

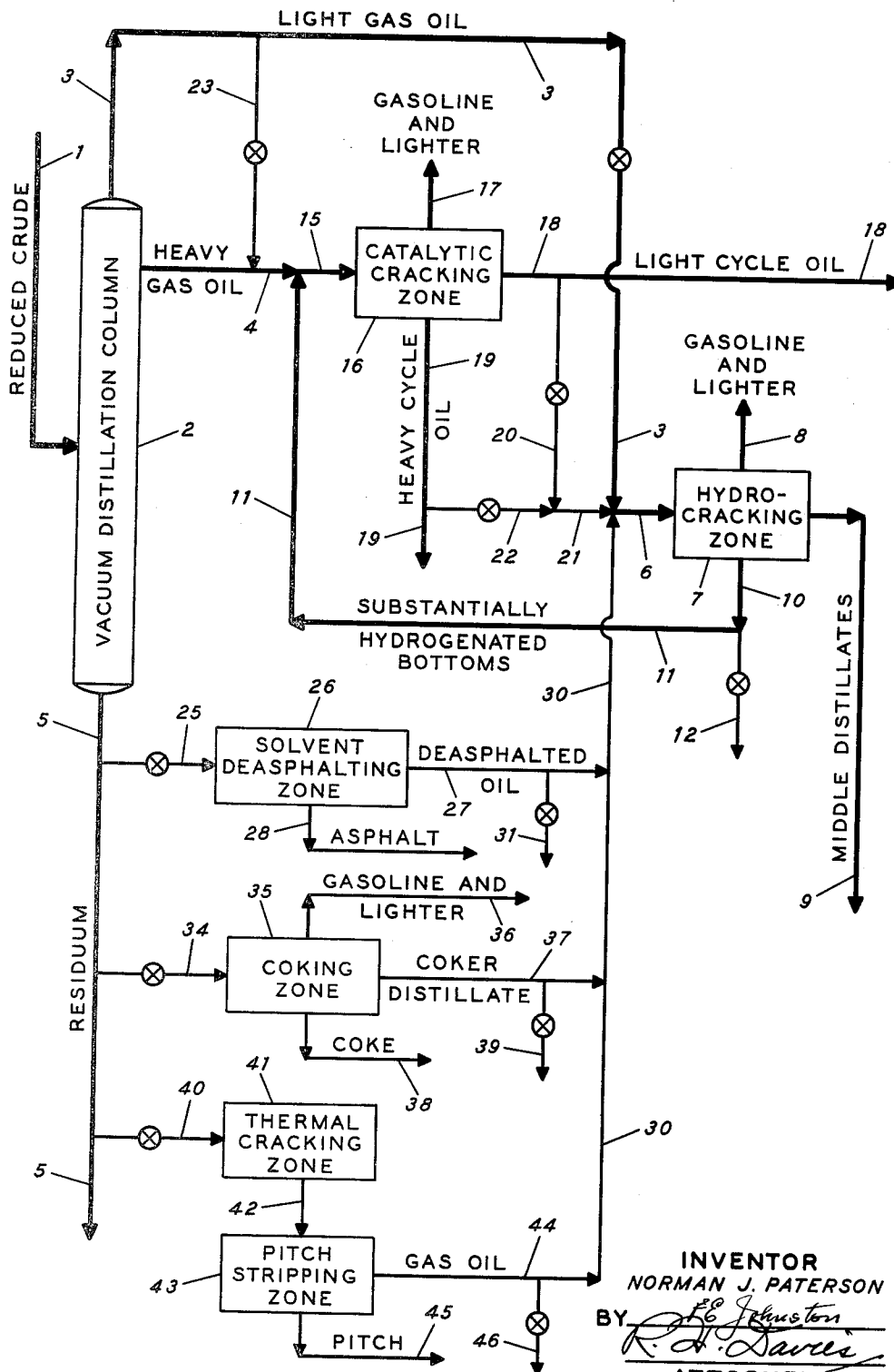
April 12, 1966

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3,245,900

HYDROCARBON CONVERSION PROCESS

Filed Dec. 26, 1963



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3,245,900

HYDROCARBON CONVERSION PROCESS

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Filed Dec. 26, 1963, Ser. No. 333,666

6 Claims. (Cl. 208—56)

This invention relates to processes for the catalytic conversion of hydrocarbons boiling in the gas oil boiling range along with residual fractions, to produce more valuable products, including gasoline and middle distillates. More particularly, the invention relates to combinations of catalytic cracking and hydrocracking.

DEFINITIONS

This specification contains numerous terms to characterize various hydrocarbon feeds and fractions thereof, and various conversion products. Therefore, those terms which do not have well established meanings in the art will be defined in order to facilitate understanding the subsequent description of the present invention.

The term "gas oil" is a broad general term that includes a variety of stocks. It includes any fraction distilled from petroleum, which has an initial boiling point of at least about 300° F. The gas oil can be further divided into overlapping boiling ranges. For example, a light gas oil boils between about 300° and 650° F., a medium gas oil boils between about 600° and 750° F., and a heavy gas oil boils between about 600° and 1100° F.

The term "catalytic cycle oil" also is a broad general term which covers a variety of stocks. Thus, a light catalytic cycle oil is predominantly a synthetic product boiling above the end point of gasoline and below the fresh feed initial boiling point, i.e., between about 430° and 650° F., which is highly aromatic. A heavy catalytic cycle oil boiling between about 650° and 850° F. consists of converted and unconverted portions of the fresh feed stocks and is commonly recycled with the fresh feed to increase the yield of lighter and more valuable products in a catalytic cracking unit.

The following are the terms for the products produced in the catalytic cracking and hydrocracking zones and operating definitions.

The gasoline produced in the catalytic cracking zone is usually a fraction which boils from about C₅ to 430° F. The gasoline produced in the hydrocracking zone is usually taken off a fractionating column in two fractions; one fraction is termed "light gasoline," which boils from about C₅ to 180° F., and the other fraction is termed "heavy gasoline," which boils from about 180° to 400° F. and is conventionally used as reformer charge stock.

The term "hydrocracking zone bottoms fraction," as used herein, is one which is essentially free from any synthetic hydrocracking zone product. In the case of gasoline production, it boils above the end point of gasoline, for example 400° F.

The term "conversion barrels" which is used to evaluate the conversion capacity of a catalytic cracker is the product of the fresh feed rate in barrels per stream day multiplied by the liquid volume percent conversion (distillate plus loss (D+L) below 430° F.) and divided by 100.

The term "gasoline efficiency" defines the ratio of gasoline barrels divided by conversion barrels times 100 and is expressed as a percentage.

The term "coke burning rate," as pounds of coke burned, expresses an operating limitation on a catalytic cracking unit and defines the pounds of coke burned in the regenerator per operating hour to regenerate the catalyst for reuse in the reactor. Thus, it is understood, but not generally expressed, that the catalyst is regenerated

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to a constant carbon or coke level after each regeneration. The ability of the regenerator on a catalytic cracking unit to remove the deposited coke on the spent catalyst limits the throughput and conversion that may be taken in the catalytic reactor.

PRIOR ART PROBLEMS

Today, the refiner has an increasingly difficult task in meeting the market demand for sufficient quantities of high-octane gasoline. To meet this demand, the refiner is looking for new processes and ways to realize maximum utilization of old conversion processes. Hydrocracking is one relatively new process to which the refiner has been giving a great deal of attention. This process is able to convert a wide range of high boiling feeds into gasoline, middle distillates and other useful products. At the same time, the older processes, such as catalytic cracking, have been studied for ways to increase the conversion to gasoline and to increase the capacity of existing units. One way to increase the capacity of many existing catalytic crackers would be to decrease the production of coke, a carbonaceous deposit which is deposited on the cracking catalyst. This follows, since the coke burning capacity of the catalytic cracking regenerators is often the bottleneck that limits additional throughput and conversion barrels. Thus, two of the major problems that face the refiner are (1) to convert more feed to gasoline and (2) to produce less coke in a given catalytic cracker.

A point is reached in an existing unit where conversion barrels cannot be increased further because of limitations on the coke-burning capacity of the catalyst regenerators. It would be desirable, if a method were available, to increase the conversion barrels to gasoline in an existing catalytic cracker operating at coke-burning limits.

Although recycling heavy catalytic cycle oil with a fresh feed is a common prior art method of increasing the yield of gasoline, it creates new problems of its own. Such heavy cycle oils undergo some change during passage through the catalytic cracking zone, which makes the material less desirable for re-cracking than the original feed stock. By recycling a portion or all of the heavy cycle oil to the catalytic cracking zone, there is a build-up of high-molecular weight polynuclear aromatics containing two or more benzene rings and a condensed nucleus. Such polynuclear aromatics which build up in the heavy cycle oil decompose during the catalytic cracking or re-cracking operation to form large quantities of coke. It is well known to improve the quality of catalytic recycle stocks by solvent extraction, whereby the less refractory raffinate from the solvent extraction zone is returned to the catalytic cracking zone and the less desirable aromatic extract is disposed of. Although this treatment removes polynuclear aromatics from the recycle stock thus permitting higher conversions to gasoline, it also down-grades a portion of the feed stock components to fuel oil and less valuable uses. It would be desirable, if a method were available, to deactivate the recycle stocks, as far as their tendency to form coke is concerned, and to activate the recycle stocks, as far as their tendency to form high octane gasoline is concerned.

Many of the prior art methods for increasing conversion to gasoline apply to decreasing the coke production as well. One such method has been to partially hydrogenate the total catalytic cracking feed stock. In this method, the gasoline yield and the gasoline to coke production ratio is increased by partially hydrogenating the highly-condensed aromatics in the catalytic cracking feed stocks. However, this improvement in an existing catalytic cracker results in at least two new problems of its own: (1) it requires a hydrogenation unit large enough to handle at least the total catalytic cracking feed stock and (2) the hydrogenation unit requires either frequent

regeneration and/or addition of fresh hydrogenation catalyst which results in considerable down time. Another method for improving catalytic cracking operations by reducing coke and light gas production is to partially hydrogenate cycle oils from the cracking operation before they are recycled to the catalytic cracking reactor. By this method, olefins, nitrogen, and other harmful materials are hydrogenated and aromatics are partially saturated; they are thus prevented from contributing to the coke production on their recycle to the catalytic cracking zone. The improvement made by this latter method is to reduce that portion of the coke production attributed to the recycle and the small increase in gasoline in the catalytic cracker that results from the conversion of recycle stock. It would be desirable, if a method were available, to increase the conversion to gasoline from the whole feed and, at the same time, decrease the production of coke in an existing catalytic cracker operation which is limited in conversion by coke burning in the regenerator.

Inherent in one of the major problems facing the refiner, i.e., to convert more catalytic cracking feed to gasoline, is how to decrease the production of lower value stocks, such as fuel gas and bunker fuel oil.

The production of fuel gas has been increased by the recent shift in the demand for petroleum products. Until recently, light straight-run and cracked gas oils, boiling from about 450° and 700° F., have been used in domestic heating oils. The demand for such heating oils has decreased, because of the increasing use of natural gas in many marketing areas. Thus, they have been diverted to catalyst cracking feed stocks where they tend to produce a high ratio of butanes and lighter gases, when such feeds are processed at conventional conversion levels. One of the prior art methods for relieving this problem is to pass the lighter gas oils to a hydrocracking zone instead, where they can be converted almost completely into gasoline. However, unless a sufficient supply of relatively low-cost hydrogen is available, the process is, in many cases, uneconomical compared to disposing of the lighter gas oils at distress prices or to including them into catalytic cracking feed stocks, and accepting the lower realization from the increased fuel gas. It would be desirable, if a method were available, to partially convert the lighter gas oils into gasoline in hydrocracking and catalytically cracking zones without the increase in fuel gas production.

Likewise, the disposal of heavy gas oils into bunker fuel oils has also become uneconomical in recent years, due to the competition from natural gas. In addition, air pollution restrictions in many areas, which prevent the burning of bunker fuel due to its high sulfur content, the almost complete dieselization of railroads, and the competitive advantages afforded low-cost foreign bunker fuels, make disposal of heavy gas oil economically unattractive. One of the problems that the refiner faces, by including heavy gas oils into catalytic cracking feeds, is that they produce more coke at a given conversion level, compared to lighter gas oils. Another problem is the effect caused by including these heavier gas oils on product distribution due to the presence of metallic contaminants and catalyst poisons. Generally, the percentage of metallic contaminants and catalyst poisons in the heavier gas oil increases with boiling range, along with nitrogen content. Methods have been used to purposely undercut the end point of the heavy gas oil from the crude distillation columns to maintain the catalytic cracking feed stocks at low contaminant levels. However, this results in diverting valuable catalytic cracking stocks to lower value fuel oil. Other prior art methods for handling the heavier gas oils have been to process the crude vacuum residuum in a solvent deasphalting zone to produce a deasphalted oil having a low carbon and metals content. This method is applicable only to a narrow range of crude oils, and it is not possible to pro-

duce a deasphalted oil directly suitable for catalytic cracking from the majority of the available crude oils, unless a very low yield is taken in the deasphalting zone, or expensive treating processes are used to remove metallic contaminants. Still other prior art methods for producing additional quantities of catalytic cracking feed stocks have been to hydrocrack the reduced crude oil. These methods have not been successful to date, due to the high coke formation and poisoning of the hydrocracking catalyst with metallic contaminants in the feed stocks. It would be desirable, if a method were available, to handle heavy contaminated refractory oil, or portions thereof, in a hydrocracking zone to upgrade them so they can be processed in a catalytic cracking zone.

OBJECTS

The following objects of the present invention include the solutions to the above-mentioned problems:

It is a general object of this invention to provide a process wherein increased quantities of catalytic cracking feed stock can be effectively converted into gasoline.

Another object of the present invention is to provide a process for handling said increased quantities of catalytic cracking feed stocks, so that a larger portion may be converted into gasoline and middle distillates instead of all, or a substantial portion thereof, being diverted to less valuable uses, as has been necessary heretofore.

Another object of the present invention is to provide a process for increasing the capacity of an existing catalytic cracker, which is conversion limited due to being at the limits of the coke burning capacity, so that the yield of gasoline may be maximized.

Another object of the present invention is to provide a process combination, consisting of a catalytic cracker and a hydrocracker, whereby the gasoline-producing efficiency of the catalytic cracker is improved to an extent heretofore not possible.

Another object of the present invention is to provide a process for handling many types of heavy gas oil feeds in a hydrocracking zone.

STATEMENT OF INVENTION

In accordance with one embodiment of the present invention, there is provided a method of decreasing coke production and increasing gasoline production in a catalytic cracking process which comprises increasing the feed rate to said catalytic cracking process by adding to each volume of original feed between 0.1 and 1 volume of a substantially hydrogenated bottoms stream produced by hydrocracking a selected hydrocracker feed characterized by containing at least 20 volume percent aromatics and boiling entirely above 300° F., said hydrocracking being carried out at conditions whereby at least 30 volume percent of said selected feed is converted to recovered products boiling below the initial boiling point of said selected feed with adequate hydrogen being consumed, above 750 s.c.f./bbl., to substantially hydrogenate the material not so converted and thereby to produce said hydrogenated bottoms, and said catalytic cracking process being carried out at conditions which would accomplish conversion of between 30 and 80 percent of said original feed in the absence of said added bottoms stream, whereby total coke production with the increased feed rate is less than with the original feed, and gasoline production is increased by an amount which is greater than the amount which would be obtained by catalytically cracking the added bottoms stream alone at said conditions.

In accordance with a more preferred embodiment of the present invention, there is provided a method of decreasing coke production and increasing gasoline production in a catalytic cracking process which comprises increasing the feed rate to said catalytic cracking process by adding to each volume of original feed between 0.1 and 1 volume of a substantially hydrogenated bottoms stream produced by hydrocracking a selected hydro-

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cracker feed characterized by containing at least 20 volume percent aromatics and boiling above about 650° F., said hydrocracking being carried out at conditions whereby at least 30 volume percent of said selected feed is converted to recovered products including a product in the gasoline boiling range and a product in the middle distillate boiling range, boiling below the initial boiling point of said selected feed with adequate hydrogen being consumed, above 750 s.c.f./bbl., to substantially hydrogenate the material not so converted and thereby to recover said hydrogenated bottoms boiling entirely above the end point of the middle distillate product, and said catalytic cracking process being carried out at conditions which would accomplish conversion of between 30 and 80 percent of said original feed in the absence of said bottoms stream, whereby total coke production with the increased feed rate is less than with the original feed, and gasoline production is increased by an amount which is greater than the amount which would be obtained by catalytically cracking the added bottoms stream alone at said conditions.

BASIS FOR THE PRESENT INVENTION

The following is a discussion of the bases for setting the operating limits of the present invention. One of the essential aspects of the present invention is the addition of the highly-hydrogenated hydrocracker bottoms to the catalytic cracker. This bottoms stream is of a higher hydrogen content than the fresh gas oil feed to the catalytic cracker. It is found that, under similar operating conditions of catalyst temperature and space rate, the bottoms stream will convert more readily into lighter products than the fresh gas oil feed. Further, it has been found that under operating conditions to give equal coke productions, the highly-hydrogenated bottoms will show a large increase in conversion relative to the fresh gas oil feed. Still further, it has been found that, at conditions to give equal conversion, the highly-hydrogenated bottoms will result in only a fraction of the coke production that will occur with the fresh gas oil feed alone. In fact, at equal operating conditions in the catalytic cracker, the highly-hydrogenated bottoms will not only show a much lower coke production but also a large increase in conversion, compared with the fresh gas oil feed alone.

One of the operating limits of the present invention is that the hydrocracker feed contains at least 20 volume percent aromatics and preferably between about 20 and 45 volume percent aromatics. This is to ensure a reasonable saturation of aromatics in the feed. Under the hydrogenation conditions in the hydrocracker, the aromatics are converted partly to naphthenes and partly from polynuclear aromatics to mononuclear aromatics. Contrary to other reported processes, better results are obtained when the hydrocracker bottoms stream is substantially hydrogenated rather than partially hydrogenated. It has been found that the addition of highly paraffinic bottoms from highly paraffinic hydrocracker feeds, i.e. ones containing less than 20% aromatics, do not give the foregoing results of increased conversion and decreased coke production.

Another operating limit of the present invention is the concentration of the hydrocracker bottoms stream in the original catalytic cracker feed. The minimum concentration is set at 0.1 volume of said bottoms in each volume of said feed. If the combined stream is composed of less than 0.1 volume of bottoms, the dispersion of this highly-hydrogenated stream is not complete enough to cause the reduction in coke production or to increase the conversion to gasoline. The maximum concentration is set at 1.0 volume of said bottoms in each volume of said feed. If the combined stream is composed of more than 1 volume bottoms per volume of catalytic cracker feed, the conversion and fresh gas oil feed ca-

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capacity of the catalytic cracker will start to decrease due to overcracking in the catalytic cracker.

Preferably, the concentration of bottoms fraction is 0.1 to 0.2 volume per volume of catalytic cracker feed. By operating within these limits, the production of light gases in the catalytic cracker is not increased markedly and the advantages of the present invention are maintained, i.e. increased conversion and decreased coke production.

If light straight-run gas oil or catalytic cycle oil is available as hydrocracker feed, it is desirable to maximize the conversion to gasoline in the hydrocracker up to the availability of hydrogen. This reduces the production of light gases in the catalytic cracker, when the hydrocracker bottoms stream is combined with the fresh gas oil feed. However, the hydrocracker conversion should not exceed a certain maximum, if all the bottoms stream is combined with the fresh gas oil catalytic cracker feed. This maximum conversion is determined so that sufficient unconverted hydrocracker feed, the bottoms stream, is available to provide at least the 0.1 volume minimum concentration.

If deasphalted oil, coker gas oil, or other heavy feed is available as hydrocracker feed, it is desirable to minimize the conversion below feed initial. However, at least 30 volume percent of the heavy feed must be converted to products boiling below the initial boiling point of the feed. This 30 volume percent limitation is so that the hydrocracking conditions are severe enough to produce a substantially hydrogenated bottoms.

It is necessary to specify the minimum hydrogen consumption figures for each type of hydrocracking zone feed, in order to achieve substantially complete hydrogenation of aromatics. Table I gives the minimum hydrogen consumption and the preferred range, in order to convert at least 30% of the hydrocracking zone feed to products boiling below the initial boiling point of said feed.

Table I

Hydrocracking zone feed	Hydrogen consumption, s.c.f./bbl. feed	
	Minimum	Preferred range
Light straight-run gas oil and light and heavy catalytic cycle oil	750	750-1,600
Heavy straight-run and solvent deasphalted oil	1,000	1,250-2,500
Coker and pitch stripper gas oil	1,200	1,500-3,500

DRAWING

The present invention will be more clearly understood and further objects and advantages thereof will be apparent from the following description when read in connection with the accompanying drawing. The drawing is a simplified flow diagram illustrating process units and flow paths suitable for carrying out the process of the present invention.

PROCESS UNITS AND OPERATING CONDITIONS, GENERAL

The process of the present invention includes the combination of a catalytic cracking zone and a hydrocracking zone. The latter may be either a one- or two-stage system with the necessary source of hydrogen. Suitable catalysts and operating conditions for these two zones are described below. Included, within each of the two zones, are conventional means for separating the effluent from the catalytic cracking and hydrocracking zone into various products. The usual means of achieving the desired separation is to use fractionating columns operating at conventional conditions.

The present invention makes use of feed preparation units, such as a solvent deasphalting unit, a delayed or

fluid coking unit, and a thermal viscosity-breaking or recycle-cracking unit with auxiliary pitch stripping units. As these units do not form a specific part of the invention, their detailed description will not be included. However, it is intended that these units will operate under conventional conditions used in the petroleum industry.

CATALYTIC CRACKING ZONE

The catalytic cracking zone fresh feed stock may be any of the conventional types such as, for example, a straight-run gas oil obtained from a crude distillation column. The boiling range of the feed stock is generally between about 400° and 1100° F.

The catalytic cracking zone can be operated at conventional catalytic cracking conditions and may employ fluidized beds or moving compact beds of solid catalysts. For example, a temperature of about 850° to 1000° F., a pressure of about 10 to 50 p.s.i.g. and at a 30% to 80% per-pass conversion, with a conventional catalyst such as silica-alumina, silica-zirconia, silica-magnesia, or the like and natural and treated clays and combinations thereof. The catalytic cracking zone is preferably operated above 50% per pass conversion when gasoline is to be maximized and below 50% per pass conversion when middle distillates are to be maximized.

HYDROCRACKING ZONE

The feed to the hydrocracking zone may be any suitable gas oil such as straight-run gas oil, deasphalted oil, coker gas oil, pitch stripper gas oil, or catalytic cycle oil. In the case in which gasoline is the only major product, the hydrocracking zone feed should have an initial boiling point above about 300° F. In the case in which both middle distillates and gasoline are to be produced, the feed should have a substantial portion boiling above 650° F. In both cases, the feed to the hydrocracking zone should contain at least 20% aromatics.

Depending upon the particular hydrocracking catalyst used, the hydrocracking zone feed may contain relatively large quantities of nitrogen. There is no nitrogen limit on the hydrocracking zone bottoms which are passed to the catalytic cracking zone, although under the severe hydrogenative conditions employed, the nitrogen is reduced to very low levels.

In general, the hydrocracking zone operation becomes much more economical with feeds containing less than 200 p.p.m., preferably below 100 p.p.m., and much more preferably below 10 p.p.m. of nitrogen. A low level of nitrogen in the feed permits the hydrocracking reaction to be conducted at lower temperatures than feeds containing larger amounts of nitrogen compounds.

In cases in which the hydrocracking zone feeds are not inherently low in nitrogen, the feed should be pretreated by a suitable denitrification process. The given feed may be treated with hydrogen in the presence of a sulfactive hydrogenation catalyst, e.g. Group VI and VIII metals or compounds thereof on alumina or other support, at elevated temperatures and pressures, to remove nitrogen compounds therefrom.

The usual hydrocracking zone operating conditions comprise from about 2000 to 30,000 s.c.f. hydrogen/bbl. of total feed and, preferably, from about 2000 to 15,000, at an LHSV of from about 0.2 to 15 and, preferably, from about 0.4 to 3.0, at a pressure of at least 1000 p.s.i.g. and, preferably, from about 1000 to 3000 p.s.i.g., and a temperature in the range of from about 400° to 950° F. The preferred initial on-stream temperature is from about 500° to 650° F., with progressive increase to about 750° to 950° F., so as to maintain catalyst activity at a controlled level.

The catalyst employed in the hydrocracking zone is one wherein a material having a hydrogenating-dehydrogenating activity is deposited or otherwise combined with a catalyst support. The cracking component may comprise any one or more non-acidic, weakly acidic or strongly

acidic materials such as silica, alumina, bauxite, silica-alumina, silica-magnesia, silica-alumina-zirconia and the like, as well as various acid-treated clays and similar materials. The hydrogenating-dehydrogenating component of the catalyst can be selected from any one or more of the various groups VI, VII and VIII metals, as well as the oxides and sulfides thereof, alone or together with promoters and stabilizers that may have, by themselves, negligible catalytic effect. Examples of suitable hydrogenating-dehydrogenating components are the oxides and sulfides of molybdenum, tungsten, vanadium, chromium, and the like, as well as metals such as iron, nickel, cobalt and platinum. More than one hydrogenating-dehydrogenating component may be present, and favorable results may be obtained with catalysts containing composites of two or more of the oxides of molybdenum, cobalt, chromium, tungsten, nickel, tin and zinc, and with mixtures of said oxides with fluorine. The amount of the hydrogenating-dehydrogenating component can be varied within wide limits from about 0.5 to 30% based on the weight of the entire catalyst.

The proper selection of operating conditions and catalysts should be correlated with the particular type of hydrocracking feed stock used in order to convert at least 30% of the fresh feed to products boiling below the initial boiling point of the feed. In general, an acidic catalyst such as nickel sulfide and/or cobalt sulfide combined with such acidic materials as conventional cracking catalysts containing silica-alumina, silica-alumina-zirconia, acid treated clays and the like is used for hydrocracking the hydrofined light straight-run and catalytic cycle oils; a non-acidic catalyst such as a combination of cobalt sulfide and/or nickel sulfide with molybdenum sulfide and with such non-acidic materials as alumina (preferred), bauxite, silica, and zirconia is used for hydrocracking the coker and pitch stripper gas oils; and a weakly acidic catalyst such as a combination of nickel sulfide and/or molybdenum sulfide with tungsten sulfide and with such weakly acidic materials as silica-magnesia is used for hydrocracking heavy straight-run and deasphalted oils.

DESCRIPTION OF DRAWING

Referring now to the drawing, a reduced crude oil feed is supplied through line 1 to vacuum distillation column 2. A light gas oil is removed from the overhead of column 2 through line 3. A heavy gas oil is removed as a side cut from column 2 through line 4. Residuum is removed from the bottom of column 2 and withdrawn for subsequent processing through line 5. The light gas oil in line 3 is passed through line 6 to hydrocracking zone 7. Gasoline and lighter fractions comprising C₄-hydrocarbons are removed from zone 7 and withdrawn from the system through line 8. Middle distillates are removed from zone 7 in the event they are produced and withdrawn from the system through line 9. A substantially hydrogenated bottoms stream is removed from zone 7 through line 10 and passed through line 11. A portion of the bottoms stream may be withdrawn from the system through line 12. The heavy gas oil in line 4 is combined with the substantially hydrogenated bottoms stream in line 11 and passed through line 15 to catalytic cracking zone 16. Gasoline and lighter fractions are removed from zone 16 and withdrawn from the system through line 17. Light cycle oil is removed from zone 16 through line 18. Heavy cycle oil is removed from catalytic cracking zone 16 through line 19. The light cycle oil in line 18 may be withdrawn from the system or passed through lines 20, 21 and 6 to hydrocracking zone 7. Heavy cycle oil in line 19 may be withdrawn from the system or separately passed through lines 22, 21 and 6 to hydrocracking zone 7 or combined with the light cycle oil in line 20 and passed to the hydrocracker.

Alternatively, the light gas oil in line 3 may be passed through line 23 and combined with heavy gas oil in line 4. The combined light and heavy gas oil may be passed

through line 4, combined with the substantially hydrogenated bottoms stream in line 11, passed to catalytic cracking zone 16, and catalytically cracked into gasoline and lighter and light and heavy cycle oil. In this alternative, the light cycle oil and/or the heavy cycle oil is passed to hydrocracking zone 7 and hydrocracked into the product fractions indicated in the preceding paragraph.

One method of processing the residuum in line 5 is to pass the residuum through line 25 to solvent deasphalting zone 26. A deasphalted oil is removed from zone 26 through line 27. Asphalt is removed from zone 26 and withdrawn from the system through line 28. In this method, the deasphalted oil in line 27 is passed through line 30 and line 6 to hydrocracking zone 7 and hydrocracked into the fractions indicated above. A portion of the deasphalted oil may be withdrawn from the system through line 31.

Another method of processing the residuum in line 5 is to pass the residuum through line 34 to coking zone 35. Gasoline and lighter fractions are removed from zone 35 and withdrawn from the system through line 36. Coker distillate is removed from zone 35 through line 37. Coke is removed from zone 35 and withdrawn from the system through line 38. In this method, the coker distillate in line 37 is passed through line 30 and line 6 to hydrocracking zone 7. A portion of the coker distillate may be withdrawn from the system through line 39.

Still another method of processing the residuum in line 5 is to pass the residuum through line 40 to thermal cracking zone 41. The effluent from thermal cracking zone 41 is passed through line 42 to pitch stripping zone 43. A gas oil is removed from zone 43 through line 44. Pitch is removed from zone 43 and withdrawn from the system through line 45. In this method, the gas oil in line 44 is passed through line 30 and line 6 to hydrocracking zone 7. A portion of the gas oil may be withdrawn from the system through line 46.

EXAMPLES

To demonstrate the advantages of substantially hydrogenating the hydrocracker bottoms, compared to partially hydrogenating the bottoms, and the effectiveness of the more severe hydrogenation on the subsequent catalytic cracking operations, two pilot plant experiments were performed as described in Example Nos. 1 and 2.

Example No. 1.—A 40.8° API West Texas sweet crude was distilled to obtain the following fractions:

Table II

	Stock A Light gas oil	Stock B Heavy vacuum gas oil
Boiling range, ° F.....	400-720	700-1,020
Gravity, ° API.....	35.6	29.5
Aniline point, ° F.....	127	186
Sulfur, wt. percent.....	0.38	0.65
Nitrogen, p.p.m.....	400	650
Aromatics, vol. percent.....	28	33

The light gas oil was first processed in a 2-stage hydrocracking unit. A first stage was provided to reduce the total nitrogen in the feed to below about 0.5 p.p.m. by hydrofining over a nickel sulfide (6½ wt. percent Ni) and molybdenum sulfide (22 wt. percent Mo) on activated alumina, an acid catalyst support. The first stage effluent was scrubbed with freshly distilled water to remove hydrogen sulfide and ammonia and then was fed to a second stage or hydrocracking unit using nickel sulfide (6 weight percent Ni) on silica-alumina, an active acidic cracking catalyst support. The following operating conditions were maintained in the second or hydrocracking stage to convert 70 volume percent of feed

to products boiling below 400° F. on a once-through basis:

Table III

Space velocity (volume of oil per hour per volume of catalyst) --	1.5.
Pressure	1500 p.s.i.g.
Catalyst temperature	500-700° F.
Hydrogen consumption	1400 s.c.f./bbl. of raw feed to first stage.

The following yields, based on raw feed, were obtained:

Table IV

	Vol. percent
Propane minus	2.0
Butanes	9.5
Light gasoline (C ₅ -180° F.)	20.0
Heavy gasoline (180-400° F.)	47.9
400° F.+ bottoms	30.0

The properties of the hydrocracker bottoms, compared to the original raw feed, are as follows:

Table V

	Raw feed to hydrocracker	400° F.+ hydrocracker bottoms
Gravity, ° API.....	35.6	40.0
Aniline point, ° F.....	127	160
Nitrogen, p.p.m.....	400	0.2
Sulfur, wt. percent.....	0.40	Nil
Aromatics, vol. percent.....	28	3
Boiling range, ° F.....	400-720	400-690

A series of catalytic cracking runs were made in a laboratory fluid bed catalyst testing unit, using commercial plant synthetic silica-alumina equilibrium catalyst of 30 activity (D+L). All runs were made at 900° F. catalyst temperature, 1.5 space velocity and catalyst to oil ratio of 1.5. In these runs the cracking cycle between regeneration was 12 minutes.

The following three stocks were catalytically cracked at constant conditions:

- The 700° to 1020° F. heavy vacuum gas oil—stock B
- The 400° F.+ hydrocracker bottoms from the 400° to 720° F. light gas oil—stock C
- A blend of 100 parts of heavy vacuum gas oil, stock B, and 11.5 parts of the 400° F.+ hydrocracker bottoms, stock C.

Table VI.—Pilot catalytic cracking once-through yields at 900° F.

	Stock B Heavy Vacuum Gas Oil	Stock C Hydrocracker 400° F.+ bottoms	Blend of 100 parts Stock B; 11.5 parts Stock C
Conversion, vol. percent below 430° F. (D+L).....	50	70	56.7
C ₃ as a gas, EFO LV, percent.....	6.0	7.0	6.2
Butanes, LV, percent.....	11.0	15.0	12.5
C ₅ , 430° F., LV, percent.....	40.0	56.0	45.4
430° F.+ cycle oil, LV, percent.....	50.0	30.0	43.3
Coke, wt. percent.....	3.8	2.8	3.5

The yield of products from a given feed stock is a function of conversion obtained at the specified temperature and operating conditions and with a specific catalyst. The conversions obtained under these specified conditions outlined above are a direct indication of the relative ease of cracking the various feed stocks. It will be noted that the 400° F.+ hydrocracker bottoms is much less refractory to cracking under the conditions of test compared to the heavy vacuum gas oil. Likewise, the conversion and yields of gasoline obtained on the blend of

heavy vacuum gas oil and hydrocracker bottoms are higher at the same conditions than would be estimated by a direct linear addition of the yields obtained on the components at the same operating conditions. A more meaningful comparison is to compare product distribution obtained from various feed stocks at constant conversion. At a constant conversion of 55 volume percent ($D+L$) the above cracking data shows the following product distribution:

Table VII

	Stock B	Stock C	Blend
Conversion, vol. percent ($D+L$).....	55.0	55.0	55.0
C_3 , as a gas, EFO LV, percent.....	8.5	5.0	5.8
Butanes, LV, percent.....	11.5	13.2	11.8
C_5 , 430° F. gasoline, LV, percent.....	42.5	45.1	44.6
Coke, wt. percent.....	5.2	1.2	3.3

The above data show that the addition of the severely hydrogenated hydrocracker bottoms in concentrations as low as about 0.1 volume bottoms per volume of heavy vacuum gas oil feed, caused an increase in conversion, reduction in coke production, an increase in gasoline production and an increase in gasoline efficiency that is greater than was obtained by processing the added hydrocracker bottoms alone at the same operating conditions. That is, it would be expected that the blend, composed 0.115 volume of stock C (10.3 vol. percent of total blend) and one volume stock B, would yield 42.8% gasoline and 4.8% coke instead of the observed 44.6% gasoline and 3.3% coke. Thus, it is as if 65% of stock C converted to gasoline, while producing no coke and further reduced coke production of stock B to 3.7%. With the reduction in the production of coke by the addition of hydrocracker bottoms, an appreciable increase in fresh feed rate to a commercial catalytic cracker would be possible in cases where the unit is limited in fresh feed capacity by the ability of the regenerator to burn the coke and to regenerate the catalyst. The increased gasoline yield following the increased conversion and the increased gasoline efficiency may indicate some form of hydrogen transfer from the hydrocracker bottoms, which may effectively correct the coke-forming reactions, thus permitting increased conversion and efficiency of conversion.

These surprising results are not obtained if the hydrocracker bottoms stream is not substantially hydrogenated, as would be the case if the hydrocracker feed were merely hydrofined with less than 30% conversion to lower boiling distillates, and as would also be the case if the hydrocracker feed were hydrocracked to greater than 30% conversion to lower boiling distillates but with insufficient hydrogen consumption, for example due to hydrocracking at too high a temperature or at too low a hydrogen partial pressure. This distinction exists even though such hydrofined or hydrocracked streams might themselves produce less coke and more gasoline than the usual catalytic cracker feed.

Example No. 2.—To illustrate the latter advantages of the process, the following results were obtained in comparing catalytic cracking of heavy gas oils ordinarily fed to a catalytic cracker (case I) with catalytic cracking of a blend composed 65% of said gas oils and 35% of incompletely hydrogenated bottoms from hydrofining deasphalted oil at hydrogen consumption of 500 s.c.f./bbl. and 15% conversion below feed initial over cobalt (2 weight percent)—molybdenum (7 weight percent)—alumina (case II) and comparing with a preferred embodiment of the present invention, where the catalytic cracking of a blend composed of 65% of said gas oils and 35% of hydrogenated bottoms from hydrocracking the deasphalted oil at hydrogen consumption of 2000 s.c.f./bbl. and 50% conversion below feed initial (case III) over nickel sulfide (14 weight percent Ni) and tungsten sulfide (14 weight percent W) on silica-magnesia (30

weight percent MgO based on the support), a weakly acidic catalyst support.

These are pilot plant data at identical conditions of temperature (950° F.), cycle length (5 minutes), catalyst/oil ratio (3.7), and space velocity (3.2), with no recycle oil.

Table VIII

Case.....	I	II	III
Conversion, vol. percent.....	40	41	50
Coke, wt. percent.....	4.1	3.9	3.3

Although conversion increased slightly, and coke decreased, with the inclusion of hydrofined bottoms in case II, the improvement is small and reflects only that the hydrofined oil is a better feed for catalytic cracking than the usual heavy gas oils. The results are only what would be expected. That is, the total feed to the cracker cannot be increased significantly and the hydrofined bottoms obtained at considerable expense merely displaces a substantially equivalent amount of heavy gas oil ordinarily in the feed. The displaced heavy gas oil can only be disposed of as low value fuel oil. In contrast, the results in case III show that it would be possible to add the substantially hydrogenated hydrocracker bottoms to the ordinary heavy gas oil feed without displacing an equivalent amount of heavy gas oil. Consequently, the total feed rate can be increased while still obtaining greater conversion with less coke make.

Now, to illustrate various embodiments of the invention as they may be practiced commercially and to show the advantages obtained thereby as compared to known and proposed processes, Examples Nos. 3–6 are presented:

Example No. 3.—In one of the preferred embodiments of the present invention, an 80,000 b.p.s.d. Mid-Continent refinery processes the same West Texas crude referred to in Example No. 1. A conventional FCCU (fluidized catalytic cracking unit) in the refinery, which processes 23,000 b.p.d. of heavy vacuum gas oil at 70% conversion, has a coke burning capacity of 25,000 pounds per operating hour and a total reactor feed capacity of 37,500 b.p.s.d. The refinery also includes a two-stage catalytic reformer of 22,000 b.p.s.d. capacity when operating to a reformate octane of 99 F–1+3 ml. T.E.L. per gallon and an alkylation plant to process mixed C_3 and C_4 olefins from the FCCU.

Reduced crude from the bottom of the atmospheric distillation column in the refinery is passed through a heater to a vacuum column where the following fractions and bottoms are produced:

	B.p.s.d.	Boiling range, ° F.
Light vacuum gas oil.....	10,000	400–720
Heavy vacuum gas oil.....	23,000	700–1020
Vacuum residuum.....	3,000	1020+

This example compares the disposal of 10,000 b.p.s.d. of the surplus light vacuum gas oil either as distress No. 2 oil at a low profit (base case IV), with conversion of the vacuum gas oil to gasoline in a recycle two-stage hydrocracking unit (case V), and with one of the preferred embodiments of the present invention, comprising once-through hydrocracking the light vacuum gas oil and blending the substantially hydrogenated unconverted bottoms with the heavy vacuum gas oil feed to the FCCU (case VI).

In case VI the light vacuum gas oil having an aromatics content of 28 percent is contacted in a two-stage hydrocracking unit. The first-stage hydrofiner operates at 650° to 850° F., a hydrogen partial pressure of 1000 to 2500 p.s.i.g., an LHSV of 0.4 to 1.0 with a sulfided non-acidic catalyst, i.e. nickel-molybdenum on alumina as used in Example No. 1. The second-stage hydrocracker operates with the same acidic catalyst and at the same conditions

as shown in Table III of Example No. 1. The hydrogen consumption is 1,250 s.c.f./bbl. of fresh feed. The properties of the hydrocracker bottoms compared to the raw feed are shown in Table V of Example No. 1. The hydrocracker zone effluent is separated by means of a fractionating column to give the following products:

	B.p.s.d.
Gas as EFO -----	200
Isobutane -----	670
Normal butane -----	280
Light gasoline (C ₅ -180° F.) -----	2,000
Heavy gasoline (180-400° F.) -----	4,790
Bottoms 400° F.+ -----	3,000

The isobutane is utilized in the alkylation plant, the normal butane and light gasoline to gasoline blending and the heavy gasoline is reformed in a new catalytic reformer. In this case no hydrogen plant is required since there is sufficient hydrogen from the existing reformer operating on straight-run naphtha and the new reformer operating on the heavy hydrocracked naphtha. The hydrocracker bottoms is combined with the 23,000 b.p.s.d. of heavy vacuum gas oil from the vacuum column of the crude unit to give a combined fresh feed stream to the catalytic cracker of 26,000 b.p.s.d. This increased feed to the catalytic cracker is processed at 75 percent conversion without exceeding the coke burning capacity of 25,000 lbs. per operating hour. The catalytic cracker effluent is passed to a fractionating column where light gas is separated, a C₃ and C₄ cut is separated and sent to the alkylation plant and the C₅-430° F. whole gasoline sent to gasoline blending. A portion of the light cycle oil is blended with straight-run middle distillate and the balance of the cyclic oil is blended with the vacuum residuum to produce bunker fuel oil.

Table IX.—Example No. 3

	Case IV	Case V	Case VI
Refinery products, b.p.s.d.:			
Refinery fuel gas, as EFO -----	4,000	3,920	4,200
Motor gasoline (10 lb. R.V.P.) (97 F-1+3 ml. T.E.L./gal) -----	47,665	56,770	57,780
No. 2 heating oil -----	13,000	13,000	13,000
Bunker fuel oil (150 SSF/122°F.) -----	6,900	6,900	6,900
Distress No. 2 oil -----	10,000		
Total refinery products -----	81,565	80,670	81,880
New refinery equipment added:			
2-stage hydrocracker (for conversion of light vacuum oil) -----			
1st stage, b.p.s.d. -----		10,000	10,000
2d stage, b.p.s.d. -----		16,665	10,000
Conversion below 400° F., overall, percent -----		100	70
Single stage reformer -----		7,200	4,200
Hydrogen plant, million s.c.f./d. -----		5.0	
FCC operation:			
Fresh feed, b.p.s.d. -----			
Heavy vacuum gas oil -----	23,000	23,000	23,000
Hydrocracker bottoms -----			3,000
Total fresh feed -----	23,000	23,000	26,000
Heart cut recycle -----	14,500	14,500	11,500
Average reactor temp., °F. -----	900	900	900
Coke burning rate, lbs./hr. -----	25,000	25,000	24,000
Conversion, vol. percent -----	70	70	75
Conversion barrels -----	16,100	16,100	19,500
C ₅ , 430° F. catalytic gasoline, b.p.s.d. -----	10,700	10,700	13,300
Gasoline efficiency:			
Gasoline b.p.s.d. -----			
Conversion Barrels ×100 -----	66.5	66.5	68.2

Compared to the base case IV, the addition of the once-through hydrocracking in case VI by adding 3,000 b.p.s.d. of hydrocracker bottoms to the FCC fresh feed results in an increase fresh feed rate of about 11% on the FCCU, an increase in conversion, a decrease in coke production, a reduction in heart cut recycle rate and an increase in gasoline, C₃ and C₄ production over and above the volume that would be produced if the hydrocracker bottoms

had been separately catalytically cracked at the same conditions. The increase in conversion and conversion barrels is accompanied by an increase in gasoline efficiency.

Considerable savings are realized in the conversion of the 10,000 b.p.s.d. by case VI which is one of the preferred embodiments of the present invention over the installation of a recycle hydrocracker to convert the gas oil to gasoline by extinction recycle (case V). The second stage of the hydrocracker is reduced in size, 10,000 b.p.s.d. versus 16,665 b.p.s.d. A smaller single stage catalytic reformer to process the hydrocracked gasoline is required; 4200 b.p.s.d. versus 7200 b.p.s.d. and a hydrogen plant is not required in contrast to the 5 million s.c.f./d. unit that is required in case V. In addition, case VI on an overall refinery basis produces about 1,000 b.p.s.d. more gasoline than case V.

Example No. 4.—Utilizing the same 80,000 b.p.s.d. West Texas crude referred to in Example No. 1, another method of disposing of the 10,000 b.p.s.d. of surplus light vacuum gas oil is illustrated. In this example, the existing FCCU now processes both the light and heavy vacuum gas oil at reduced conversion. In the base case VII, the 10,000 b.p.s.d. of light vacuum gas oil is processed along with the 23,000 b.p.s.d. of heavy vacuum gas oil at 55% conversion. The excess cycle oils over and above fuel oil requirements must again be disposed of as distress No. 2 or heavier fuel oils. In another case (case VIII) the 8,000 b.p.s.d. of distress cycle oil is processed in a two-stage hydrocracker to extinction recycle for gasoline production. This is compared with another preferred embodiment of the present invention (case IX) where the excess cycle oil, 8,000 b.p.s.d., is once-through hydrocracked at 44.5% conversion with the bottoms boiling above feed initial being processed in the FCCU at an overall increase in catalytic conversion.

The properties of this cycle oil are:

Gravity -----	26.8° API.
Aniline point -----	130° F.
Sulfur -----	0.50 wt. percent.
Nitrogen -----	200 p.p.m.
Aromatics -----	45 vol. percent.
Boiling range -----	400-850° F.

Processing the 8,000 b.p.s.d. cycle oil to gasoline in the two-stage hydrocracker would require an 8,000 b.p.s.d. first-stage along with a 13,335 b.p.s.d. second-stage operating at 60 percent conversion below 400° F., with extinction cycle to gasoline, a 4.0 million s.c.f./d. hydrogen plant and a new catalytic reformer of 5,800 b.p.s.d. capacity to process the heavy gasoline from the hydrocracker.

In case IX, the light and heavy vacuum gas oils are charged to the FCCU as explained above. In this preferred method of operation, 8,000 b.p.s.d. of catalytic cycle oil are processed through a two-stage hydrocracker on a once-through basis at 44.5 percent conversion below 400° F. The feed to the first and second stages is 8,000 b.p.s.d. The operating conditions and the catalysts for the two stages are the same as in Example No. 3. Hydrogen consumption is 1,600 standard cubic feet per barrel of raw feed and under these circumstances the existing and a new 2,880 b.p.s.d. catalytic reformer are sufficient to supply hydrogen for the hydrocracker. The effluent from the hydrocracker reactor is separated into the following products:

	B.p.s.d.
Gas as EFO -----	100
Isobutane -----	380
Normal butane -----	160
Light gasoline (C ₅ -180° F.) -----	900
Heavy gasoline (180-400° F.) -----	2,880
Bottoms (400° F.+) -----	4,500

The inspections on the hydrocracker bottoms compared to hydrocracker feed are as follows:

	Hydrocracker feed	Hydrocracker 400+ bottoms
Gravity, ° API.....	26.8	34.0
Aniline point, ° F.....	130	160
Sulfur, wt. percent.....	0.45	Nil
Nitrogen, p.p.m.....	200	0.2
Aromatics, vol. percent.....	45	12
Boiling range, ° F.....	430-800	430-750

In case IX, once-through hydrocracking the 8,000 b.p.s.d. of cycle oil at a conversion of 44.5 volume percent to materials below feed initial (400° F.) and adding the hydrocracker bottoms to the FCCU fresh feed gives an increase of about 13% in total fresh feed rate, a decrease in coke, an increase in conversion, and higher gasoline, C₃ and C₄ production over and above the volume that would be realized if the hydrocracker bottoms had been separately cracked at the same conditions. Again the increase in conversion barrels is accompanied by an increase in gasoline efficiency.

As in case VI of Example No. 3, considerable equipment savings are realized by case IX over the recycle hydrocracker case (case VIII) in the reduced size of the second stage of the hydrocracker, the single-stage reformer and the hydrogen plant. In addition, case IX on an overall refinery basis produces about 1400 b.p.s.d. more gasoline over the more extensive extinction recycle hydrocracking of case VIII.

Table X.—Example No. 4

	Case VII	Case VIII	Case IX
Refinery products, b.p.s.d.:			
Refinery fuel gas, as EFO	4,120	4,400	4,040
Motor gasoline (10 lb. R.V.P.) (97 F-1+3 ml. T.E.L./gal.)	49,100	57,930	59,310
No. 2 heating oil	13,000	13,000	13,000
Bunker fuel oil (150 SSF/122° F.)	6,900	6,900	6,900
Distress cycle oil or No. 2 oil	8,000		
Total refinery products	81,120	82,230	83,250
New refinery equipment added:			
2-stage hydrocracker (for conversion of cycle oil)—			
1st stage, b.p.s.d.		8,000	8,000
2nd stage, b.p.s.d.		13,335	8,000
Conversion below 400° F., overall, percent.		100	44.5
Single stage reformer		5,800	2,880
Hydrogen plant, million s.c.f./d.		4.0	
FCC operation:			
Fresh feed, b.p.s.d.—			
Heavy vacuum gas oil	23,000	23,000	23,000
Light vacuum gas oil	10,000	10,000	10,000
Hydrocracker bottoms			4,500
Total fresh feed	33,000	33,000	37,500
Heart cut recycle	4,500	4,500	
Average reactor temp., ° F.	900	900	900
Coke burning rate, lbs./hr.	25,000	25,000	24,000
Conversion, vol. percent.	55.0	55.0	60.0
Conversion barrels	18,150	18,150	22,500
C ₃ , 430° F. catalytic gasoline, b.p.s.d.	12,600	12,600	15,000
Gasoline efficiency	69.4	69.4	70.2

Example No. 5.—A California refinery processing 100,000 b.p.s.d. of 21.0° API, San Joaquin Valley, California crude oil and equipped with a conventional fluid catalytic cracking unit is selected as another illustrative example of the application of the present invention. The catalytic cracking unit has a coke burning capacity of 36,000 pounds per operating hour. Maximum reactor throughput is 42,000 b.p.s.d. at an average reactor temperature on the high nitrogen California stocks of 950° F. The total reactor feed comprises 30,000 b.p.s.d. of heavy California vacuum gas oil at 63 volume percent conversion, and a heart cut recycle of 12,000 b.p.s.d. Unconverted cycle oil from the catalytic cracker along with vacuum column pitch is processed in a single coil thermal cracker. The balance of the pitch, 5,000 b.p.s.d., is utilized for asphalt production. The refinery is equipped with a two-stage

catalytic reformer for upgrading the octane numbers of heavy straight-run naphtha and dehexanized thermal naphtha. Capacity of the catalytic reformer is 12,000 b.p.s.d. when operating at an average reformate severity octane of 97 F-1+3 ml. T.E.L. per gallon. The mixed C₃ and C₄ olefins from the catalytic and thermal cracker are alkylated for the production of whole alkylate for motor gasoline.

In this example, it is desired to reduce the production of bunker fuel oil from the current operation by the installation of a fluid coking unit to process the 25,000 b.p.s.d. of vacuum residuum. Utilizing a recycle ratio of 35-45 percent of high boiling (over 1025° F.) coker products and a catalyst temperature of 950° F. results in the production of 10,000 b.p.s.d. of heavy coker gas oil boiling from 620-1025° F. which is added to the straight-run gas oil as feed stock in the FCCU (case X). After the installation of the fluid coker, the lighter coker gas oil (350-620° F.) and the light catalytic cycle oil along with the medium cycle oil (630-750° F.) are processed in a separate hydrocracker to gasoline. The heavy catalytic cycle oil (750° F.+) is disposed of as bunker fuel oil.

Operating in still another preferred embodiment of the present invention (case XI), the 10,000 b.p.s.d. of heavy coker gas oil is rerun, preferably in a vacuum column to give 6,500 b.p.s.d. of overhead of 900° F. end point and 3,500 b.p.s.d. of bottoms having the following inspections:

Gravity	10.0° API.
Aniline point	142° F.
Carbon residue	3.6 wt. percent.
Nitrogen	13,000 p.p.m.
Sulfur	3.5 wt. percent.
Aromatics	48 vol. percent.
Boiling range	900-1025° F.

This material is diluted with 2,500 b.p.s.d. of heavy catalytic cycle oil of 650-850° F. boiling range and processed in a once-through single-stage hydrocracker over a sulfided nonacidic catalyst such as the nickel-molybdenum on alumina used in Example No. 1 at 650-725° F., 2,000 p.s.i.g., a hydrogen consumption of 1,500 s.c.f./bbl. of feed, and over 30 percent conversion below feed initial. This will result in denitrification, desulfurization and aromatics saturation. Topping out the material boiling below 430° F., there will result 5,500 b.p.s.d. of hydrocracker bottoms of the following inspections:

Gravity	27.0° API.
Aniline point	190° F.
Aromatics	3.0 vol. percent.
Nitrogen	1.0 p.p.m.
Sulfur	Nil.
Boiling range	440-990° F.

The analyses of the FCC operations, which demonstrate the essence of the present invention, clearly show the value of adding the hydrocracker bottoms to the FCC fresh feed in case X in an amount about equal to 17.5 percent of the whole fresh feed to the unit. Thus, in attempting to maximize reduction of coker fuel oil in the refinery by the addition of the fluid coker in case X and including all the heavy coker gas oil (620-1025° F.) in the FCC feed results in a loss of conversion barrels and 10 lb. R.V.P. gasoline production. Removing the highly refractive portion of the heavy coker gas oil (900-1025° F.) by vacuum distillation results in an increase in conversion barrels but no substantial improvement in gasoline production. However, in case XI, hydrocracking the heavy end of the coker gas oil (920-1025° F.) in a once-through single-stage hydrocracker at about 35 percent conversion to materials boiling below the feed stocks initial, and adding the hydrocracker bottoms to the FCC fresh feed results in a large increase in conversion barrels and 10 lb. R.V.P. gasoline produc-

tion. This increase in conversion and gasoline production and decrease in catalytic coke production is greater than would have been realized if the hydrocracker bottoms from the heavy coker gas oil had been catalytically cracked at the same conditions by itself.

Table XI.—Example No. 5

	Case X	Case XI
Refinery products, b.p.s.d.:		
Refinery fuel gas, as EFO	9,930	11,900
Motor gasolines (10 lb. R.V.P.) (96 F-1+3 ml. T.E.L./gal.)	63,000	70,050
Diesel fuel oil	20,000	20,000
Bunker fuel oil (225 SSF/122° F.)	5,800	
Coke, tons/D./as EFO	750/3750	750/3750
Asphalt	5,000	5,000
Total Refinery Products	107,480	110,700
New refinery equipment added, b.p.s.d.:		
Part 1—		
Fluid coker, fresh feed	25,000	25,000
Single Stage Hydrocracker:		
Heavy coker gas oil		3,500
Heavy catalytic cycle oil		2,500
		6,000
Part 2—		
Two-stage hydrocracker (for conversion of light coker and cycle oils to gasoline)	19,000	15,000
Single stage reformer	13,000	10,000
Hydrogen plant, million s.c.f./d.	23.0	21.0
FCC operation:		
Fresh feed, b.p.s.d.:		
Heavy vacuum gas oil	30,000	30,000
Heavy coker gas oil (620-1025° F.)	10,000	
Rerun heavy coker gas oil (620-900° F.)		6,500
Hydrocracker bottoms		5,500
Total fresh feed	40,000	42,000
Heart cut recycle		
Total reactor feed	40,000	42,000
Average reactor temp., °F.	950	940
Coke burning rate, lbs./hr.	36,500	34,000
Conversion, vol. percent.	46.8	60
Conversion barrels	18,720	25,200
C ₅ , 430° F. catalytic gasoline b.p.s.d.	13,170	18,400
Gasoline efficiency, percent.	70.3	73.0

Example No. 6.—In still another preferred embodiment of the present invention, the 100,000 b.p.s.d refinery outlined in Example No. 5 is taken as the basis for an additional illustrative example. The refinery operates a 30,000 b.p.s.d. FCCU at 63 percent conversion, 950° F. reactor temperature and 40 percent heart cut recycle as represented by case XII. The single coil thermal cracking unit operates at a 2.0:1 recycle ratio to fresh feed, the recycle boiling range being 450-750° F. The cracked tar contains about 50 percent of unconverted gas oil boiling in the range of 550-1000° F. This gas oil, which is a mixture of unconverted catalytic and thermal gas oil, is quite refractory to further thermal cracking.

To reduce fuel oil in the refinery, a vacuum pitch stripper is utilized to strip the cracked tar to 550-1000° F. pitch stripper gas oil and an equal amount of 225° F. softening point pitch as indicated in case XIII. The inspections on the pitch stripper gas oil are:

Gravity	11.5° API.
Aniline point	117° F.
Carbon residue	1.60 wt. percent.
Sulfur	1.50 wt. percent.
Nitrogen	7,000 p.p.m.
Heavy metals:	
Ni	1.8 p.p.m.
V	0.1 p.p.m.

In case XIV, a preferred embodiment of the present invention, the pitch stripper gas oil is split to give about 40 percent overhead or 4,700 b.p.s.d. and 7,050 b.p.s.d. of an 825° F.+ heavier cut and this heavier cut is processed in a single-stage hydrocracking unit. This unit operates with the same nonacidic catalyst as in Example No. 1 at 650-725° F., 2,000 p.s.i.g., and a hydrogen consumption of 1,500 s.c.f./bbl. of feed and about 35 per-

cent conversion to materials below the feed initial (825° F.). This will result in denitrification, desulfurization and aromatics saturation. After topping out the material boiling up to 430° F. there will result 6,500 b.p.s.d. of hydrocracked bottoms having the following inspections:

Gravity	22.5° API.
Aniline point	175° F.
Aromatics	8 vol. percent.
Nitrogen	1.5 p.p.m.
Sulfur	0.005 wt. percent.
Boiling range	450-985° F.

Table XII.—Example No. 6

	Case XII	Case XIII	Case XIV
Refinery products, b.p.s.d.:			
Refinery fuel gas, as EFO	7,700	5,500	6,385
Motor gasoline (10 lb. R.V.P.) (96 F-1+3 ml. T.E.L./gal.)	49,000	46,055	53,600
Diesel fuel oil	20,000	20,000	20,000
Bunker fuel oil (225 SSF/122° F.) (100 SSF/122° F.)	22,185	25,200	17,515
Asphalt	5,000	5,000	5,000
Total refinery products	103,865	101,755	102,500
New refinery equipment added, b.p.s.d.:			
Pitch stripper		23,000	23,000
Single stage hydrocracker for conversion of the pitch stripper gas oil			7,050
FCC operation:			
Fresh feed, b.p.s.d.—			
Heavy vacuum gas oil	30,000	30,000	30,000
Light pitch stripper gas oil			4,500
Hydrocracker pitch stripper gas oil bottoms			6,500
Pitch stripper gas oil		11,750	
Total fresh feed	30,000	41,750	41,000
Heart cut recycle	12,000		
Total reactor feed	42,000	41,750	41,000
Average reactor temp., °F.	950	950	950
Coke burning rate, lbs./hr.	36,000	36,000	35,500
Conversion, vol. percent.	63	45	60
Conversion barrels	18,900	18,785	24,600
C ₅ , 430° F. cat. gasoline	12,850	11,990	17,220
Gasoline efficiency, percent.	68	67	70

In case XIV, once-through hydrocracking the 7,050 b.p.s.d. of pitch stripper gas oil at a conversion of 35 volume percent to materials below the feed initial (825° F.) and adding the hydrocracker bottoms to the FCCU fresh feed results in large increase in conversion barrels and 10 lb. R.V.P. gasoline production over case XII. Again, this increase in conversion and gasoline production and decrease in catalytic coke production is greater than would have been realized if the hydrocracker bottoms from the pitch stripper gas oil had been catalytically cracked alone at the same conditions.

Specific modes of operation for the process of the present invention have been described but many variations could be made in those modes without departing from the spirit of this invention. However, such variations as fall within the scope of the appended claims are intended to be embraced thereby.

I claim:

1. In a catalytic cracking process wherein 30 to 80 volume percent of a first hydrocarbon feedstock other than substantially hydrogenated nonsynthetic materials from a hydrocracking zone is converted per pass in a catalytic cracking zone at particular catalytic cracking conditions including a pressure of about 10 to 50 p.s.i.g., a temperature of about 850° to 1000° F. and a maximum feed rate in volumes of said first feedstock per volume of catalyst per hour, to products including gasoline, with consequent coke laydown on the catalytic cracking catalyst and wherein said feed rate is at a maximum because said catalytic cracking zone is at the limits of its coke burning capacity, the method of increasing the gasoline

production rate without substantially changing said conditions which comprises:

- (a) continuing to convert said first hydrocarbon feedstock in said catalytic cracking zone at said maximum feed rate,
 - (b) adding to each volume of said first feedstock 0.1 to 1 volume of a second hydrocarbon feedstock obtained as hereinafter specified,
 - (c) hydrocracking in a hydrocracking zone a third hydrocarbon feedstock boiling above 300° F. and containing at least 20 volume percent aromatics at a per pass conversion to synthetic products of at least 30 volume percent with a hydrocracking catalyst comprising a cracking component consisting of silica, alumina, bauxite, silica-alumina, silica-magnesia, silica-alumina-zirconia, or acid-treated clays and a hydrogenating component consisting of the metals, oxides or sulfides of molybdenum, tungsten, vanadium, chromium, iron, nickel, cobalt or platinum or mixtures thereof, under hydrocracking conditions including a pressure of at least 1000 p.s.i.g., a temperature of about 400° to 950° F., a ratio of hydrogen to total feed to said hydrocracking zone of about 2,000 to 30,000 s.c.f. per barrel, an LHSV of about 0.2 to 15, said hydrocracking conditions selected to produce a net hydrogen consumption in said hydrocracking zone of at least 750 s.c.f. per barrel of said feedstock and to substantially hydrogenate the portion of said third feedstock not converted to said synthetic products, and
 - (d) using the necessary amount of said substantially hydrogenated portion as said second feedstock,
- whereby during said catalytic cracking together of said first and second hydrocarbon feedstocks the rate of coke laydown is maintained at no more than obtained during said catalytic cracking of said first feedstock alone, and the gasoline production rate from said catalytic cracking zone is greater than the combined gasoline production rate that would be obtained by separately catalyti-

cally cracking said first and second hydrocarbon feedstocks under said catalytic cracking conditions.

2. The method as in claim 1, wherein said third hydrocarbon feedstock is a straight-run gas oil which boils between about 350° and 750° F. and wherein the hydrocracking zone hydrogen consumption is between about 750 and 1,600 s.c.f. per barrel of said third feedstock.

3. The method as in claim 1, wherein said third hydrocarbon feedstock is a gas oil from a catalytic cracking zone which boils between about 350° and 900° F. and wherein the hydrocracking zone hydrogen consumption is between about 750 and 1,600 s.c.f. per barrel of said third feedstock.

4. The method as in claim 1, wherein said third hydrocarbon feedstock is a deasphalted oil from a solvent deasphalting zone and wherein the hydrocracking zone hydrogen consumption is between about 1,250 and 2,500 s.c.f. per barrel of said third feedstock.

5. The method as in claim 1, wherein said third hydrocarbon feedstock is a gas oil from a coking zone which boils between about 300° and 1100° F. and wherein the hydrocracking zone hydrogen consumption is between about 1,500 and 3,500 s.c.f. per barrel of said third feedstock.

6. The method as in claim 1, wherein said third hydrocarbon feedstock is a gas oil from a pitch stripping zone which boils between about 300° and 1100° F. and wherein the hydrocracking zone hydrogen consumption is between about 1,500 and 3,500 s.c.f. per barrel of said third feedstock.

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UNITED STATES PATENT OFFICE
CERTIFICATE OF CORRECTION

Patent No. 3,245,900

April 12, 1966

Norman J. Paterson

It is hereby certified that error appears in the above numbered patent requiring correction and that the said Letters Patent should read as corrected below.

Column 2, line 43, for "high-molecular eight" read -- high-molecular weight --; column 3, line 2, for "considerble" read -- considerable --; line 32, for "catalyst" read -- catalytic --; column 9, line 68, for "anon-acidic" read -- a non-acidic --; column 10, TABLE VI, first column, line 4 thereof, for "C₃," read -- C₃- --; same table, first column, line 8 thereof, for "C₅, 430° F." read -- C₅- 430° F. --; column 11, TABLE VII, first column, line 2 thereof, for "C₃," read -- C₃- --; same table, first column, line 4 thereof, for "C₅, 430°F." read -- C₅- 430° F. --; column 13, line 15, for "plaint" read -- plant --;

Signed and sealed this 15th day of August 1967.

(SEAL)
Attest:

EDWARD M. FLETCHER, JR.
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