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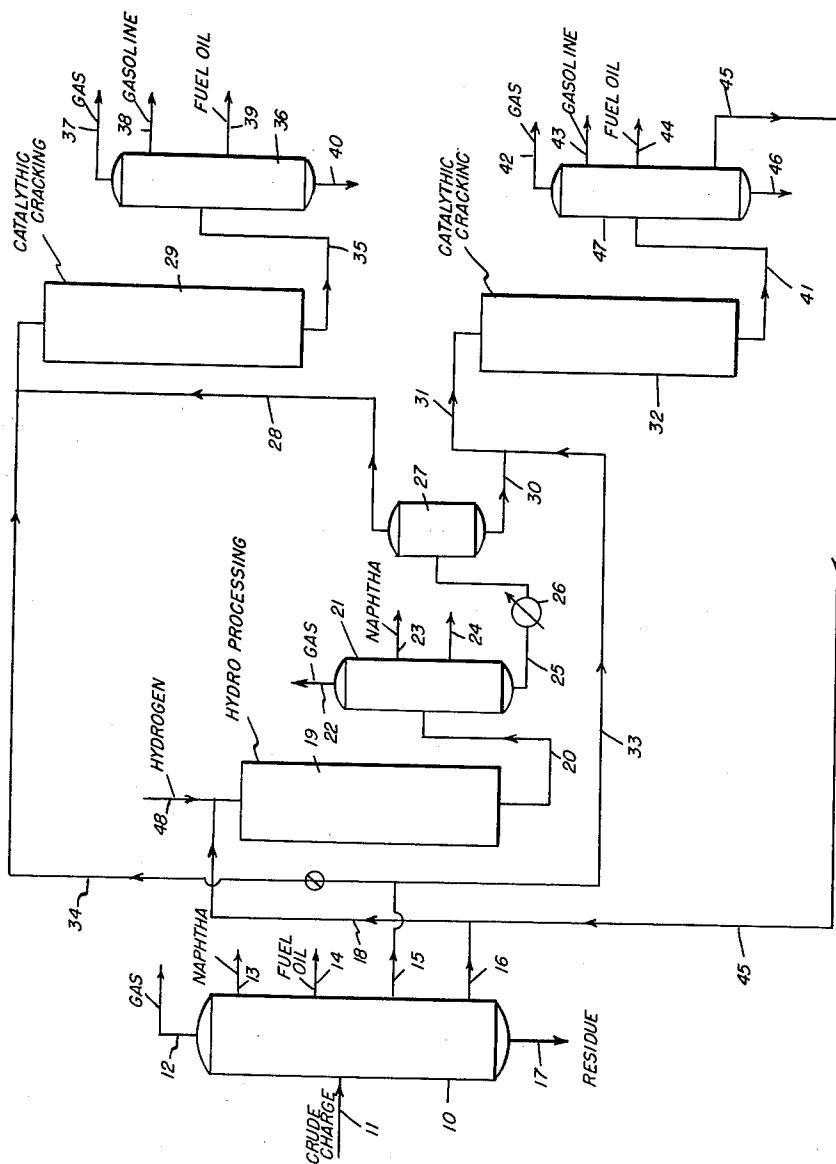
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COMBINATION CATALYTIC CRACKING-HYDROPROCESSING OPERATION

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2 Sheets-Sheet 1

FIG. 1



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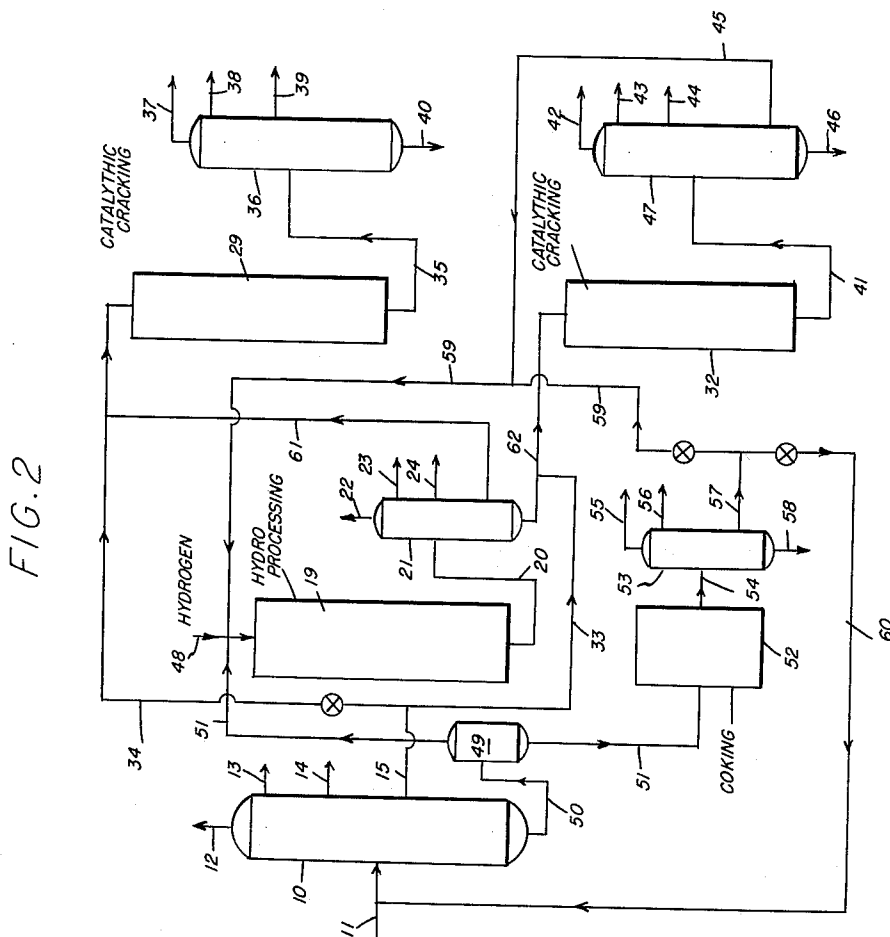
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2 Sheets-Sheet 2



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COMBINATION CATALYTIC CRACKING-HYDROPROCESSING OPERATION

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This invention relates to the conversion of high boiling hydrocarbons such as are derived from crude petroleum, shale oil, and like materials, into lower boiling hydrocarbons, particularly gasoline and fuel oil. More precisely, it relates to the conversion of high boiling hydrocarbons to gasoline-containing products by a combination hydrogenation-catalytic cracking operation.

In present day oil refineries the principal process by which high boiling hydrocarbons are converted into gasoline-containing products is catalytic cracking. The two principal catalytic cracking processes are Thermoform catalytic cracking, in which a solid particle-form catalyst is moved through the reaction zone as a compact bed, and fluid catalytic cracking, in which a powdered solid catalyst exists as a fluidized bed in the reaction zone.

Most present day refineries also employ catalytic reforming. In this process the octane number of low octane naphthas is increased so that the product is suitable for blending into gasoline of the high octane number required by today's motor vehicles. In addition to increasing the octane level of the liquid hydrocarbons, catalytic reforming produces substantial quantities of hydrogen.

The prior art indicates a number of ways in which the hydrogen produced in reforming may be employed elsewhere in the refinery to increase either the yield or quality of the desired refinery products. The prior art suggests that some of this hydrogen may be employed to treat domestic heating oils. Also, it has been pointed out that increased yields of gasoline and fuel oil from catalytic cracking may be obtained by hydrogenating the feed stock to the catalytic cracking operation. In addition, the prior art suggests that material heavier than gasoline and fuel oil produced in catalytic cracking may advantageously be hydrogenated and then returned to the catalytic cracking operation.

One factor that must be considered, however, is that the reforming operation does not, in the conventional refinery, produce sufficient hydrogen to accomplish all the hydrogenation reactions that might be desired. Careful studies must therefore be made to determine which of the available stocks are most advantageously reacted with the available hydrogen. One such study is given in Table I below.

TABLE I

Charge Stock	Straight Run Gas Oils				Catalytic Gas Oil	
	Light		Heavy			
Condition.....	(1)	(2)	(1)	(2)	(1)	(2)
Conversion to Gasoline and Lighter, vol. percent.....	40.4	47.6	36.2	50.4	20.1	39.1
Yield based on charge:						
Dry Gas, Wt. percent.....	3.6	3.8	4.4	5.9	3.3	5.2
Excess Butane, vol. percent.....	7.0	8.3	4.8	3.4	3.3	7.5
10 p.s.i. Ried Vapor Pres- sure Gasoline.....	33.5	42.0	28.8	44.8	13.0	30.4
Catalytic Gas Oil.....	59.6	52.4	68.8	49.6	79.9	60.9
Coke, wt. percent.....	1.9	1.0	5.1	2.1	2.4	1.4

¹ Raw.

² Hydrotreated.

Table I gives various yields from Thermoform catalytic cracking of three different charge stocks. Each charge stock was catalytically cracked raw and also was catalytically cracked after being hydrotreated. The catalytic cracking with each charge stock was conducted under the

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same conditions with the raw and hydrotreated stocks. It is apparent from the data in Table I that while hydrotreating improves the catalytic cracking yields of all three charge stocks, the improvement obtained on the heavy straight run gas oil and the catalytic gas oil is much more substantial than that obtained from hydrotreating the light straight run gas oil. This shows up particularly in the more greatly increased conversion, higher gasoline production and reduced coke make with these two stocks. It is, therefore, apparent that if hydrogen is available to improve catalytic cracking operations it is best used on the heavy straight run gas oil and the catalytic gas oil rather than light straight run gas oil.

There is one further problem, however, with the catalytic gas oil which, of course, is a high boiling portion of the catalytically cracked product. It is well known that material which has once been catalytically cracked is much more difficult to crack catalytically a second time. This is due to the build-up of certain materials called refractory components in the catalytic cracker product. These components are generally aromatics from which the alkyl side chains have been removed in the cracking operation. When recycling of material to the catalytic cracker is practiced, there is a tendency for these refractory components to build up in the recycle stream, making it more and more difficult to crack. The use of a hydrogenation step on this recycle stream improves this situation somewhat by hydrogenating these refractory materials. Nevertheless, there remains the possibility of a continuing build-up of them in the recycle stream.

A combination catalytic cracking-hydrogenation operation in which the available hydrogen is utilized on the charge stocks to the catalytic cracking unit which can most advantageously employ the hydrogen is the subject of this invention. In addition, this invention involves re-cracking of the catalytically cracked gas oil without the danger of build-up of refractory components.

A major object of this invention is the conversion of high boiling hydrocarbons to lower boiling hydrocarbons in an efficient, economical manner.

Another object of this invention is to provide for hydrogenation of charge stocks to catalytic cracking operations in the most efficient manner where there is not sufficient hydrogen available to hydrogenate all charge stocks to the catalytic cracking operation.

Another object of this invention is to provide for hydrogenation and re-cracking of catalytic gas oil in a catalytic cracking operation without build-up of refractory components in the gas oil over the course of time.

These and other objects of the invention will be apparent from the following description of the invention.

Before proceeding with this description certain terms employed herein will be defined. The term "hydroprocessing" is used herein to refer to a reaction between hydrogen and hydrocarbon charge in which there is a net consumption of hydrogen by the charge. It includes operations which involve only hydrogenation, such as the removal of impurities from the hydrocarbon charge and the hydrogenation of unsaturated components therein without material alteration in the boiling range of the charge, as well as operations wherein hydrogenation is accompanied by the production, through cracking, of substantial quantities of hydrocarbon material boiling below the charge.

The term "gas oil" is used herein to refer to the distillate material in a hydrocarbon fraction which boils above the gasoline boiling range. A "full range gas oil" is one which has a boiling range extending from the upper end of the gasoline boiling range to the temperature at which further distillation cannot be accomplished without thermal cracking, even by the use of vacuum. A "light gas oil" is the lower boiling portion of the full range gas

oil and typically might boil continuously between 450 and 650° F. A "heavy gas oil" is the higher boiling portion of the full range gas oil and typically might boil continuously from 650 to 1050° F.

The foregoing objectives are accomplished in this invention by hydroprocessing a high boiling hydrocarbon charge stock in a hydroprocessing zone to produce a hydroprocessed product. The hydroprocessed product is separated into a lower boiling portion and a higher boiling portion. The higher boiling portion is catalytically cracked in a first catalytic cracking zone to produce a gasoline-containing product. Catalytic gas oil which has a final boiling point below the initial boiling point of the high boiling hydrocarbon supplied to the hydroprocessing zone is separated from the catalytically cracked product of the first catalytic cracking zone and is recycled to the hydroprocessing zone to produce additional hydroprocessed product. The lower boiling portion of the hydroprocessed product is passed to a second catalytic cracking zone to be catalytically cracked therein. All of the products of the second catalytic cracking are removed from the system.

The phrase "removed from the system" is used herein to indicate that the products referred to are not recycled to either the hydroprocessing zone or either of the catalytic cracking zones which form a part of this invention. It should not, of course, be understood to mean that the products referred to may not undergo other further processing. The terms "initial boiling point" and "final boiling point" refer to the first and last temperatures observed in distillation of a material at standard pressure. With some heavy materials which crack before they are completely distilled it will be necessary to distill under vacuum and correct to standard conditions.

This invention will be best understood by referring to the attached drawings, of which:

FIGURE 1 is a diagrammatic flow plan of one operation within the broad scope of this invention, and

FIGURE 2 is a diagrammatic flow plan of a second operation within the broad scope of this invention.

Like parts in both drawings bear like numerals.

Considering, initially, FIGURE 1, a crude charge stock, such as a whole petroleum crude, is charged to fractionator 10 through line 11. In fractionator 10 the charge is separated into a plurality of fractions. The light gases may be removed overhead through line 12 and a naphtha cut taken through line 13. In a typical installation this naphtha might distill between 50 and 380° F. Frequently the naphtha will be further divided into a light naphtha, boiling, for example, from 50° F. to 180° F., which may be blended directly into the gasoline pool and a heavy naphtha, boiling, for example, from 150° F. to 380° F., which will be reformed in order to increase its octane number before blending into the gasoline pool.

A distillate fuel oil cut may also be taken through line 14. In a typical case this fuel oil might distill from 400° F. to 550° F. If desired, the domestic fuel oil cut need not be taken from fractionator 10 but the fuel oil may be included as a part of the light gas oil cut taken through line 15. This light gas oil might, for example, distill from 450° F. to 650° F. and it is handled in a manner described below.

Another fraction removed from fractionator 10 through line 16 is a heavy gas oil which, in a typical case, would distill from 650° F. to 900° F. As a final fraction a bottoms cut is taken through line 17.

The heavy gas oil flows from line 16 into line 18 and then into the hydroprocessing zone 19. There it is joined by hydrogen admitted through line 48. Typically, this hydrogen will be supplied by a reforming unit which, for the sake of simplicity, is not shown on FIGURE 1.

Hydroprocessing unit 19 may be operated in any conventional manner either to hydrogenate unsaturates and impurities only, or to produce substantial quantities of naphtha together with the hydrogenation. The condi-

tions of conversion in the hydroprocessing zone may vary widely, depending on the degree of conversion, the charge stock and the catalyst employed. Temperatures of conversion within the range 500° F. to 1000° F. are usual, although higher temperatures have been suggested. In general, it is preferred to maintain the conversion temperature within the range 500° F. to 900° F. because at the higher temperatures there is excessive production of gas (C₁, C₂ and C₃ hydrocarbons). Hydrogen pressures from a few hundred pounds per square inch to a few hundred atmospheres have been suggested. In general, however, pressures within the range 500 to 3000 pounds per square inch have found the most favor. The space velocity in volumes of hydrocarbon reactant (as 60° F. liquid) per volume of catalyst per hour should normally be within the range 0.1 to 10 and the molar ratio of hydrogen to hydrocarbon in the reaction zone should usually be within the range 2 to 80.

The hydroprocessing zone may employ any catalysts suitable to hydrogenate or hydrocrack hydrocarbons. Among the preferred catalysts is that described and claimed in United States patent application Serial Number 760,646, filed September 12, 1958. This catalyst is a composite of 15 to 40 percent by weight silica, 3 to 20 percent by weight molybdenum trioxide, 1 to 8 percent by weight cobalt oxide and the remainder alumina. Another preferred catalyst is that described in United States Patent 2,945,806, a continuation-in-part of copending application Serial Number 418,166, filed March 23, 1954, now abandoned. This catalyst is made up of 0.05 to 20 percent by weight of certain specified metals such as platinum, deposited on a base, such as silica-alumina, having a specified minimum cracking activity.

The hydroprocessed product flows from unit 19 through passage 20 into distillation column 21. From column 21 a gas stream is removed by means of line 22. This gas stream comprises predominantly hydrogen and usually after purification, may be recycled to hydroprocessing unit 19. A naphtha cut is taken through passage 23 and, if desired, a fuel oil cut may be taken at line 24. In most cases, however, any fuel oil boiling range material will be removed as part of the bottoms product through line 25. Typically, this bottoms product will boil within the range 380° F. to 850° F. In any case it is at least a part of the material boiling above naphtha, that is, above about 380° F., which is passed on through line 25.

The material in line 25 is heated by heater 26 and then flashed in flash drum 27 to effect its separation into a high boiling component and a lower boiling component. The lower boiling component is removed from vessel 27 as vapor overhead and passes by means of line 28 into a first catalytic cracking zone 29. The high boiling product passes by means of lines 30 and 31 into a second catalytic cracking reactor 32.

The temperature at which the split between material which passes overhead and that which is removed as bottoms should be controlled in the manner defined below. In a typical installation the material taken overhead through line 28 might boil within the range 380° F. to 650° F., while the material taken through line 30 might boil within the range 650° F. to 900° F.

Also supplied to catalytic cracker 32, by means of lines 15, 33 and 31, is the straight run light gas oil from fractionator 10. If there is an excess of this gas oil the excess may be passed to catalytic cracker 29 through line 34. A less preferred alternative is to supply all of the straight run light gas oil to catalytic cracker 29.

Catalytic cracking units 29 and 32 may be operated in any conventional manner to convert the feed stock thereto to lower boiling products. They may employ fluidized beds or moving compact beds of solid catalysts. A variety of such catalysts are known in the art. They include natural and treated clays and synthetic associations of silica, alumina, magnesia and combinations thereof. Suitable

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facilities for continuous regeneration of the cracking catalyst will be a part of these units.

Reaction conditions in the catalytic cracker may be conventional, including a reaction temperature within the range 600 to 1000° F., a reaction pressure within the range 5 to 30 p.s.i.g., a catalyst to oil weight ratio within the range 0.5 to 20 and a space velocity within the range 1 to 10 volumes of charge (measured as 60° F. liquid) per hour per volume of catalyst.

The product of unit 29 is removed through passage 35 to fractionator 36. In the fractionator the product is separated into conventional gas, gasoline and fuel oil products which are taken through lines 37, 38 and 39. Any material heavier than fuel oil is removed from the system through line 40. This material is neither recycled to the catalytic cracker nor to the hydroprocessor. In view of the relatively low boiling range of the charge stock to catalytic cracker 29, the quantity of material produced which is heavier than fuel oil normally will be negligible.

The product of catalytic cracker 32 passes by means of line 41 into fractionator 47. Gas and gasoline are taken as products at 42 and 43. A domestic fuel oil cut may also be removed at 44 but in most cases it will be desirable to take fuel oil boiling range material as a part of the catalytic gas oil product removed at 45 which is passed through lines 45 and 18 into hydroprocessing zone 19 to be converted therein. This catalytic gas oil will boil within the range about 400° F. to 650° F. A bottoms product is taken at 46.

It is an important element of this invention that the final boiling point of the catalytic gas oil in line 45 is below the initial boiling point of the heavy gas oil which also passes to hydroprocessing unit 19. Generally, the final boiling point of the catalytic gas oil should not be more than 100° F., and preferably not more than 25° F., below the initial boiling point of the straight run gas oil. The foregoing limitations may, of course, be achieved either by a control on the operation of fractionator 10 to regulate the initial boiling point of heavy straight run gas oil or by control of the operation of fractionator 47 to regulate the final boiling point of the catalytic gas oil.

In addition, the temperature below which material is taken overhead from flash chamber 27 and above which it is taken as bottoms through line 30 should be controlled so as to be near the final boiling point of the catalytic gas oil and the initial boiling point of the heavy straight run gas oil. Of course, there may be some overlap in boiling between the overhead taken at line 27 and the bottoms taken at line 30. For this invention, however, the initial boiling point of the material taken at line 30 should be above the final boiling point of the catalytic gas oil. Preferably, the initial boiling point of the material taken through line 30 should not be more than 100° F. above the final boiling point of the catalytic gas oil to avoid production of large quantities of less desirable heavy hydrocarbons in cracking unit 29. In a less desirable operation within the broad scope of this invention, the initial boiling point of the material in line 30 may be below the final boiling point of the catalytic gas oil but it should be less than 50° F. therebelow.

It will be apparent that by controlling the various boiling ranges in the manner defined, there can be no build-up of heavy refractory catalytically cracked materials in the catalytic gas oil. There can be no substantial quantity of hydrocarbons which have once been through cracking unit 32 returned to cracking unit 32. Substantially all of the catalytic gas oil after hydroprocessing goes to catalytic cracker 29 and, since all of the products of cracker 29 are removed from the system, these materials will never be returned to cracker 32.

Another advantage of this system may be noted. Catalytic cracker 29 handles only lower boiling material which can readily be vaporized under temperature and pressure conditions usually employed in catalytic cracking. If

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there is any heavier material which must be fed as a liquid, it is supplied to catalytic cracker 32. Thus, only catalytic cracker 32 need be equipped to handle liquid feed. Also, the two crackers may be operated under different sets of conditions to accommodate the two different charge stocks and produce greater yields of gasoline and fuel oil than if the total hydroprocessor effluent were supplied to one or more crackers in parallel.

FIGURE 2 illustrates the use of coking in combination with this invention and also shows operations which are alternatives for several of the process steps of FIGURE 1. That part of the operation of FIGURE 2 which is the same as the operation of FIGURE 1 will not be discussed here, for the sake of brevity. The same reference numerals are used for the various portions of FIGURES 1 and 2 which operate the same.

In the operation of FIGURE 2 all of the material heavier than light gas oil from fractionator 10 is supplied to a vacuum separator 49 through line 50. A reduced pressure, for example 70 millimeters of mercury, suitable to vaporize the maximum amount of material without thermal cracking, is maintained in separator 49. The overhead is the heavy gas oil which is passed to hydroprocessor 19 through line 51. The heavy bottoms are passed to coking unit 52 through line 53.

The coking operation may be conducted in any desired manner. It may employ a moving bed or a fluidized bed of hot solids. It may be conducted in the more conventional delayed coking manner. Coking is a thermal cracking operation conventionally conducted at temperatures within the range 850° F. to 1400° F. and at pressures within the range 1 to 50 p.s.i.g.

One product of the coking unit will be solid petroleum coke. The other products, liquid and vapor, are passed to a fractionator 53 by means of line 54. Gas is removed at line 55 and naphtha at line 56. A coker gas oil which might distill within the range about 380° F. to 900° F. is removed at line 57 and a heavy bottoms material is taken at line 58.

The coker gas oil may be handled in one of two ways. It is desirable that all of the coker gas oil be subjected to hydroprocessing since its quality as a catalytic cracking charge stock can be greatly improved in this manner. Thus, if sufficient hydrogen is available, all of the coker gas oil should be passed to hydroprocessing unit 19 via lines 57 and 59. If the hydrogen available is not sufficient for this, the coker gas oil may be supplied to fractionator 10 via lines 57, 60 and 11. The coker gas oil will then be fractionated with the crude charge and the lighter portion will pass directly to one or both of the catalytic cracking units 29 and 32 while the heavier portion will be passed to the hydroprocessing unit 19 through line 50, unit 49 and line 51.

In the process of FIGURE 2, fractionator 21 functions to split the hydroprocessed product boiling above naphtha in the manner described above, into a lighter portion which goes to catalytic cracker 29 through lines 61 and 34 and a heavier portion which goes to catalytic cracker 32 through line 62. This method of separation may be more expensive than the flash drum type employed in FIGURE 1 but it will separate the two portions more precisely.

As in FIGURE 1, the process of FIGURE 2 separates from the product of catalytic cracker 32, a catalytic gas oil having a final boiling point below the initial boiling point of the heavy gas oil in line 51. This catalytic gas oil goes to hydroprocessor 19 through lines 45 and 59.

It will be immediately apparent that this invention is not limited to any particular type of apparatus. Also, where single reaction units have been indicated, such as 19, 29 and 32, two or more units may be operated in parallel.

The catalytic cracking units will have some provision for continuous regeneration of the catalyst used. Similarly, where hydroprocessing is conducted in the presence

of a catalyst, there will usually be some means for periodic regeneration needed.

Hydroprocessing may be conducted without a catalyst as in the well-known hydrogen donor diluent process. Preferably, however, hydroprocessing is conducted over a catalyst, such as those previously mentioned, in a fixed bed.

Example I

This example involves a design for processing a California crude pursuant to this invention. About 18,130 barrels per day of light straight run gas oil with an initial boiling point of 450° F. and a final boiling point of 650° F. and 10,110 barrels of heavy straight run gas oil with an initial boiling point of 650° F. and a final boiling point of 900° F. were processed. About 16,330 barrels per day of the light straight run gas oil were passed to two Thermoform catalytic cracking (TCC) units operated in parallel in the position of unit 32 in the drawings. The remaining 1800 barrels per day of light straight run gas oil were passed to a third Thermoform catalytic cracker in the position of unit 29 in the drawings. All of the cracking units were operated under the following conditions:

Space velocity, vol. oil/hr./vol. catalyst-----	0.625
Oil inlet temperature, ° F.-----	930
Catalyst inlet temperature, ° F.-----	1025
Catalyst to oil volumetric rate-----	2.0

A catalytic gas oil amounting to 10,761 barrels per day with an initial boiling point of 400° F. and a final boiling point of 600° F. would be combined with the heavy gas oil and hydroprocessed under the following conditions:

Space velocity, vol. of charge per hr. per vol. of catalyst-----	2.0
Reaction temperature with fresh catalyst ---° F.---	805
Reaction pressure, p.s.i.g.-----	2000
Hydrogen circulation, cubic feet per barrel of feed -	2500

The catalyst used in the hydroprocessing comprised the oxides of cobalt, molybdenum, silicon and aluminum. The hydroprocessor effluent boiling above naphtha was subjected to a flash separation to produce 9,271 barrels per day of heavy material boiling from 600° F. to 750° F. and 10,008 barrels per day of light material boiling from 400 to 600° F. The heavy material went to the two cracking units occupying position 32, while the light portion went to the unit occupying position 29. The total products of this combination were as follows:

Dry gas, thousands of pounds per day -----	1160
Cracking unit coke, pounds per day -----	290
Butanes, barrels per day -----	4200
C ₄ free TCC gasoline, barrels per day -----	15,070
Hydroprocessed naphtha, barrels per day -----	1940
No. 2 fuel oil, barrels per day -----	4910
Bottoms, barrels per day -----	1560

Example II

In the Example I the limiting factor on the quantity of material that could be processed was the ability of the catalyst regeneration system to burn off coke and regenerate the catalyst. The quantity of charge that was processed was the quantity which kept the Thermoform catalytic cracking regenerators operating at full capacity. For purposes of comparison, this example gives the yields of products from the same TCC units operated with an amount of charge sufficient to keep the regenerators full but without the hydroprocessing step. Without hydroprocessing the system was able to handle only 19,200 barrels per day of straight run gas oil rather than the

28,240 barrels per day of Example I. The operating conditions in all three TCC units were about as follows:

Space velocity, vol. charge/hr./vol. catalyst -----	1.0
Oil inlet temperature, ° F. -----	930
Catalyst inlet temperature, ° F. -----	1025
Catalyst to oil volume ratio -----	2.0
Recycle ratio, vol. recycle/vol. fresh feed -----	1.1

The products produced were as follows:

Dry gas, thousands of pounds per day -----	670
TCC coke -----	310
Butanes, barrels per day -----	1860
C ₄ free TCC gasoline, barrels/day -----	10,100
No. 2 fuel oil, barrels/day -----	4610
Bottoms, barrels/day -----	1210

It will be seen that without the hydroprocessor the TCC units produce almost 5000 barrels per day less gasoline and the system does not have the advantage of 1940 barrels per day of naphtha from the hydroprocessor which may be reformed to gasoline.

This invention should be understood to include all changes and modifications of the examples of the invention, herein chosen for purposes of disclosure, which do not constitute departures from the spirit and scope of the invention.

I claim:

1. A process for the efficient conversion of a hydrocarbon charge stock boiling in the range from about 650° F. to about 1050° F. to lower boiling products with a minimum build-up of refractory components which comprises:

- (1) reacting said hydrocarbon charge stock in the presence of hydrogen utilizing reaction conditions wherein said reaction results in a net consumption of hydrogen to produce a hydroprocessed product;
- (2) separating the portion of the hydroprocessed product boiling above about 380° F. from said hydroprocessed product into a lower boiling fraction boiling in the range from about 380° F. to about 650° F. and a higher boiling fraction boiling in the range from about 650° F. to about 900° F.;
- (3) catalytically cracking said lower boiling portion of said hydroprocessed product to a gasoline-containing product, all of which is removed from the system;
- (4) [a] catalytically cracking said higher boiling hydroprocessed fraction in a separate catalytic cracking zone to lower boiling gasoline-containing products;
- [b] separating a catalytic gas oil boiling in the range from about 400° F. to about 650° F. from said gasoline-containing product;
- [c] recycling said catalytic gas oil to the hydroprocessing step (1); and
- [d] processing said catalytic gas oil through the sequence of steps (1), (2), (3), (4a), (4b) and (4c).

2. In a continuous process for the efficient conversion of a hydrocarbon charge stock boiling in the range from about 650° F. to about 1050° F. to lower boiling products with a minimum build-up of refractory components which comprises:

- (1) reacting said hydrocarbon charge stock in the presence of hydrogen utilizing reaction conditions wherein said reaction results in a net consumption of hydrogen to produce a hydroprocessed product;
- (2) separating the portion of the hydroprocessed product boiling above about 380° F. from said hydroprocessed product into a lower boiling fraction boiling in the range from about 380° F. to about 650° F. and a higher boiling fraction boiling in the range from about 650° F. to about 900° F.;

- (3) catalytically cracking said lower boiling portion of said hydroprocessed product to a gasoline-containing product all of which is removed from the system;
- (4) [a] catalytically cracking said higher boiling hydroprocessed fraction in a separate catalytic cracking zone to lower boiling gasoline-containing products; 5
- [b] separating a catalytic gas oil boiling in the range from about 400° F. to about 650° F. from said gasoline-containing product; 10
- [c] recycling said catalytic gas oil to the hydroprocessing step (1); and
- [d] processing said catalytic gas oil in the presence of the hydrocarbon charge stock through 15

the sequence of steps (1), (2), (3), (4a), (4b) and (4c).

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