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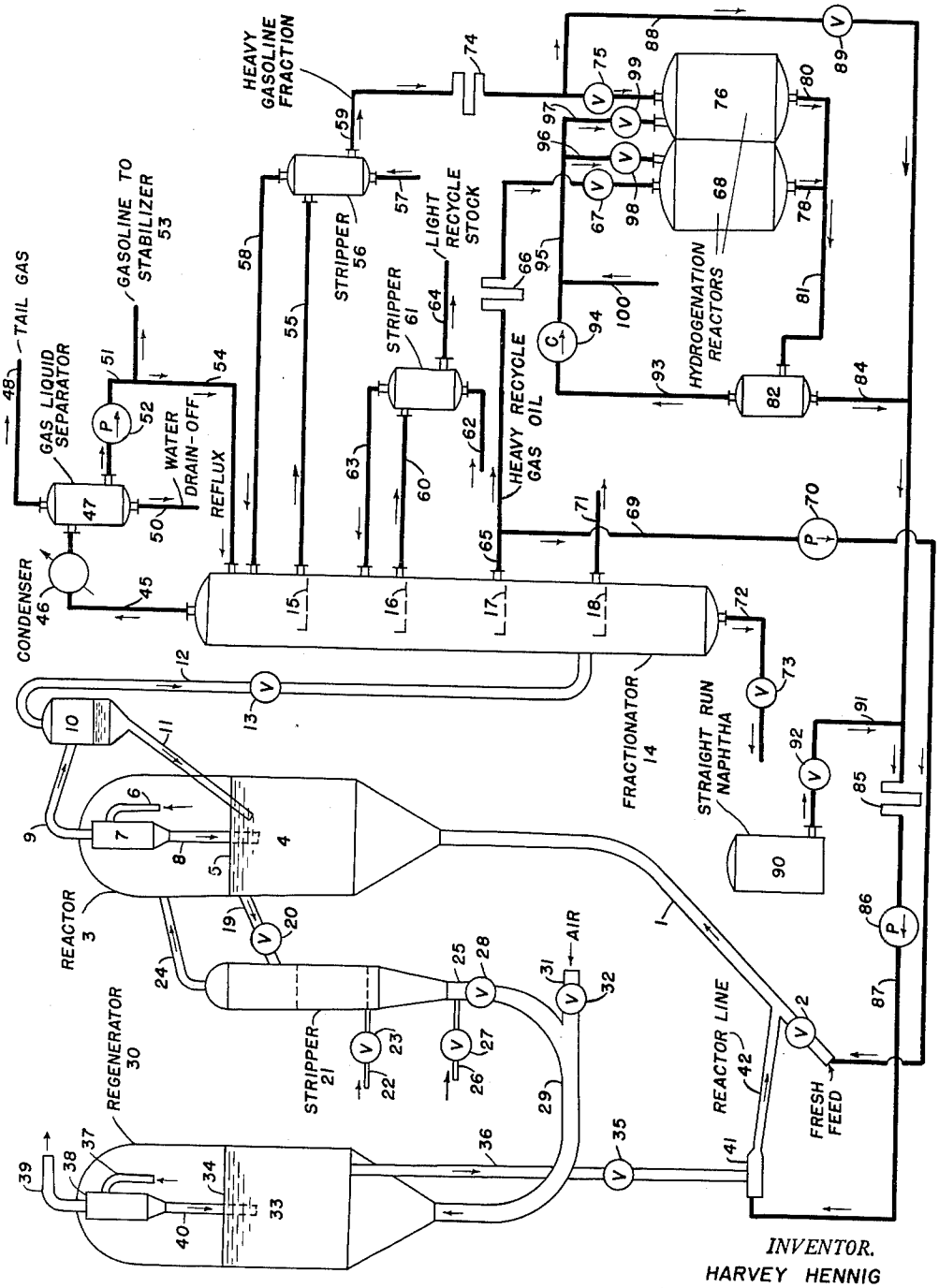

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CATALYTIC CRACKING PROCESS WITH THE PRODUCTION OF HIGH OCTANE GASOLINE

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The present invention relates to a process for converting hydrocarbon oil into substantial yields of high octane gasoline, and, more particularly, it relates to a method of converting cracking stocks in a fluidized cracking system, using the circulating-catalyst type of operation, wherein utilization is made of the heat content of the freshly regenerated catalyst as it leaves the regeneration zone, that is, before contact with incoming fresh feed, to supply the heat required to crack a mixture of a hydrogenated heavy recycle gasoline fraction and a heavy straight-run naphtha fraction (the mixture boiling in the range of about 350°–425° F., either hydrogenated or not), said heat supply being provided in a transfer-line reactor at a temperature above those maintained in the main cracking zone. In another aspect of the invention, a heavy gas-oil fraction boiling in the range of about 460° to 800° F. is recovered from the fractionator, separately hydrogenated at about 600° to 800° F. and mixed with a hydrogenated recycle gasoline fraction having a boiling range of about 350°–425° F., and the mixture is separately preheated to a temperature of about 450° to 800° F. and sent to said transfer-line reactor. In still another embodiment of the invention, the afore-mentioned mixture is mixed with a heavy-straight-run naphtha, having a boiling range of about 350° to 425° F., prior to preheating and transfer to said transfer-line reactor. Another variation of the invention comprises the process in which said heavy gas-oil and heavy recycle gasoline fractions are preheated and sent in admixture to a common hydrogenation reactor and the hydrogenated mixture is preheated and sent to said transfer-line reactor. In still another embodiment the heavy gasoline fraction is hydrogenated and sent to the catalyst transfer-line during the reaction. In still a further embodiment of the invention the heavy gasoline fraction may be separated into two parts and one part thereof hydrogenated and recycled with the unhydrogenated remainder. The invention also includes as another feature of this latter step the additional feature of adding a heavy straight-run naphtha to the recycled portions returning to the reactor. This invention is an improvement over the process disclosed in United States Patent 2,890,164, dated June 9, 1959, by Byron B. Woertz.

In the Woertz patent it is stated that modern internal combustion engines with ever-increasing compression ratios and power outputs require gasolines which burn quickly and efficiently, leaving a minimum of carbonaceous deposits in the combustion chamber. It is now known that certain heavy gasoline ends are responsible for much of the spark-plug fouling experienced today. Refiners have employed several expedients directed to overcoming this problem, including adjustment of the cracking conditions to fit the inherent characteristics of the charge

stock, and recycling certain fractions of the products to the main reaction zone. Thus, a feed hydrocarbon oil may be subjected to fractionation into light, medium, and heavy fractions, and each fraction separately subjected to cracking conditions optimum for the hydrocarbons concerned. In other methods, intermediate conversion products are separated from the cracked products in a fractionation zone and recycled back to the incoming charge-oil line wherein they are mixed directly with partially reactivated catalyst or catalyst which has been subjected to complete regeneration. In still other methods, the intermediate boiling fractions from the cracked products are subjected to cracking in a separate cracking zone using a portion of regenerated catalyst, and the products are either transmitted to a common fractionator, or separately fractioned.

The crude gasoline fractions resulting from the cracking of heavier crude oil fraction, such as virgin gas-oil, contain certain heavy gasoline ends, that is, material boiling between about 350–425° F., which are thermally stable and resistant to further cracking in the main reaction zone. In accordance with my invention, an additional advantage is gained through hydrogenating this material boiling between about 350–425° F., prior to separate contact with hot, regenerated catalyst, just prior to the entrance therein of the main oil charge. Heavy recycle gasoline fractions are separated from the cracked products, hydrogenated under conditions established to reduce the olefin content and partially saturate the aromatics, and the hydrogenated product is recycled to a hot catalyst transfer-line sufficiently far upstream from the point at which the fresh feed is charged to accomplish substantial cracking of the hydrogenated product between the points of feed entry.

As normally initially produced, the said heavy gasoline fraction may amount to about 4 to 10 volume percent of the fresh feed. If recycled without hydrogenation and in the conventional manner with the fresh feed, so little cracking takes place that the heavy recycle gasoline accumulates up to an extent of about 20 to 40 volume percent, based on fresh feed. The combination of hydrogenation and reaction in the transfer-line reactor, in accordance with this invention, cracks the recycle gasoline fraction to such an extent that only about 10 to 25 volume percent, based on fresh feed, is recycled.

With the process of this invention, advantage is taken of the presence of hot, regenerated catalyst particles within the transfer line closer to the regenerator, the higher-than-reactor temperature prevailing in the catalyst transfer-line, and the longer overall time of contact with the catalyst (including the time within the transfer-line zone, the time the recycled fraction is in contact with incoming oil and hot regenerated catalyst in the line leading to the reactor, and the time within the main catalyst zone). Of prime importance is the transfer-line temperature, which is in excess of reactor temperature. This process is applicable to moving-bed, fluid-type catalytic cracking processes, or other circulating catalysts processes, including Thermofor catalytic cracking.

The principal object of this invention is to provide a process for the conversion of hydrocarbons in the presence

of a suspended catalyst, wherein the hydrocarbons are caused to pass through a catalytic cracking zone, the cracked products are fractionated to yield (*a*) tail gas, (*b*) a gasoline fraction of enhanced octane number, (*c*) a heavy gasoline fraction boiling in the range of about 350–425° F., (*d*) a light gas-oil boiling above about 400° F., (*e*) a heavy cycle stock or gas oil boiling above about 460° F., (*f*) a decanted oil, and, (*g*) a slurry oil. The heavy gasoline fraction is separately hydrogenated, the hydrogenated product is preheated, if necessary, and sent to a transfer-line reactor conveying hot, regenerated catalyst to said reaction zone, and the cracked and hydrogenated products are mixed with incoming feed at the end of the transfer-line reactor and pass to the main reactor.

A further object of this invention is to provide a process wherein the above-mentioned heavy cycle stock or gas-oil fraction (*e*) is also hydrogenated under separate conditions, and then is recycled with said hydrogenated heavy gasoline fraction. It is to be understood that the heavy gasoline fraction (*c*) and the heavy gas-oil fraction (*e*) which are hydrogenated and recycled may be derived either from a single feed oil or from the cracked products of different feed oils, and may be used to supplement the gasoline yield and octane number value obtained from feed oil from a second source. Likewise, the heavy gasoline ends, resulting after separation of the desired product having an end point of about 350° F., and obtained from the combined cracked products of the afore-mentioned utilization of two-source feeds, may also be hydrogenated and recycled in accordance with this invention. In addition, heavy straight-run naphtha from a different source may be recycled, with or without prior hydrogenation.

In general, the present invention comprises not only an improved method of operating a catalytic cracking process wherein it is desired to produce a gasoline having an end-point of about 350° F., and wherein economic loss and the necessity for changes in equipment are minimized, but it comprises a method for utilizing the more thermally stable, heavier fractions and straight-run fractions to advantage in increasing yields per pass, and in increasing the octane number of the lower end-point product. This invention can be applied to any catalytic cracking process having particular application to continuous processes and will be described, although without being limited thereby, in connection with a fluid-type operation using cross-flow between a fluid reactor zone and a fluid catalyst regeneration zone. In these operations the reactor and regenerator conditions are selected in accordance with the feed characteristics, the degree of cracking desired per pass, and the desired character of the product. For purposes of this invention, reactor temperatures may range from above 750° F. to as high as 950° or 1000° F., using atmospheric to superatmospheric pressures, and space velocities from 0.5 to 10 parts of oil per hour per part of catalyst.

The feed oil may comprise any hydrocarbon material capable of producing gasoline-boiling-range hydrocarbons on being subjected to cracking conditions. The feed oils may be obtained from various crudes, comprise those hydrocarbons having boiling ranges between about 400° to 1000° F., and include such material as virgin gas-oils, heavy cycle stocks, and mixtures thereof, which, among other materials, are capable of producing heavy cracked gasoline hydrocarbons boiling in the range of 350° to 425° F. for recycle in accordance with the invention. Preferred feeds are distillates from Gulf Coast and Mid-Continent crudes which do not have excessive carbon contents, that is, the Conradson carbon residue should be below about 1.0 weight percent.

The conditions used in regenerating the catalyst are designed to oxidize carbonaceous materials and other combustible contaminants from the surfaces of the catalyst particles and revivify the catalyst to its original activity. For this purpose, temperatures in the regeneration zone are generally higher than those in the reactor, being in the range of 1000° to 1250° F., with pressures comparable to those in the reactor, namely, atmospheric to superatmospheric. Any oxidizing medium may be used, including air and other mixtures of oxygen and inert gases. The only limitation attaching to the regeneration conditions is that they must be more severe than the reactor conditions so that advantage can be taken of the greater heat content of the regenerated catalyst.

Any suitable cracking catalyst may be used, such as bauxite, alumina, acid-treated kaolin, silica, fuller's earth, acid-treated bentonite, diatomaceous earth, synthetic silica-alumina, and natural clays and the like which are adapted to fluidization. These catalytic materials may be used singly or in admixture, and may include various known promoters.

In order to better explain the invention, reference is made to the drawing which is a schematic flow diagram showing one system applicable to the process. Only the essential reactors, vessels and conduits are shown in order that one skilled in the art can follow the process and understand the invention more readily. Various pumps, gauges, pressure-control valves, temperature- and flow-control devices, etc., that are required for actual operation have been omitted .

Hydrocarbon feed, which may be preheated sufficiently to vaporize a substantial portion thereof, is introduced into line 1, through valve 2, wherein it mixes with hot, regenerated catalyst.

The mixture of feed and catalyst passes to reactor 3 maintained under cracking conditions. A fluidized reaction zone 4 is maintained with upper level 5 within reactor 3 by means of suitable space velocities and pressure controls well-known in the art. After sufficient residence time within zone 4, the reaction products, some catalyst, and any unreacted feed are withdrawn through line 6 into separator 7 wherein the entrained catalyst particles are separated by cyclonic action for return to the reaction zone through line 8. Reaction products and any remaining catalyst fines pass through line 9 and enter separator 10 which removes most of the remaining portion of catalyst fines and returns them to the reactor by line 11. Reaction products substantially freed from catalyst fines pass through line 12 and valve 13 into fractionator 14 near the bottom. Fractionator 14 may be any type of fractionator suitable for fractional distillation, such as that shown fitted with plates 15, 16, 17 and 18, for the separation of the various fractions in accordance with the characteristics of the reaction products.

A portion of the catalyst is separated from reaction zone 4 and passed through line 19 and valve 20 into stripper 21. Steam is introduced into stripper 21 at line 22 controlled by valve 23. Any entrained reaction products in the catalyst are conveyed by line 24 back to the reactor. Aerating steam may also be introduced in stand-pipe 25, through line 26 and valve 27. Valve 28 in stand-pipe 25 controls the flow of used catalyst through line 29 into regenerator 30. Regenerator 30 may be an ordinary fluid-type regenerator wherein air is introduced at line 31, controlled by valve 32. The mixture of catalyst and air maintains a fluid bed 33, having an upper level 34 in regenerator 30, which is controlled by the rates of flow of material input and output in the regeneration zone. The rate of removal of regenerated catalyst is controlled by valve 35 in stand-pipe 36. The mixture of combustion gases and catalyst within regenerator 30 passes via line 37 into separator 38 wherein the combustion gases from the regeneration are removed and conveyed from the system by line 39 and any entrained catalyst is returned to the regenerator by line 40. Hot regenerated catalyst leaving stand-pipe 36, controlled by valve 35, is continuously passed into injector 41 and is forced into transfer-line reactor 42.

Returning to fractionator 14, overhead materials comprising gas and gasoline fractions boiling up to about 350° F., are removed by means of line 45 and pass through

condenser 446 into gas-liquid separator 47. Tail-gas is removed at line 48 and any water present (as a result of superheated steam used to conduct the distillation) is collected in the bottom of separator 47 for withdrawal through line 50. Liquid gasoline product is withdrawn through 51 by pump 52. This product may be sent to a stabilizer (not shown) via line 53 and a part thereof may be returned by line 54 to the top of fractionator 14 as reflux. Plate 15 serves to remove a heavy gasoline fraction, in liquid form, boiling between about 350–425° F., which passes through line 55 to stripper 56. Steam is introduced into stripper 56 by line 57, and any lighter hydrocarbons present are removed and conducted by line 58 back to fractionator 14. The stripped, liquid, heavy gasoline fraction is removed at line 59. The boiling range of the heavy gasoline fraction must be within about 350–425° F. in order to obtain the increased yields of the present invention and, accordingly, the boiling range of the original fraction withdrawn at line 59 should be checked, and if found to contain components boiling materially outside this range, it should be refractionated in a separate fractionator (not shown). The initial boiling point of this fraction may be as high as 375° F. but the end boiling point should not be over about 425° F.

Plate 16 serves to remove a light cycle stock boiling above about 400° F. This is sent through line 60 into stripper 61 operating by means of steam introduced in line 62. Light ends from this fraction are sent back to fractionator 14 by line 63 and the stripped light recycle stock is withdrawn at line 64. The end point of the recycle stock is about 600° F.

Heavy recycle gas-oil, boiling above 460° F., is removed at plate 17 by means of line 65, and sent through heat-adjusting coil 66 and valve 67 to hydrogenation reactor 68. Ordinarily, and for purposes of comparison of the results obtainable by this invention, heavy recycle gas-oil alone is recycled via line 69 and pump 70 directly to join fresh feed entering in line 1.

A decanted oil stream is withdrawn from plate separator 18 through line 71, and a slurry oil is withdrawn from the bottom of the fractionator via line 72, controlled by valve 73.

The heavy gasoline fraction withdrawn through line 59 passes through heat-adjusting coil 74 and to valve 75 to hydrogenation reactor 76. Hydrogenated product leaves reactor 68 via line 78 and reactor 76 via line 80 and the combined stream passes via line 81 to gas-liquid separator 82. Liquid from separator 82 passes into line 84, through preheater coil 85, by means of pump 86 and line 87 into injector 41 and reactor transfer line 42. Reactors 68 and 76 are arranged so that they may be operated as two separate zones or as one common zone. The operation as two separate zones allows the application of different hydrogenation conditions to the recycled fractions being treated. Operation of these reactors as a joint reactor allows the mixture of heavy gasoline fraction (line 59) and heavy recycle gas oil (line 65) to be treated to common hydrogenation conditions. By-pass line 88, controlled by valves 89, is provided to by-pass reactor 76 and convey all or part of the heavy gasoline fraction to line 84. Vessel 90 is used to supply a heavy straight-run naphtha to the process by means of line 91, controlled by valve 92 in accordance with one embodiment of the invention. Gas-liquid separator 82 is provided to separate hydrogen from the liquid hydrogenated products from reactors 68 and 76. The separated hydrogen gas leaves separator 82 by line 93, and is returned to reactors 68 or 76 by compressor 94 and line 95 and branch lines 96 and 97 controlled by valves 98 and 99. Make-up hydrogen is introduced through line 100.

Conditions of hydrogenation in reactors 68 and 76 are maintained sufficient to reduce the olefinic content and partially saturate the aromatic content of the respective feeds being treated therein. For this purpose, temperatures of at least 600° F. at 400 p.s.i. with a minimum space velocity of about 0.5 and a hydrogen to hydrocarbon ratio of about 1–2 has been found to be necessary. Table I sets forth the range and specific conditions for reactors 68 and 76 found applicable for purpose of obtaining substantially increased yields of high octane gasoline in accordance with this invention.

Table I

Reactor Feed Condition	Reactor 76, Heavy Gasoline Fraction		Reactor 68, Heavy Recycle Gas Oil	
	Range	Specific	Range	Specific
Temperature, ° F.	600–750	700	600–800	725
Pressure, p.s.i.	400–850	650	400–1,000	700
Sp. Vel., Liq. Vol./Vol./Hr.	0.5–4	1.5	0.5–4	1.0
Hydrogen/Hydrocarbon mole ratio	1–10	5	2–10	6

Catalyst Composition (for both)	Range	Specific
as CoO, wt. percent	2–4	3
as MoO_3, wt. percent	8–15	12
Alumina (containing 1–5% silica)	balance	

The hydrogenated recycle fractions along with any untreated fractions from lines 88 or 91 enter injector 41 at a temperature of about 450° to 800° F. This temperature is adjusted by coil 85. The regenerated catalyst is at a substantially higher temperature than the recycled fractions, making it possible to employ a minimum amount of heating in coil 85. The ratio of regenerated catalyst to recycled fractions in reactor line 42 is high so that substantially complete vaporization of the hydrocarbons occurs with very little reduction in overall temperature.

The conditions within line 42 are maintained at regeneration temperatures, or slightly lower, and the catalyst-hydrocarbon mixture is sent therethrough at a linear velocity of about 6 to 10 feet per second, with residence times varying from about 0.5 to 5.0 seconds. Fluidized conditions are maintained by sloping line 42 downwardly about 1 foot per every 10 to 20 ft. in length. This allows a residence time sufficient to pre-crack the hydrogenated recycled materials.

At the end of line 42, the reaction mixtures meets the incoming feed in preheated condition in line 1, and the combined mixture passes into reaction zone 4.

By proceeding in the manner described, the instant process exhibits certain advantages over prior art methods. The hydrogenation of the olefinic content of the selected recycle stocks minimizes the tendency of the olefin content of these stocks to crack and thus decreases the product content of olefins and more unsaturated products, which lead to polymers, gums and coke in the main reactor. Thus, the recycle stocks are slightly deactivated as far as their tendency to form coke is concerned, and activated as far as their tendency to form gasoline hydrocarbons of high octane value is concerned. In addition, the olefin content of the finished gasoline is reduced without a corresponding reduction in octane number.

An additional feature of the process is that the hydrogenated recycle stocks are subjected to higher-than-normal reaction temperatures and catalyst-to-oil ratios, thereby achieving greater cracking of the more refractory components. The net results are more efficient use of heat, less catalyst fouling, higher catalyst activity in zone 4, and a superior product distribution in comparison to that obtained if the transfer-line reactor were not used, or the hydrogenated recycle stocks were sent directly to zone 4. The following table of results illustrates the various benefits from my process.

Table II

Example	I	II	III	IV	V	VI	VII
Feeds, in volumes (liquid):							
Fresh Feed—Virgin Gas Oil	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Fresh Feed—Heavy Straight-Run Naphtha					10.7	10.7	10.7
Recycle Heavy Gas Oil	40.0	40.0	40.0		40.0	40.0	
Recycle Heavy Gas Oil—Hydrogenated				37.0			37.0
Recycle Heavy Gasoline			8.1			20.6	
Recycle Heavy Gasoline—Hydrogenated		15.2	8.1	15.8	27.9	20.6	28.5
Total Feed	140.0	155.2	156.2	152.8	178.6	181.2	176.2
Feed to Transfer-Line Reactor		15.2	16.2	52.8	38.6	41.2	76.2
Products:							
Gas and Coke (Wt. Percent Fresh Virgin Gas Oil)	19.3	20.8	21.0	18.8	22.7	23.3	20.7
Butanes and Butylenes, Vol. Percent of Fresh Virgin Gas Oil	16.7	18.0	18.0	20.8	19.9	19.9	22.7
Gasoline (350° F., E.P.) Vol. Percent Fresh Virgin Gas Oil	[1] 40.8	43.8	43.6	45.5	47.8	47.4	49.5
Gasoline (350-425° F.) Vol. Percent Fresh Virgin Gas Oil	[1] 7.6						
Light Cycle Stock, Vol. Percent Fresh Virgin Gas Oil	[1] 22.0	24.3	24.4	21.7	27.6	27.7	25.0
Heavy Oils, Vol. Percent Fresh Virgin Gas Oil	5.0	5.0	5.0	5.0	5.0	5.0	5.0

[1] As normally fractionated, the products are:
Gasoline (400° F., E.P.) ..... 45.9 Vol. Percent.
Light Cycle Stock ..... 24.5 Vol. Percent.

For comparison, Example I shows results with typical normal operation, without any of the embodiments of this invention.

Example II shows results with a heavy gasoline fraction which is hydrogenated and sent to catalyst transfer-line.

Example III shows the results from splitting the heavy gasoline fraction into two parts and hydrogenating only one part thereof, but recycling both parts.

Example IV, shows the results with a hydrogenated heavy gas-oil recycle stocks and a hydrogenated heavy gasoline fraction.

Example V shows the results from the embodiment of Example II with the additional feature of adding a heavy straight-run naphtha to the hydrogenated heavy gasoline fraction.

Example VI shows the results from the embodiment of Example III with the additional feature of adding a heavy straight-run naphtha to the recycled parts.

Example VII shows the results from the embodiment of Example IV with the additional feature of adding a heavy straight-run naphtha to the combined hydrogenated stocks.

The examples demonstrate the increased gasoline yields that are made possible through the process of this invention. Also, the volume percent light cycle stock is increased. The temperature of the contact of the recycled fractions or recycled mixtures is maintained at about 100° to 300° F. higher than the temperature of the incoming mixture and about 100° to 300° F. higher than that prevailing in the main conversion zone 4 of reactor 3. This temperature differential is subject to variation outside the 100 to 300° F., range for those feed stocks containing more or less refractory hydrocarbons. In general, the conditions set forth herein will apply to those feeds from which economical yields of gasoline can be obtained. The only limitations applying to the invention appear in the appended claims.

The embodiments of the invention in which an exclusive property or privilege is claimed are defined as follows:

1. In a catalytic cracking process wherein a hydrocarbon oil having a boiling range of about 400° to 1000° F. is contacted with a cracking catalyst at a temperature of above about 750° F. to about 1000° F. in a conversion zone, used catalyst and reaction products including a gasoline fraction of enhanced octane number and an end boiling point of about 350° F. are separated, the improvement whereby gas and coke formation are reduced and the yield of said gasoline fraction is increased which comprises separating a heavy gasoline fraction boiling in the range of about 350° to 425° F. from said reaction products, separately hydrogenating said heavy gasoline fraction at a temperature of about 600° to 750° F. at a pressure of about 400 to 850 p.s.i. in the presence of a catalyst and about 1 to 10 mols of hydrogen per mol of said heavy gasoline fraction at a minimum space velocity of about 0.5 to substantially saturate the olefin hydrocarbon content thereof, separately regenerating said used catalyst at a temperature of above said reaction temperature to produce regenerated catalyst at a temperature of about 1000° to 1250° F., contacting said substantially saturated heavy gasoline fraction with said hot regenerated catalyst for about 0.5 to 5 seconds in a separate reaction zone to induce cracking of the refractory constituents therein, mixing incoming feed hydrocarbon oil with said cracked heavy gasoline fraction and regenerated catalyst and conducting the mixture to said conversion zone.

2. The process in accordance with claim 1 in which said reaction products from said conversion zone are separated into a heavy recycle gas oil boiling above about 460° F., about 40 volumes of said heavy recycle gas oil per 100 volumes of feed hydrocarbon oil are mixed with said incoming feed hydrocarbon oil and about 15.2 volumes of said substantially saturated heavy gasoline fraction per 100 volumes of said feed hydrocarbon oil are mixed with said hot regenerated catalyst in said separate reaction zone.

3. The process in accordance with claim 2 in which about 8.1 volumes of said substantially saturated heavy gasoline fraction and about 8.1 volumes of untreated heavy gasoline fraction per 100 volumes of feed hydrocarbon oil are mixed with said hot regenerated catalyst in said separate reaction zone.

4. The process in accordance with claim 2 in which about 27.9 volumes of said substantially saturated heavy gasoline fraction and about 10.7 volumes of a heavy straight-run naphtha per 100 volumes of feed hydrocarbon oil are mixed with said hot regenerated catalyst in said separate reaction zone.

5. The process in accordance with claim 4 in which about 20.6 volumes of said substantially saturated heavy gasoline fraction and about 20.6 volumes of untreated heavy gasoline fraction are mixed with said heavy straight-run naphtha passing into contact with said hot regenerated catalyst in said separate reaction zone.

6. The process in accordance with claim 1 in which said reaction products from said conversion zone are separated into a heavy recycle gas oil boiling above about 460° F., said heavy recycle gas-oil fraction is hydrogenated at a temperature of about 600 to 800° F. at a pres-

sure of about 400 to 1000 p.s.i. in the presence of a catalyst and about 2 to 10 mols of hydrogen per mol of said heavy recycle gas-oil fraction to substantially saturate the olefin hydrogen content thereof and about 37 volumes of the product along with about 15.8 volumes of said substantially saturated heavy gasoline fraction per 100 volumes of feed hydrocarbon oil are mixed with hot regenerated catalyst in said separate reaction zone.

7. The process in accordance with claim 1 in which said reaction products from said conversion zone are separated into a heavy recycle gas oil boiling above about 460° F. and about 37 volumes of said heavy recycle gas oil and about 15.8 volumes of said substantially saturated

heavy gasoline fraction per 100 volumes of said feed hydrocarbon oil are mixed with said hot regenerated catalyst in said separate reaction zone.

8. The process in accordance with claim 7 in which said heavy recycle gas-oil fraction and said heavy gasoline fraction are combined and hydrogenated in a common hydrogenation zone.

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