrogen requirements would occur if the cycle oil were not solvent extracted. In addition to reducing the sulfur content of the cycle oil, some hydrogenation and a slight amount of cracking occurs so that the hydrodesulfurized oil has a somewhat lower boiling point.

The hydrodesulfurized cycle oil extract is passed by way of line **51** into means for separating it into various boiling fractions. This means is depicted here as fractionating tower **52**. A hydrogen stream containing some of the H_2S evolved during hydrodesulfurization is taken overhead and passed by way of line **53** into valved line **54** by which it is returned to line **46** and employed in hydrodesulfurization vessel **49**. Naphtha formed during hydrodesulfurization (which may amount to about 10% of the charge to hydrodesulfurization vessel **49**, i.e. about 300 barrels per day based upon the initial charge of crude oil to the process) is removed from fractionating tower **52** and sent by way of line **56** to hydroforming. A higher boiling fraction, substantially all of which boils above about 600° F. and usually boils within the range of about 600° to 750° or 800° F., is removed from the bottom of fractionator **52** by way of line **57**. This higher boiling fraction amounts to about 800 barrels per day based upon the initial pass of crude oil through the process. A portion or all of it may be withdrawn by way of valved line **58** and employed for blending in fuel oil. Due to its reduced sulfur content and greater stability it forms a valuable component for that purpose. Because of its reduced sulfur content and because it has been partially hydrogenated, it may be passed from line **57** into valved line **59** and then sent to valved line **41** by which it may be recycled as a stock suitable for catalytic cracking.

A fraction of the hydrodesulfurized extract, substantially all of which boils within the range of about 400° to 600° F. is removed from fractionator **52** by way of line **61**. This lower boiling fraction of the hydrodesulfurized extract oil amounts to about 1,900 barrels per day based upon the initial crude oil passed on a once-through basis through the system. This oil is passed into furnace **62** wherein it is heated to the temperature needed for its hydrogenative cracking. The heated oil which may be at a temperature of about 600 to 900° F. is passed from the furnace by way of line **63** into hydrogenative cracking unit **64**. In vessel **64** this lower boiling hydrodesulfurized extract oil is contacted with hydrogen and a hydrogenative cracking catalyst at a temperature which is in the range of about 850 to 1000° F.

The catalyst is a dual-functioning catalyst which combines hydrogenation properties and cracking properties so as to cause hydrogenation of the extract oil and thereafter cracking of the oil. The hydrogenation components of such a catalyst may be the oxides and/or sulfides of the metals of group 6 and/or 8 of the periodic table (or the metals themselves). These are supported on a carrier having cracking properties such as natural and activated clays, synthetic catalytic cracking catalysts such as silica alumina, silica magnesia, silica alumina zirconia, or cracking bases such as HF promoted alumina. The catalyst may contain between 1 to 10%, preferably about 5% or thereabouts by weight, of the hydrogenation component supported in extended form upon the cracking component. The catalyst may be prepared by any of the conventional techniques such as by impregnation of the support with an aqueous solution of the hydrogenation component, by precipitation of the hydrogenation component upon the cracking support, or by co-precipitation of the hydrogenation component with the cracking component. For example a silica alumina cracking catalyst containing between 5 and 20% alumina with the remainder being silica, may be impregnated with a solution of ammonium molybdate, the impregnated catalyst dried and then calcined to convert the ammonium molybdate to molybdenum oxide; thereby producing a catalyst containing about 5% MoO_3. Other catalysts such as nickle on silica alumina, iron on silica alumina, platinum on silica alumina, platinum on fluorided alumina, cobalt molybdate on fluorided alumina, molybdenum oxide on fluorided terrana earth, and similar dual-functioning catalyst may be employed. This dual-functioning catalyst converts the polycyclic aromatics in the lower boiling extract oil to naphtha by hydrogenating one ring of the polycyclic, and thereafter by reason of the cracking component of the catalyst this hydrogenated ring is cracked whereupon the naphtha boiling range monocyclic aromatic is produced. In hydrogenative cracking vessel **64**, the lower boiling extract oil is contacted with the dual-functioning catalyst at the defined temperature (about 950° F. in this embodiment) and at a pressure of about 1000 to 5000 p.s.i.g., e.g. about 3000 p.s.i.g. while employing hydrogen in the amount of about 2000 to 6000 s.c.f./barrel of feed. A space velocity of from 1 to 20, e.g. about 5 volumes of oil per hour per volume of catalyst may be used. Conversions to lower boiling products on the order of 80% or higher are obtained, most of it being high octane naphtha having an antiknock value such as 100 F–1 or higher. Omission of the extraction step or the hydrode-

sulfurization step causes a drastic reduction in the octane number of the naphtha. Omission of the hydrodesulfurization step also causes a reduction in the extent of conversion as well as causing an increase in the rate of deactivation of the catalyst. Thus these preceding steps are essential. Fractionation of the hydrodesulfurized extract so that only the defined lower boiling fraction is charged to hydrogenative cracking is also essential to the production of high antiknock naphtha. Because the naphtha produced by hydrogenative cracking of the higher boiling fraction of the hydrodesulfurized extract has been found to have an octane number of 90 F–1 or lower, this higher boiling fraction is advantageously processed through the catalytic cracking unit wherein it yields a naphtha having an F–1 octane number of 95 or higher. In addition, this higher boiling extract fraction is much more resistant to conversion in the hydrogenative cracking unit. Only abont one-third of it is converted to lower boiling products as compared with 80% conversion of the lower boiling extracts. Its presence in the hydrogenative cracking unit **64** would thus tend to build up. Because increased coke formation is encountered through the use of this higher boiling extract fraction, the dual-functioning catalyst would have to be regenerated more frequently.

Approximately 2000 to 2500 s.c.f. of hydrogen per barrel of hydrodesulfurized extract charged to the hydrogenative cracking unit are consumed during the reaction. If the higher boiling fraction of the extract were processed through the hydrogenative cracking unit, an insufficient supply of hydrogen would exist. It would be necessary to employ generated hydrogen rather than to use the hydrogen which is produced during the hydroforming of the naphtha fraction. In the process described herein, sufficient hydrogen is generated to satisfy the total requirements of the integrated refining scheme. In this embodiment approximately four million s.c.f. of hydrogen/day (based upon the initial crude oil on a once-through basis) are consumed in hydrogenative cracking unit **64**. To maintain high catalyst activity, the hydrogen stream which is introduced into vessel **64** is relatively free of H_2S. This hydrogen stream is one which is separated from the products from the hydroforming of a desulfurized naphtha. The hydrogen stream is introduced by way of line **66** into line **61** by which it eventually reaches the hydrogenative cracking vessel **64**.

The reaction products from the hydrogenative cracking vessel **64** are removed therefrom and passed by way of line **67** to a fractionation system represented herein by fractionator **68**. Unconverted extract oil is separated as a bottom stream and passed by way of line **69** back to the hydrogenative cracking vessel **64**. The high octane naptha is removed as a side stream and passed by way of line **71** into line **27** where it is later blended with the other high octane naptha fractions produced to form the high octane gasoline. A hydrogen stream is removed overhead by way of line **72**. Because this stream will normally have a higher H_2S content than the hydrogen stream from hydroforming, it is passed to the hydrodesulfurization vessels wherein it serves as the hydrogen employed therein. If necessary, a portion of this stream may be recycled to the hydrogenative cracking vessel by way of line **73**, but it is preferred not to do so. The major portion of the hydrogen stream flowing in line **72** is diverted and passed by way of line **74** into line **54** by which it is charged to hydrodesulfurization vessel **49**. The remaining portion of the hydrogen stream is passed by way of line **75** and is employed in the hydrodesulfurization of the virgin naphtha.

The virgin naphtha removed from the crude oil in fractionating system **13** is passed by way of line **16** into furnace **76** wherein it is heated to the usual hydrodesulfurization temperature. The hydrogen stream in line **75** is also heated in the furnace tubes and is passed with the naphtha by way of line **77** into hydrodesulfurization unit **78**. In vessel **78** the virgin naphtha is desulfurized under the conditions and with the catalyst employed in hydrodesulfurization unit **49**, except that a somewhat lower temperature on the order of about 550° to 750° F. is used. Hydrogen consumption amounts to about 20 to 60 s.c.f., usually about 40 s.c.f. of hydrogen/barrel of naphtha charged. On the basis of once-through processing of the initial crude oil, as hereinbefore defined, hydrogen consumption amounts to about 300,000 s.c.f./day.

The products from hydrodesulfurization vessel **78** are passed by way of line **79** into fractionation system represented herein by fractionator **81**. A recycle hydrogen stream is removed overhead and is returned by way of line **82** to line **75**. The desulfurized naphtha is removed as a bottom stream from fractionator **81** and passed by way of line **83** through heater **84**. The small amount of naphtha produced during the hydrodesulfurization of the cycle oil extract is passed by way of line **56** into line **83**. The heated naphtha is then passed by way of line **86** into hydro-forming unit **87**.

In hydroforming unit **87**, the octane number of the naphtha is greatly improved, e.g., it is increased from about 60 F–1 up to 95 F–1 or higher. In the hydroforming reaction, naphthenes are dehydrogenated to higher octane aromatics and paraffins are cyclized to aromatics also. A substantial quantity of hydrogen is produced per barrel of naphtha charged. This may vary from about 500 to 1200 s.c.f. of hydrogen per barrel of naphtha charged. The catalysts employed in hydroforming are those such as molybdena on alumina, chromia on alumina, and platinum plus halogen on alumina or silica alumina support, etc. Of these, it is preferred to employ a platinum-type catalyst because it produces the greatest improvement in octane number of the naphtha and also results in a higher net production of hydrogen. It is particularly preferred to employ the process known as Ultraforming since it produces highest octane numbers and maximum hydrogen production, due in part to its operation at somewhat lower pressures on the order of 100 to 400 p.s.i.g. The hydroforming reaction is carried out by contacting the naphtha with the catalyst at a temperature of about 850° to 1000° F. and a pressure of about 50 to 750 p.s.i.g. A space velocity from 0.5 to 5 volumes of naphtha/hour/volume of catalyst is used. Hydrogen is introduced to the reactor at the rate of about 1000 to 6000 s.c.f./barrel of naphtha. In the embodiment described herein the naphtha is contacted with a platinum supported an alumina catalyst containing about 1% or even less of platinum at a temperature of about 925° F., a pressure of about 250 p.s.i.g., a space velocity of 1.5, with the introduction of about 3–4000 s.c.f. of hydrogen/barrel of naphtha charge. The octane number of the naphtha is improved from about 60 to about 98 F–1, and a net production of hydrogen in the neighborhood of about 1000 s.c.f./barrel of naphtha charge is obtained. Based upon the amount charged in this embodiment, 7.8 million cubic feet of hydrogen/day are produced.

The reaction products from the hydroforming step are passed by way of line **88** into a fractionating system, indicated herein by fractionating tower **89**, wherein various fractions are separated. High octane naphtha is removed from the bottom of fractionator **89** and passed by way of line **91** wherein it meets with the other high octane naptha fractions produced in the process. These fractions are blended with additional components to form the product high octane gasoline. A hydrogen stream is removed overhead from fractionator **89** by way of line **92**. A portion of this stream is recycled to the hydroforming process by way of line **92**. The remainder of the stream is diverted and passed by way of line **93** to the hydrogenative cracking vessel **64**. In startup operations it may be desirable to employ some of this hydrogen in the hydrodesulfurization units, but thereafter the integrated process functions best by charging the net production of hydrogen from the hydroformer directly to the furnace of the hydrogenative cracking vessel **64**.

It is apparent from the foregoing description, that the process of this invention provides an integrated system for producing maximum octane number naphtha in high yields in an efficient manner which eliminates the need for using outside hydrogen, and employs hydrogen produced in the integrated process in a manner which further benefits operation of the process.

Thus having described the invention what is claimed is:

1. A process for the manfacture of high octane naphtha fractions which comprises fractionating crude oil to produce naphtha and gas oil fractions therefrom, catalytically hydroforming said naphtha fraction to improve its octane number and simultaneously produce hydrogen, catalytically cracking said gas oil fraction to produce high octane naphtha and catalytic gas oil, solvent extracting said catalytic gas oil to separate an aromatics-rich hydrocarbon extract phase from an aromatics-lean hydrocarbon raffinate phase, hydrodesulfurizing said hydrocarbon extract phase, splitting the hydrodesulfurized extract phase into a lower boiling fraction substantially all of which boils below about 600° F. and a higher boiling fraction, catalytically cracking said higher boiling fraction to produce high octane naphtha, hydrogenatively cracking said lower boiling fraction in the presence of a hydrogen stream recovered from the products of said catalytic hydroforming and in the presence of a catalyst having hydrogenation and cracking properties to produce a high octane naphtha.

2. The process of claim 1 wherein a hydrogen stream is separated from the products of the hydrodesulfurization step and recycled thereto, and the makeup hydrogen which is supplied to said hydrodesulfurization step is a hydrogen stream separated from the products of the hydrogenative cracking step.

3. The process of claim 1 wherein the naphtha produced during said hydrodesulfurization step is passed to said hydroforming step.

4. A process for producing high octane gasoline boiling range hydrocarbons which comprises catalytically cracking a gas oil to produce high octane naphtha and catalytic gas oil, extracting said catalytic gas oil with a solvent which preferentially extracts aromatic hydrocarbons and thereby producing a hydrocarbon extract phase rich in aromatic hydrocarbons and a hydrocarbon raffinate phase lean in aromatic hydrocarbons, hydrodesulfurizing said hydrocarbon extract phase, splitting said hydrodesulfurized hydrocarbon extract phase into a lower boiling fraction and a higher boiling fraction, hydrogenatively cracking said lower boiling fraction by contacting it with a catalyst having hydrogenation and cracking properties in the presence of hydrogen and at a temperature of about 850° to 1000° F. and thereby producing a high octane naphtha.

5. The process of claim 4 wherein substantially all of said lower boiling hydrodesulfurized extract fraction boils between about 400° and 600° F., and wherein said higher boiling hydrodesulfurized extract is catalytically cracked.

6. A process for the manufacture of high octane naphtha fractions which comprises fractionating crude oil to produce naphtha and gas oil fractions therefrom, hydrodesulfurizing said naphtha fraction and thereafter hydroforming said hydrodesulfurized naphtha fraction to improve its octane number and simultaneously produce hydrogen, catalytically cracking said gas oil fraction to produce high octane naphtha and catalytic gas oil, solvent extracting said catalytic gas oil to separate an aromatics-rich hydrocarbon extract phase from an aromatics-lean hydrocarbon raffinate phase, charging said hydrocarbon raffinate phase to the catalytic cracking step, hydrodesulfurizing said hydrocarbon extract phase, splitting the hydrodesulfurized extract phase into a lower boiling fraction substantially all of which boils within the range

of about 400° to 600° F. and a higher boiling fraction substantiall all of which boils above about 600° F., charging said higher boiling fraction to said catalytic cracking step, hydrogenatively cracking said lower boiling fraction at a temperature between about 850° to 1000° F., a pressure between about 1000 and 5000 p.s.i.g. in the presence of hydrogen and a catalyst having hydrogenation and cracking properties and thereby producing high octane naphtha, recovering a hydrogen stream from the products of the hydroforming step and charging said hydrogen stream to the hydrogenative cracking step, recovering a second hydrogen stream from the products of the hydrogenative cracking step and charging portions of said second hydrogen stream to the hydrodesulfurization steps.

7. The process of claim 6 wherein said hydroforming step employs a supported platinum catalyst, an operating temperature of about 850° to 1000° F. and a pressure of about 50 to 400 p.s.i.g.

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