Low octane naphtha and fresh gas oil are catalytically cracked with a zeolite catalytic cracking catalyst in separate elongated reaction zones yielding a naphtha having an increased octane rating. Recovering a naphtha fraction, particularly a heavy naphtha fraction, from the cracked product, and combining it with the fresh naphtha feed, further enhances the octane rating of the naphtha product. A cycle gas oil may be recovered from the reactor effluent and combined with the fresh naphtha feed or introduced into a separate (a third) elongated reaction zone to effect maximum recovery of naphtha and lighter stocks having improved product quality. Optionally, any of the feedstocks may be subjected to further cracking in a dense bed of the zeolite catalyst.

29 Claims, 2 Drawing Figures
CATALYTIC CRACKING OF NAPHTHA AND GAS OIL

CROSS-REFERENCE TO RELATED APPLICATIONS

This is related to application Ser. No. 77,480 which discloses the fluid catalytic cracking, in risers, of low octane naphtha and a recycle stream of full range naphtha or heavy naphtha and application Ser. No. 889,714 which discloses the catalytic cracking of naphtha with crystalline aluminosilicate catalyst.

BACKGROUND OF THE INVENTION

This invention relates to the catalytic cracking of hydrocarbons. In particular this invention relates to the fluid catalytic cracking of fresh gas oil and naphtha having a low octane rating and boiling in the range of 100°F-450°F.

Gasoline is frequently blended from stocks, including naphtha, the octane of which has been increased through catalytic reforming. Both virgin and cracked stocks may be upgraded by reforming operations. Catalytic reformers are usually operated to provide the desired octane improvement with the least conversion of gasoline to saturated butanes and lighter materials.

The gasoline blending pool is maintained by a variety of operations — isobutanes and butenes, for example, are charged to alkylation units and light olefins are polymerized to provide high octane blending components while the catalytic cracking of gas oil augments the supply of naphtha as well as providing additional feed for alkylation and polymerization units. Although hydrocracking provides additional quantities of gasoline blending naphthas, the heavy naphtha from hydrocracking often has a relatively low octane number.

Recently the introduction of zeolite cracking catalysts has effected significant improvements in the catalytic cracking operation. When employed for gas oil cracking in existing catalytic cracking units these highly active catalysts have produced increased throughput and improved product quality. In addition, catalytic cracking apparatus, such as disclosed in U.S. Pat. Nos. 3,433,733 and 3,448,037, have been developed specifically for use with these improved catalysts. This apparatus incorporates the concept of "riser cracking" wherein the feedstocks are cracked in elongated reaction zones or risers terminating in a tapered reactor containing a dense phase and dilute phase of catalyst. A mixture of zeolite catalyst and gas oil passes through the riser under cracking conditions which are tailored to the particular feedstock and desired products. After passing through the riser further cracking of the feed can be achieved, if necessary, in the fluidized dense phase of catalyst in the reactor.

Although the zeolite catalysts increase the supply of high quality naphtha, yields of the lighter hydrocarbons are substantially lower than from catalytic cracking with amorphous silica-alumina catalysts. In the future, therefore, the supply of isobutane, propylene and butene for alkylation production and of these and other light hydrocarbons for polymerization and petrochemical manufacture will continually decline. A process which will upgrade naphtha streams for use in gasoline blending and supply additional quantities of C4 and lighter hydrocarbons is highly desirable.

Naphtha is more difficult to crack than gas oil and up to the present time limited success has been obtained in cracking naphtha catalytically. Traditional cracking catalyst, such as silica-alumina, exhibited relatively poor selectivity and activity when employed to crack naphtha resulting in the formation of relatively large amounts of gas and coke and producing small amounts of desirable olefins and aromatics. U.S. Pat. No. 3,284,341 discloses a process for the catalytic cracking of naphtha with a silica-alumina catalyst to produce substantial quantities of olefins and aromatics by maintaining the space velocity above about 4.5, the pressure between 0 and 20 psig and the reaction temperature between 1,000°F and 1,200°F.

The new zeolite cracking catalysts are being employed extensively in gas oil cracking operations but their utility for the conversion of naphtha has yet to be fully explored. U.S. Pat. No. 3,247,098 discloses that hydrogen mordenite, a crystalline aluminosilicate, is an extremely active catalyst for the conversion of light naphtha to lighter components together with improving the octane number of the resultant naphtha. The utility of the mordenite alumino-silicate as a cracking catalyst for naphtha was found to be surprising in view of the ineffectiveness of a magnesium faujasite catalyst to satisfactorily crack naphtha. Magnesium faujasite was known to be a highly effective gas oil cracking catalyst. The development of processes which may usefully be employed for the cracking of naphtha and gas oil and which will be selective for the production of light hydrocarbons as well as naphthas having enhanced octane ratings is highly desirable.

SUMMARY OF THE INVENTION

Broadly, our invention is directed to the catalytic cracking of gas oils boiling in the range of about 430°F-1,050°F. and naphthas boiling in the range of about 100°F-450°F. to produce substantial quantities of naphtha having an octane rating significantly higher than that of the low octane naphtha feed and yielding substantial quantities of lighter hydrocarbons which may serve as feed for alkylation and petrochemical manufacture. A fluid catalytic cracking unit having multiple risers and employing zeolite cracking catalysts has been found to offer particular utility for the catalytic cracking of such feed streams. The naphtha and the gas oil are catalytically cracked in separate cracking zones under conditions specifically tailored to the particular feedstock. In addition, a full range or heavy naphtha cut may be removed from the reactor effluent for recycling to the fresh naphtha cracking zone and a cycle gas oil may also be removed from the reactor effluent for introduction into the fresh naphtha cracking zone or into a separate (third) cracking zone to effect maximum recovery of naphtha and lighter stocks having improved product quality. A cracking zone may be limited to an elongated reaction zone, also known as a riser, or a combination of an elongated reaction zone and the dense phase of catalyst in the reactor vessel.

BRIEF DESCRIPTION OF THE DRAWINGS

The present invention will be more readily understood by reference to FIG. 1 which depicts a flow diagram of a preferred embodiment of the process of the invention and to FIG. 2 which depicts an apparatus for carrying out a preferred embodiment of the invention.
DESCRIPTION OF THE PREFERRED EMBODIMENTS

Broadly, we have found that the octane rating of a low octane naphtha can be significantly improved by catalytically cracking it in a fluid catalytic cracking unit employing a multiplicity of elongated reaction zones, hereinafter called risers or riser conduits, wherein the low octane naphtha is introduced into one of the risers and fresh gas oil is introduced into a second riser. Operating conditions within the several risers are established to obtain optimum product yield and product quality from each of the individual streams. Broadly, our process contemplates a process for the catalytic cracking of naphtha and gas oil with a zeolite cracking catalyst in a fluid catalytic cracking unit comprising a reactor, a regenerator and a multiplicity of elongated reaction zones wherein said reactor contains a dense phase and a dilute phase of said catalyst and said elongated reaction zones terminate at said reactor which comprises:

a. passing a naphtha stream and a zeolite cracking catalyst through a first elongated reaction zone under naphtha cracking conditions,

b. passing a gas oil and a zeolite cracking catalyst through a second elongated reaction zone under gas oil cracking conditions,

c. discharging the effluents from said first and second reaction zones into a catalyst phase in said reactor, said effluents comprising vaporous reaction mixture and catalyst, and

d. recovering from the vaporous reaction mixture in a dilute phase of catalyst in said reactor a fraction boiling in the range of 100° to 450°F, having an octane rating higher than said naphtha stream and a fraction boiling below 100°F.

The catalyst phase into which the effluent from a particular riser discharge is either the dense phase or the dilute phase of catalyst in the reactor which leads to a number of optional embodiments incorporating various combinations of riser cracking and dense phase cracking, i.e., bed cracking.

In the simplest embodiment, the cracking of both naphtha and gas oil is restricted to the risers by discharging the effluent from both risers into the dilute phase of catalyst in the reactor vessel. In this situation the reactor vessel is utilized as a disengaging zone with little or no cracking taking place therein.

In another embodiment, the gas oil is subjected to further cracking in the dense catalyst phase. This is achieved by discharging the effluent from the naphtha riser into the dilute phase of catalyst and the effluent from the gas oil riser into the dense phase of catalyst, the vaporous reaction mixture from the gas oil riser passes through the dense phase of catalyst under catalytic cracking conditions effecting an additional conversion of 5 to 30 volume percent and discharges into the dilute phase of catalyst.

In a further embodiment, the gas oil is subjected only to riser cracking while the naphtha is cracked in both the risers through the dense phase of catalyst. The effluent from the gas oil riser is discharged directly into the dilute phase of catalyst in the reactor vessel while the effluent from the naphtha riser is discharged into the dense phase of catalyst, the vaporous reaction mixture from the naphtha riser passes through this dense phase under catalytic cracking conditions effecting an additional conversion of 5 to 30 volume percent and discharges into the dilute phase of catalyst.

In another embodiment, the gas oil and the naphtha are both subjected to riser cracking and bed cracking by discharging the effluent from both risers into the dense phase of catalyst. The vaporous reaction mixture passes through the dense phase under catalytic cracking conditions effecting an additional conversion of each stream of 5 to 30 volume percent and discharges into the dilute phase.

In another series of embodiments a cycle gas oil is recovered from the effluent from the catalytic cracking unit and recycled to the unit wherein it is subjected to riser cracking in a third elongated reaction zone and bed cracking in the dense phase of catalyst in the reactor. This embodiment envisions the following additional steps which may be optionally added to any of the above four embodiments:

e. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling above about 500°F,

f. passing said fraction of step (e) and a zeolite cracking catalyst through a third elongated reaction zone under cycle gas oil cracking conditions, and

g. discharging the effluent from said third elongated reaction zone into a dense phase of catalyst in said reactor, said effluent comprising vaporous reaction mixture and catalyst, said vaporous reaction mixture from said third elongated reaction zone passing through said dense phase under catalytic cracking conditions effecting an additional conversion of 5 to 30 volume percent and discharging into a dilute phase of catalyst.

In still another series of embodiments cycle gas oil may be recovered from the effluent from the catalytic cracking unit and recycled to the naphtha riser where it is subjected, together with the naphtha, to both riser and bed cracking. This will permit two more optional embodiments wherein the fresh gas oil is subjected to riser cracking only or a combination of riser and bed cracking.

Another group of optional embodiments envisions the use of a recycle naphtha stream. A naphtha cut which may be either a full range naphtha, i.e., about a 100° to 450°F boiling range, or a heavy naphtha, i.e., about a 250° to 450°F boiling range, may be removed from the reactor effluent and recycled for introduction into the naphtha riser wherein it is combined with the fresh naphtha and subjected to both riser and bed cracking. Fresh gas oil is introduced into a separate riser. In this two-riser concept the fresh naphtha and the recycle naphtha are subjected to both riser and bed cracking. With the fresh gas oil subjected either to riser cracking only or riser and bed cracking, two more additional embodiments are possible. The following additional steps relating to the naphtha recycle are added:

j. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling between 100° and 450°F, and

k. introducing the fraction from step (j) into the first elongated reaction zone effecting conversion of said fraction.

Two more optional embodiments envision two recycle streams, cycle gas oil and recycle naphtha, and the use of three risers. Both recycle streams are recovered from the reactor effluent as explained above in describing other embodiments. In this three-riser concept the
recycle naphtha is combined with fresh naphtha and the combined streams are passed through the naphtha riser and subjected to both riser and bed cracking. The cycle gas oil is introduced to its own riser and is cracked in both the riser and the bed of catalyst in the reactor. The fresh gas oil passes through a third riser and is subjected to riser cracking only in one embodiment or both riser and bed cracking in the other.

In accordance with this invention the fresh feedstocks for this process comprise naphtha and gas oil. The naphtha feedstocks comprise hydrocarbons boiling in the range of about 100°-450°F. and include many refinery streams having low economic value which may be upgraded through the process of our invention. Useful naphtha feedstocks are usually highly paraffinic and include such light hydrocarbon fractions as low octane naphtha, aromatic solvent extraction raffinate, thermal cracked naphtha, coker naphtha, low octane naphtha from hydrocracking operations and straight run naphthas. As used herein, the term "low octane fresh naphtha" refers to these useful feedstocks. Because of the refractory quality of these naphtha feed streams, they must be subjected to more severe conditions than is the fresh gas oil feed if significant conversion is to be obtained.

The gas oil feed in our process comprises petroleum stocks boiling from about 430° to 1,050°F. and comprises heavy atmospheric gas oil, light and heavy vacuum gas oils, gas oil from coking operations, vis-broken gas oil, deasphalted gas oil, decarbonized gas oil, hydrocracked gas oil, hydrocracked gas oil and solvent extracted gas oil. All of these feedstocks are petroleum stocks whose processing has not previously included fluid catalytic cracking and are commonly referred to, therefore, as virgin gas oils.

Products from the process of our invention include naphthas with improved octane ratings and hydrocarbons boiling below the initial boiling point of the naphtha feed which will find particular utility as feed streams for petrochemical and gasoline manufacture.

The catalyst employed in the instant invention is a cracking catalyst of the zeolite type as exemplified by those catalysts wherein a crystalline aluminosilicate is dispersed in a siliceous matrix. Among the preferred zeolites which may be usefully employed in the catalyst used in the process of our invention are those known as zeolite X and zeolite Y, including both the naturally occurring and synthetic varieties. Because of their extremely high activity, these zeolite materials are composed with a material processing a substantially lower level of catalytic activity, a siliceous matrix which may be of the synthetic, semi-synthetic or natural type. The materials may include silica-alumina, silica-gel, silicalydrilla, silica-magnesia, silica-thoria or silica-zirconia which have been successfully employed heretofore. In general, the composite crystalline zeolite catalyst comprises about 1 to 50 weight percent zeolite, about 5 to 50 weight percent alumina and the remainder silica. The crystalline aluminosilicate portion of the catalyst composition is a natural or synthetic, alkali metal, crystalline aluminosilicate which has been treated to replace all or at least a substantial portion of the original alkali metal ions with other ions such as hydrogen and/or a metal or combination of metals such as barium, calcium, magnesium, manganese or rare earth metals, for example, cerium, lanthanum, neodymium, praseodymium, samarium and yttrium. The crystalline zeolites contemplated above may be represented by the formula

\[ M_{2n}O \cdot Al_2O_3 \cdot x SiO_2 \cdot y H_2O \]

where \( M \) represents hydrogen or a metal, \( n \) its valence, \( x \) has a value ranging from 2 to 10 and \( y \) ranges from 0 to 10, in dehydrated zeolites \( y \) will be substantially 0. In the instant invention the preferred crystalline zeolites are either natural or synthetic zeolite X or zeolite Y. In highly preferred embodiments \( m \) is selected from the group consisting of hydrogen, calcium, magnesium and the rare earth metals.

Those skilled in the art will readily appreciate that the catalytic cracking equipment currently employed in the petroleum industry which incorporates the concept of "riser cracking" may be utilized to practice some embodiments of our invention. A particularly preferred apparatus is that described in U.S. Pat. No. 3,433,733 wherein the charge stocks are cracked in two elongated reaction zones or "risers" terminating in a tapered reactor chamber wherein further dense bed cracking may take place. One riser passes through the tapered wall of the reactor terminating in a downwardly directed outlet while the second riser passes through the bottom of the reactor discharging upwardly into the reactor. The cracking apparatus incorporates a stripping section beneath the dense fluid bed wherein entrained and adsorbed hydrocarbons are displaced from the catalyst by means of steam as the catalyst passes from the reactor to the regeneration vessel. As with conventional fluid catalytic cracking equipment, the catalyst is contacted with an oxygen-containing gas in the regenerator to effect combustion of at least a portion of the deposited coke. The regenerated catalyst is then reintroduced into the bottom of the risers at a point where the feedstocks are introduced. By introducing fresh gas oil to one riser and cycle gas oil to the other, it is possible to operate the individual risers under conditions specifically tailored to obtain the desired cracking of the particular feedstock passing therethrough. The overall effect, of course, is to achieve optimum performance of both the fluid catalytic cracking unit and the zeolitic cracking catalyst being employed.

Those embodiments of this invention requiring two risers may be conducted, for example, in the apparatus described in U.S. Pat. No. 3,433,733 by D. P. Bunn, Jr. et al. and discussed above. When the embodiment necessitates the use of three risers, an apparatus such as depicted in FIG. 2 may be employed which is an improvement over the apparatus of Bunn et al. Broadly, this improved embodiment may be described as follows: In an apparatus for fluid catalytic cracking of a type comprising:

- a reactor chamber having a tapered portion of greater diameter at the upper portion than at the lower portion,
- a stripping chamber,
- a regenerator chamber,
- a multiplicity of riser conduits entering and passing through the walls of said reactor chamber and discharging into said reactor chamber,
- means to introduce steam into the lower portion of the reactor chamber,
- means to withdraw vapors from the stripping chamber,
- means to withdraw gaseous products and steam from the upper portion of the reactor chamber,

3,784,463
h. means to withdraw solids from the lower portion of the reactor chamber and to discharge the same into the stripping chamber,

i. means to introduce steam into the lower portion of the stripping chamber,

j. means to withdraw solids from the lower portion of the stripping chamber and to discharge the same into the regenerator chamber,

k. means to introduce combustion gas into the lower portion of the regenerator chamber,

l. means to withdraw flue gas from the upper portion of the regenerator chamber,

m. means to withdraw regenerated catalyst from the regenerator chamber and to discharge the same into the inlet of the riser conduits, and

n. means to introduce oil feed into the inlet of said riser conduits, the improvement which comprises:

1. a first riser conduit entering and passing through a wall of the reactor chamber and discharging into the reactor chamber,

2. a second riser conduit entering and passing through a wall of the reactor chamber and discharging into the reactor chamber at a point above the discharge point of the first riser, and

3. a third riser conduit entering and passing through a wall of the reactor chamber and discharging into the reactor chamber at a point intermediate the discharge points of the first and second riser conduits.

In preferred embodiments, the first riser conduit may pass through the bottom wall of the reactor chamber, the second riser conduit through the side wall of the reactor chamber or the third riser conduit through the side wall of the reactor chamber. In other preferred embodiments, the first riser conduit may terminate in the reactor chamber in an upwardly directed outlet, the second riser conduit in a downwardly directed outlet or the third riser conduit in a downwardly directed outlet.

In the naphtha riser, the operating conditions contemplated herein include a temperature of 750°-1,300°F., preferably 900°-1,000°F., a conversion per pass of 25-80 volume percent, preferably 30-60 volume percent, and a vapor velocity of 15-50 feet/second, preferably 20-40 feet/second. In the fresh gas oil riser, the operating conditions include a temperature of 840°-1,100°F., preferably 890-1,000°F., a conversion per pass of 30-80 volume percent, preferably 40-65 volume percent, and a vapor velocity of 15-50 feet/second, preferably 20-40 feet/second. When cycle gas oil is introduced into a third riser the operating conditions in this riser include a temperature of 800°-1,100°F., preferably 850°-1,000°F., a conversion per pass of 20-70 volume percent, preferably 30-50 volume percent, and a vapor velocity of 15-50 feet/second, preferably 20-40 feet/second. When cracking in the dense phase of catalyst, i.e., bed cracking, is employed the operating conditions in the bed include a temperature of 800°-1,150°F., a conversion of 5-30 volume percent and a vapor velocity of 0.5-4 feet/second, preferably 1.3-2.2 feet/second. When the term conversion is used herein in connection with virgin gas oil or cycle gas oil a 430°F conversion is intended and is defined as 100 minus the volume percent of product boiling above 430°F. When conversion is used herein in connection with naphtha or recycle naphtha a 115°F conversion is intended and is defined as 100 minus the volume percent of product boiling above 115°F.

The invention may be understood from the following detailed description taken with reference to FIG. 1 which illustrates and exemplifies a means by which the process of the present invention may be practiced. By describing our invention in this manner it is not intended to restrict the invention thereby since modifications to the following description may be made within the scope of the claims without departing from the spirit thereof.

In this description a preferred embodiment incorporating three risers is employed. A heavy naphtha recycle is combined with fresh naphtha and introduced to one riser, fresh gas oil is fed to another riser and cycle gas oil to the third riser. The combined naphtha streams and the cycle gas oil are subjected to both riser and bed cracking while the fresh gas oil only undergoes riser cracking.

Fresh naphtha obtained from a variety of sources, not shown, passes through line 10 and is combined with a heavy naphtha stream passing through line 12. The heavy naphtha is a cracked stream and is recovered from an effluent stream as described hereinafter. These combined streams are introduced into riser 14 where they are brought into contact with hot regenerated zeolite cracking catalyst from standpipe 16. The resulting suspension of catalyst-naphtha vapor passes up riser 14 to reactor 18. The reactor contains a bed of catalyst particles disengaging as a catalyst disengaging space and is referred to as the dilute phase of catalyst. The effluent from riser 14 discharges into the lower portion of reactor 18 and passes upward through the dense phase of catalytic cracking catalyst 20 effecting further conversion of the combined naphtha streams.

A fresh gas oil stream is introduced through line 24 to riser 26 where it is contacted with regenerated zeolite cracking catalyst from standpipe 28. The resultant catalyst-in-gas oil vapor passes up riser 26 to reactor 18 discharging into the dilute phase of catalyst 22 wherein the catalyst particles disengage from the vapor phase mixture and fall into the bed of catalyst 20. An intermediate gas oil stream, obtained from the effluent from the catalytic cracking unit as described hereinafter, passes through line 30 into riser 32 where it is contacted with hot regenerated zeolite cracking catalyst from standpipe 34. The resultant suspension of catalyst-in-cycle gas oil vapor passes up riser 32 to reactor 18 discharging into the dense phase of catalyst 20. The vapor mixture passes upwardly through the bed of catalyst effecting some further conversion of the cycle gas oil.

Cracked products disengage from the catalyst in the dilute phase of catalyst 22 above the catalyst dense phase 20. These vapors, together with any entrained catalyst, pass through cyclone separators, not shown, wherein this catalyst is substantially separated from the vapors and returned to the catalyst bed. Effluent gases containing the cracked products pass through line 36 to fractionator section 38 wherein the vapor mixture is separated into various products. These products include: (1) a stream of C4 and lighter hydrocarbons passing through line 40, (2) a light naphtha side stream passing through line 42, (3) a heavy naphtha side stream recovered as a product through line 44 or recy-
cled to the fresh naphtha feed line through line 12, (4) a light gas oil side stream passing through line 46, (5) an intermediate gas oil recovered through line 30 as a cycle gas oil stream and recycled as feed to the cata-
lytic cracking unit, and (6) a heavy gas oil bottoms product recovered through line 48.
Catalyst is withdrawn from the bottom of the reactor through slide valves 50 and 52 passing into stripping zone 54 containing baffles 56. Steam is introduced into the lower portion of stripper 54 to remove adsorbed and entrained hydrocarbons from the catalyst as it passes through the stripper. Stripped catalyst is with-
drawn from the bottom of the stripping zone through spent catalyst standpipe 58 and discharges into regen-
erator 60. The catalyst forms a dense bed within regen-
erator 60 and is regenerated therein by contacting it with air to remove the carbon from the catalyst surface. Regenerated catalyst is withdrawn from the bottom of regenerator 60 through standpipes 16, 28 and 34 to supply the hot regenerated catalyst to risers 14, 26, 26 and 32, respectively, as previously described.

The process described above and illustrated in FIG. 1 may be conducted in the apparatus depicted in FIG. 2. Referring to FIG. 2, those items identified by numbers with prime notations designate corresponding portions of the equipment illustrated in FIG. 1 by the same numbers but without the prime notation. Additional numbers are employed in FIG. 2 to designate portions of the apparatus not illustrated or detailed in FIG. 1.

The charge stocks, fresh naphtha and recycle naphtha, cycle gas oil and fresh gas oil in lines 16', 30' and 24', respectively, are introduced into risers 14', 32' and 26', respectively, where they are contacted with zeolite cracking catalyst from standpipes 16', 34' and 28', respectively. The mixtures of oil and catalyst pass through their respective risers discharging into reactor 18' where risers 26' and 32' terminate in downwardly directed outlets 19. The serrated edge provides smooth flow of the effluent from the risers, particularly when the dense bed level 21 fluctuates near the downwardly projecting portion of the riser. The effluents passing from risers 14' and 32' pass upwardly through the dense phase bed in reactor 18' effecting further conversion of the cycle gas oil and naphtha streams. The vaporous effluent from riser 26' discharges into vapor space 22' (also known as the dilute phase of catalyst). Those skilled in the art will readily appreciate that by varying level 21 of the dense phase of catalyst numerous combinations of riser cracking and bed cracking may be obtained for the feedstocks passing through risers 26' and 32'. Thus, with the dense bed level 21 covering the outlets of both risers 26' and 32', the cycle gas oil and fresh gas oil will be riser cracked and bed cracked. With the catalyst level as indicated in FIG. 2, the fresh gas oil will be only riser cracked while the cycle gas oil will be riser cracked and bed cracked. Dropping the level below the outlet of riser 32' results in the cycle gas oil and fresh gas oil only being riser cracked. In all these situations the naphtha stream would be subjected to both riser and bed cracking.

Particles of catalyst in vapors passing through the dil-
ute phase of catalyst 32' disengage from the vapors and fall back into the dense bed of catalyst. The vapors and entrained catalyst pass through cyclone 35 where the entrained catalyst is separated and returned to dense bed 20' through dip leg 37 and valve 39. A single cyclone has been shown for clarity but it will be under-
stood that several cyclones may be assembled in series to achieve the desired separation. Effluent gases pass from cyclone 35 into plenum chamber 41 from where the gases are discharged from the reactor through line 43 to product recovery, not shown, where the conver-
sion products are recovered and separated into desired products and recycle streams, usually by distillation fa-
cilities well known in the art.

Steam introduced to steam ring 45 in the lower por-
tion of the reactor strips the catalyst as it passes down-
wardly through the bottom portion of the reactor through slide valves 50' and 52' into stripping chamber 54'. Stripping chamber 54' is provided with baffles 56' attached to the wall of stripper 54' and riser 14'. As the catalyst cascades down over the baffles steam intro-
duced through steam rings 53 and 55 displaces and re-
moved adsorbed and entrained hydrocarbon vapors which pass upwardly through the stripper venting through stripper vent line 57 and discharge into vapor space 22' of reactor 18'.

Stripped catalyst is withdrawn from the bottom of stripping chamber 54' passing through line 58' and dis-
charges into regenerator 60'. Air introduced through air ring 62 at the bottom of regenerator 60' burns the carbon from the surface of the catalyst in dense bed 64. The resulting flue gas passes upwardly entering cyclone 70 where entrained catalyst is separated and returned to catalyst bed 64 through dip leg 72. Flue gas passes from the cyclone into plenum chamber 74 and through flue gas line 76 to vent facilities. Regenerated catalyst is withdrawn from the bottom of the regenerator at rates controlled by slide valves 80, 82 and 84 to supply hot regenerated catalyst to standpipes 28', 34' and 16', respectively, as described above.

The following exemplifies the practice of our invention and its advantages over alternative or prior art processes. A series of four runs are performed with a fluidized catalytic cracking unit having two feed risers. The same zeolite-containing cracking catalyst is em-
ployed in all runs. The catalyst consists of a 1:1 weight blend of a zeolite cracking catalyst and a high alumina amorphous cracking catalyst. The zeolite catalyst comprises 18 weight percent of a type X zeolite in a silica-
alumina matrix and has a rare earth content of about 2.9 weight percent. The amorphous silica-alumina cata-
lyst has a high alumina content, a surface area of about 140 m²/g, and a pore volume of about 0.44 cc/g. In each of the runs fresh naphtha and fresh gas oil serve as the feedstocks. The properties of the naphtha, a heavy straight run gasoline, are set forth in Table I while those of the gas oil are presented in Table II.

**TABLE I**

<table>
<thead>
<tr>
<th>Feedstock</th>
<th>Analysis</th>
<th>Heavy Straight</th>
<th>Run Gasoline</th>
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<tr>
<td>Gravity, °API</td>
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<tr>
<td>Sulfur, X-Ray, wt. %</td>
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<td>Research Octane, clear</td>
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<tr>
<td>Research Octane, plus 3 cc TEL</td>
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TABLE I-Continued

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<tr>
<th>Feedstock Analysis</th>
<th>Heavy Straight Run Gasoline</th>
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<td>FIA-MS, Vol. %</td>
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<td>Paraffins</td>
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(1) Octane rating with addition of 3 cc tetraethyllead per gallon

TABLE II

<table>
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<tr>
<th>Feedstock Analysis</th>
<th>Gas Oil</th>
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<tr>
<td>Gravity, °API</td>
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<tr>
<td>10 mm Vacuum Distillation, °F.*</td>
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<tr>
<td>10</td>
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<tr>
<td>50</td>
<td>577</td>
</tr>
<tr>
<td>90</td>
<td>632</td>
</tr>
<tr>
<td>EP</td>
<td>767</td>
</tr>
</tbody>
</table>

°API Atmospheric equivalent temperatures

In Runs 1 and 3 the gas oil and naphtha are combined in a ratio of 3:1 and introduced into a single riser. Both runs are made at a riser temperature of 920°F. Runs 1 and 3 are the same in all respects except that in Run 3, a heavy naphtha fraction is recovered from the reactor effluent and recycled to the feed riser. In Runs 2 and 4, demonstrating the process of our invention, the gas oil and naphtha are segregated and introduced into separate risers. Again, both runs are made with riser temperatures of 920°F. The effluent from the naphtha riser is subjected to additional bed cracking at 920°F. to obtain increased conversion of the naphtha. Run 2 employs no naphtha recycle, but in Run 4 a heavy naphtha cut is taken from the reactor effluent and recycled to the naphtha riser.

Operating data and results of these runs are presented in Table III, below.

TABLE III

<table>
<thead>
<tr>
<th>Run No.</th>
<th>Total Fresh Feed Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>75:25 wt. Blend</td>
</tr>
<tr>
<td></td>
<td>Gas Oil &amp; Naphtha</td>
</tr>
<tr>
<td></td>
<td>1</td>
</tr>
<tr>
<td></td>
<td>2</td>
</tr>
<tr>
<td></td>
<td>3</td>
</tr>
<tr>
<td></td>
<td>4</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Feed to:</th>
<th>75:25 wt. Blend</th>
</tr>
</thead>
<tbody>
<tr>
<td>Naphtha Recycle, Vol. % Total Fresh Feed</td>
<td>0.0</td>
</tr>
<tr>
<td>Temperature — All Risers, °F. 430°F, Conversion, Vol. %</td>
<td>920</td>
</tr>
<tr>
<td>Total Fresh Feed</td>
<td>78.2</td>
</tr>
<tr>
<td>Yields, Wt. %</td>
<td>4.8</td>
</tr>
<tr>
<td>Dry Gas (C3 and lighter)</td>
<td>1.9</td>
</tr>
</tbody>
</table>

Notes: 1. Net yield. Does not include material consumed in producing alkylate.
2. Debutanized naphtha.
3. RON + 3 cc TEL — Research Octane Number with addition of 3 cc tetraethyllead per gallon.
ond mixture through said second elongated reaction zone under gas oil cracking conditions.

e. discharging the effluents from said first and second reaction zones into a catalyst phase in said reactor, said effluents comprising vaporous reaction mixture and catalyst, and

f. recovered from the vaporous reaction mixture in a dilute phase of catalyst in said reactor a fraction boiling in the range of 100° to 450°F, having an octane rating higher than said fresh naphtha stream and a fraction boiling below 100°F.

2. A process according to claim 1 wherein the naphtha cracking conditions of step (a) include a temperature of 750°-1,300°F., a conversion of 25-80 volume percent and a vapor velocity of 15-50 feet/second and the gas oil cracking conditions of step (b) include a temperature of 840°-1,100°F., a conversion of 30-80 volume percent and a vapor velocity of 15-50 feet/second.

3. A process according to claim 2 wherein the effluents from the first and second elongated reaction zones are discharged into a dilute phase of catalyst.

4. A process according to claim 2 wherein the effluent from the first elongated reaction zone is discharged into a dilute phase of catalyst and the effluent from the second elongated reaction zone is discharged into a dense phase of catalyst, said vaporous reaction mixture from said second elongated reaction zone passing through said dense phase under catalytic cracking conditions effecting an additional conversion of 5-30 volume percent and discharging into a dilute phase of catalyst.

5. A process according to claim 2 wherein the effluent from the second elongated reaction zone is discharged into a dense phase of catalyst and the effluent from the first elongated reaction zone is discharged into a dense phase of catalyst, said vaporous reaction mixture from said first elongated reaction zone passing through said dense phase under catalytic cracking conditions effecting an additional conversion of 5-30 volume percent and discharging into a dilute phase of catalyst.

6. A process according to claim 2 wherein the effluents from the first and second elongated reaction zones are discharged into a dense phase of catalyst, said vaporous reaction mixture from said first and second reaction zones passing through said dense phase under catalytic cracking conditions effecting an additional conversion of 5-30 volume percent and discharging into a dilute phase of catalyst.

7. A process according to claim 3 including the following additional steps:

g. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling above about 500°F.,

h. introducing the fraction from step (g) and a third portion of freshly-regenerated zeolite cracking catalyst into a third elongated reaction zone to form a third mixture consisting essentially of said fraction from step (g) and said third portion of catalyst,

i. passing reactants consisting essentially of third mixture through said third elongated reaction zone under cycle gas oil cracking conditions, and

j. discharging the effluent from said third elongated reaction zone into a dense phase of catalyst in said reactor, said effluent comprising vaporous reaction mixture and catalyst, said vaporous reaction mixture from said third elongated reaction zone passing through said dense phase under catalytic cracking conditions effecting an additional conversion of 5-30 volume percent and discharging into a dilute phase of catalyst.

8. A process according to claim 4 including the following additional steps:

g. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling above about 500°F.,

h. introducing the fraction from step (g) and a third portion of freshly-regenerated zeolite cracking catalyst into a third elongated reaction zone to form a third mixture consisting essentially of said fraction from step (g) and said third portion of catalyst,

i. passing reactants consisting essentially of said third mixture through said third elongated reaction zone under cycle gas oil cracking conditions, and

j. discharging the effluent from said third elongated reaction zone into a dense phase of catalyst in said reactor, said effluent comprising vaporous reaction mixture and catalyst, said vaporous reaction mixture from said third elongated reaction zone passing through said dense phase under catalytic cracking conditions effecting an additional conversion of 5-30 volume percent and discharging into a dilute phase of catalyst.

9. A process according to claim 5 including the following additional steps:

g. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling above about 500°F.,

h. introducing the fraction from step (g) and a third portion of freshly-regenerated zeolite cracking catalyst into a third elongated reaction zone to form a third mixture consisting essentially of said fraction from step (g) and said third portion of catalyst,

i. passing reactants consisting essentially of said third mixture through said third elongated reaction zone under cycle gas oil cracking conditions, and

j. discharging the effluent from said third elongated reaction zone into a dense phase of catalyst in said reactor, said effluent comprising vaporous reaction mixture and catalyst, said vaporous reaction mixture from said third elongated reaction zone passing through said dense phase under catalytic cracking conditions effecting an additional conversion of 5-30 volume percent and discharging into a dilute phase of catalyst.

10. A process according to claim 6 including the following additional steps:

g. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling above about 500°F.,

h. introducing the fraction from step (g) and a third portion of freshly-regenerated zeolite cracking catalyst into a third elongated reaction zone to form a third mixture consisting essentially of said fraction from step (g) and said third portion of catalyst,

i. passing reactants consisting essentially of said third mixture through said third elongated reaction zone under cycle gas oil cracking conditions, and
j. discharging the effluent from said third elongated reaction zone into a dense phase of catalyst in said reactor, said effluent comprising vaporous reaction mixture and catalyst, said vaporous reaction mixture from said third elongated reaction zone passing through said dense phase under catalytic cracking conditions effecting an additional conversion of 5–30 volume percent and discharging into a dilute phase of catalyst.

11. A process according to claim 4 wherein the catalytic cracking conditions in the dense phase include a temperature of 800°–1,150°F and a vapor velocity of 0.5–4 feet/second.

12. A process according to claim 5 wherein the catalytic cracking conditions in the dense phase include a temperature of 800°–1,150°F and a vapor velocity of 0.5–4 feet/second.

13. A process according to claim 6 wherein the catalytic cracking conditions in the dense phase include a temperature of 800°–1,150°F and a vapor velocity of 0.5–4 feet/second.

14. A process according to claim 7 wherein the cycle gas oil cracking conditions of step (f) include a temperature of 800°–1,100°F, a conversion of 20–70 volume percent and a vapor velocity of 15–50 feet/second and the catalytic cracking conditions in the dense phase include a temperature of 800°–1,150°F and a vapor velocity of 0.5–4 feet/second.

15. A process according to claim 8 wherein the cycle gas oil cracking conditions of step (f) include a temperature of 800°–1,100°F, a conversion of 20–70 volume percent and a vapor velocity of 15–50 feet/second and the catalytic cracking conditions in the dense phase include a temperature of 800°–1,150°F and a vapor velocity of 0.5–4 feet/second.

16. A process according to claim 9 wherein the cycle gas oil cracking conditions of step (f) include a temperature of 800°–1,100°F, a conversion of 20–70 volume percent and a vapor velocity of 15–50 feet/second and the catalytic cracking conditions in the dense phase include a temperature of 800°–1,150°F and a vapor velocity of 0.5–4 feet/second.

17. A process according to claim 10 wherein the cycle gas oil cracking conditions of step (f) include a temperature of 800°–1,100°F, a conversion of 20–70 volume percent and a vapor velocity of 15–50 feet/second and the catalytic cracking conditions in the dense phase include a temperature of 800°–1,150°F and a vapor velocity of 0.5–4 feet/second.

18. A process for the catalytic cracking of naphtha and gas oil with a zeolite cracking catalyst in a fluid catalytic cracking unit comprising a reactor, a regenerator and a multiplicity of elongated reaction zones wherein said reactor contains a dense phase and a dilute phase of said catalyst and said elongated reaction zones terminate at said reactor which comprises:

a. introducing a low octane fresh naphtha stream, the hydrocarbon fraction boiling above about 500°F, of step (h) herein and a first portion of a freshly regenerated zeolite cracking catalyst into a first elongated reaction zone to form a fourth mixture consisting essentially of said fresh naphtha stream, said fraction boiling above about 500°F and said first portion of catalyst,

b. passing reactants consisting essentially of said fourth mixture through said first elongated reaction zone under naphtha cracking conditions,
mixture through said first elongated reaction zone under naphtha cracking conditions,
c. discharging the effluent from said first elongated reaction zone into a dense phase of said catalyst in said reactor, said effluent comprising vaporous reaction mixture and catalyst,
d. passing vaporous reaction mixture from step (c) through said dense phase under catalytic cracking conditions and discharging into a dilute phase of catalyst,
e. introducing a virgin gas oil and a second portion of freshly-regenerated zeolite cracking catalyst into a second elongated reaction zone to form a seventh mixture consisting essentially of said virgin gas oil and said second portion of catalyst,
f. passing reactants consisting essentially of said seventh mixture through said second elongated reaction zone under gas oil cracking conditions,
g. discharging the effluent from said second elongated reaction zone into a catalyst phase in said reactor, said effluent comprising vaporous reaction mixture and catalyst,
h. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling between 100° and 450°F, and
i. recovering as products from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a fraction boiling in the range of 100° to 450°F, having an octane rating higher than said fresh naphtha stream and a fraction boiling below 100°F.
23. A process according to claim 22 wherein the naphtha cracking conditions of step (b) include a temperature of 750°F-1,300°F, a conversion of 25-80 volume percent and a vapor velocity of 15-50 ft/sec,
the gas oil cracking conditions of step (f) include a temperature of 840°F-1,100°F, a conversion of 30-80 volume percent and a vapor velocity of 15-50 ft/sec and the catalytic cracking conditions in the dense phase include a temperature of 800°F-1,150°F, a vapor velocity of 0.5-4 ft/sec, and an additional conversion of 5-30 volume percent.
24. A process according to claim 23 wherein the effluent from the second elongated reaction zone is discharged into a dilute phase of catalyst.
25. A process according to claim 23 wherein the effluent from the second elongated reaction zone is discharged into a dense of catalyst, said vaporous reaction mixture from said second elongated reaction zone passing through said dense phase under catalytic cracking conditions and discharging into a dilute phase of catalyst.
26. A process according to claim 24 including the following additional steps:
j. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling above 500°F,
k. introducing said fraction of step (j) and a third portion of freshly-regenerated zeolite cracking catalyst into a third elongated reaction zone to form an eighth mixture consisting essentially of said fraction from step (j) and said third portion of catalyst,
l. passing reactants consisting essentially of said eighth mixture through said elongated reaction zone under cycle gas oil cracking conditions, and
m. discharging the effluent from said third elongated reaction zone into a dense phase of catalyst in said reactor, said effluent comprising vaporous reaction mixture and catalyst, said vaporous reaction mixture from said third elongated reaction zone passing through said dense phase under catalytic cracking conditions and discharging into a dilute phase of catalyst.
27. A process according to claim 25 including the following additional steps:
j. recovering from the vaporous reaction mixture in the dilute phase of catalyst in said reactor a hydrocarbon fraction boiling above 500°F,
k. introducing said fraction of step (j) and a third portion of freshly-regenerated zeolite cracking catalyst into a third elongated reaction zone to form an eighth mixture consisting essentially of said fraction from step (j) and said third portion of catalyst,
l. passing reactants consisting essentially of said eighth mixture through said elongated reaction zone under cycle gas oil cracking conditions, and
m. discharging the effluent from said third elongated reaction zone into a dense phase of catalyst in said reactor, said effluent comprising vaporous reaction mixture and catalyst, said vaporous reaction mixture from said third elongated reaction zone passing through said dense phase under catalytic cracking conditions and discharging into a dilute phase of catalyst.
28. A process according to claim 26 wherein the cycle gas oil cracking conditions of step (1) include a temperature of 800°F-1,100°F, a conversion of 20-70 volume percent and a vapor velocity of 15-50 ft/sec.
29. A process according to claim 27 wherein the cycle gas oil cracking conditions of step (1) include a temperature of 800°F-1,100°F, a conversion of 20-70 volume percent and a vapor velocity of 15-50 ft/sec.

* * * * *