IMPROVED SHORT CONTACT TIME FCC PROCESS

FIELD OF THE INVENTION

[0001] This invention relates to a short contact time (SCT) fluid catalytic cracking (FCC) process for improving the yield of distillates and gasoline. More particularly, an SCT FCC process combines staging the FCC conversion process with interstage molecular separation of multi-ring aromatic species wherein separation of FCCU bottoms and recycling the separated stream containing saturates and 1- and 2-ring aromatics to the FCC unit results in improved yields of gasoline and other distillates while decreasing bottoms yields.

BACKGROUND OF THE INVENTION

[0002] Conversion of petroleum feeds to more valuable products is important to petroleum processes such as fluidized catalytic cracking (FCC) and coking. In the FCC process, high molecular weight feeds are contacted with fluidized catalyst particles in the riser reactor of the FCC unit. The contacting between feed and catalyst is controlled according to the type of product desired. In catalytic cracking of the feed, reactor conditions such as temperature and contact time are controlled to maximize the products desired and minimize the formation of less desirable products such as light gases and coke.

[0003] Since contacting between catalyst and feed in the FCC reactor is typically in the order of a few seconds, an important factor governing the efficiency of the cracking process is the catalyst. Catalysts for the FCC process are well known and may be either amorphous or crystalline. Catalyst entering the FCC reactor is typically fluidized using steam, hydrocarbon gases generated during the

cracking process or some combination thereof. The reaction of catalyst and feed generates large volumes of gaseous hydrocarbons and spent catalyst bearing coke deposits. The gas/solid mixture is passed to separators, typically cyclones, where spent catalyst is separated from gases. Gases are then processed to recover desired hydrocarbons and spent catalyst sent for regeneration.

[0004] Because of the short contacting time between feed and catalyst, the condition of the feed is also important. The type of feed injection can have an impact on the product slate produced by the FCC reactor. There are two pathways for the feed to crack into gaseous hydrocarbons, i.e., catalytic and thermal. Thermal cracking in an FCC unit is generally undesirable as this type of cracking can result in the generation of light gases such as methane in addition to coke. Short contact time FCC processes have been developed for handling heavier feeds. These short contact times can result in improved gasoline yields and lower coke and dry gas make from such heavier feeds.

[0005] SCT FCC has been widely implemented as a means to decrease non-selective reactions, suppress undesirable hydrogen transfer reactions and yield a more valuable product slate. High conversion of feed is in part due to control of reaction parameters such as reactor temperature, catalyst activity, catalyst circulation rate or some combination thereof. Recent improvements to the SCT FCC process have focused on catalysts and hardware. One method for improving the SCT FCC process involves forming a falling curtain of FCC catalyst that is then contacted with feed. Other improvements involve hardware modifications such as dual risers. Still other improvements are directed to process modifications.

[0006] There is still a need to improve the SCT FCC process which improvement can be implemented using existing catalysts and without the need for extensive modifications to the FCC unit itself.

SUMMARY OF THE INVENTION

[0007] One embodiment of the invention relates to a fluid catalytic cracking process which comprises:

(a) contacting hydrocarbon feed with regenerated catalyst in a fluid catalytic cracking reactor operating under short contact time cracking conditions including a contact time of less than 5 seconds to form a cracked product;

(b) separating cracked product from catalyst;

(c) passing the separated cracked product into a fractionating zone and separating from the cracked product a bottoms stream;

(d) passing the bottoms stream to a separation zone and separating the bottoms stream into a stream containing saturates and 1- and 2-ring aromatics and a stream containing predominantly aromatics having 3 or more rings;

(e) combining the stream containing saturates and 1- and 2-ring aromatics with hydrocarbon feed in step (a) to form a combined feed and contacting combined feed with catalyst of step (a).

[0008] Another embodiment relates to a fluid cracking process which comprises:

(a) contacting hydrocarbon feed with regenerated catalyst in a fluid catalytic cracking reactor operating under short contact time cracking conditions including a contact time of less than 5 seconds to form a cracked product;

(b) separating cracked product from catalyst;

(c) passing the separated cracked product into a fractionating zone and separating from the cracked product a bottoms stream;

(d) passing the bottoms stream to a separation zone and separating the bottoms stream into a stream containing saturates and 1- and 2-ring aromatics and a stream containing predominantly aromatics having 3 or more rings;

(e) passing the stream containing saturates and 1- and 2-ring aromatics to step (a) wherein hydrocarbon feed and the stream containing saturates and 1- and 2-ring aromatics are separately contacted with regenerated catalyst.

BRIEF DESCRIPTION OF THE DRAWING

[0009] The figure is a schematic drawing of the fluid catalytic cracking process having a bottoms separation step.

DETAILED DESCRIPTION OF THE INVENTION

[0010] A conventional FCC process includes a riser reactor and a regenerator wherein petroleum feed is injected into the reaction zone in the riser containing a bed of fluidized cracking catalyst particles. The catalyst particles typically contain zeolites and may be fresh catalyst particles, catalyst particles from a catalyst regenerator or some combination thereof. Gases that may be inert gases, hydrocarbon vapors, steam or some combination thereof are normally employed as lift gases to assist in fluidizing the hot catalyst particles.

[0011] Catalyst particles that have contacted feed produce product vapors and catalyst particles containing strippable hydrocarbons as well as coke. The catalyst exits the reaction zone as spent catalyst particles and is separated from the reactor's effluent in a separation zone. The separation zone for separating spent catalyst particles from reactor effluent may employ separation devices such as cyclones. Spent catalyst particles are stripped of strippable hydrocarbons using a stripping agent such as steam. The stripped catalyst particles are then sent to a regeneration zone in which any remaining hydrocarbons are stripped and coke is removed. In the regeneration zone, coked catalyst particles are contacted with an oxidizing medium, usually air, and coke is oxidized (burned) at high temperatures such as 650°C to 760°C. The regenerated catalyst particles are then passed back to the riser reactor.

[0012] Suitable hydrocarbon feeds for the catalytic cracking process described herein include natural and synthetic hydrocarbonaceous oils boiling in the range of 221°C (430°F) to 566°C (1050°F), such as gas oil; heavy hydrocarbonaceous oils comprising materials boiling above 566°C (1050°F); heavy and reduced petroleum crude oil; petroleum atmospheric distillation bottoms; petroleum vacuum distillation bottoms; pitch, asphalt, bitumen, other heavy hydrocarbon residues; tar sand oils; shale oil; liquid products derived from coal liquefaction processes, naphtha, and mixtures thereof.

[0013] FCC catalysts may be amorphous, e.g., silica-alumina, crystalline, e.g., molecular sieves including zeolites, or mixtures thereof. A preferred catalyst particle comprises: (a) an amorphous, porous solid acid matrix, such as alumina, silica-alumina, silica-magnesia, silica-zirconia, silica- thoria, silica-beryllia, silica-titania, silica-alumina-rare earth and the like; and (b) a zeolite such as faujasite.

The matrix can comprise ternary compositions, such as silica-alumina-thoria, silica-alumina-zirconia, magnesia and silica-magnesia-zirconia. The matrix may also be in the form of a cogel. Silica-alumina is particularly preferred for the matrix, and can contain 10 to 40 wt.% alumina. As discussed, promoters can be added.

[0014] The primary catalyst zeolite component includes zeolites which are iso-structural to zeolite Y. These include the ion-exchanged forms such as the rare-earth hydrogen and ultrastable (USY) form. The zeolite may range in crystallite size from 0.1 to 10 microns, preferably from 0.3 to 3 microns. The relative concentrations of zeolite component and matrix on an anhydrous basis may vary widely, with the zeolite content ranging from 1 to 100, preferably 10 to 99, more usually from 10 to 80, percent by weight of the dry composite.

[0015] The amount of zeolite component in the catalyst particle will generally range from 1 to 60 wt.%, preferably from 5 to 60 wt.%, and more preferably from 10 to 50 wt.%, based on the total weight of the catalyst. As discussed, the catalyst is typically in the form of a catalyst particle contained in a composite. When in the form of a particle, the catalyst particle size will range from 10 to 300 microns in diameter, with an average particle diameter of 60 microns. The surface area of the matrix material after artificial deactivation in steam will be ≤ 350 m^2/g, preferably 50 to 200 m^2/g, more preferably from 50 to 100 m^2/g. While the surface area of the catalysts will be dependent on such things as type and amount of zeolite and matrix components used, it will usually be less than 500 m^2/g, preferably from 50 to 300 m^2/g, more preferably from 50 to 250 m^2/g, and most preferably from 100 to 250 m^2/g.

[0016] The cracking catalyst may also include an additive catalyst in the form of a medium pore zeolite having a Constraint Index (which is defined in United States

Patent Number 4,016,218, Haag, et al.) of 1 to 12. Suitable medium pore zeolites include ZSM-5, ZSM-11, ZSM-12, ZSM-22, ZSM-23, ZSM-35, ZSM-48, ZSM-57, SH-3 and MCM-22, either alone or in combination. Preferably, the medium pore zeolite is ZSM-5.

[0017] Short contact time means that the hydrocarbon feed will contact the cracking catalyst for less than five seconds. Preferably, in the short contact time reaction step, the hydrocarbon feed will contact the cracking catalyst for 1 to 5 seconds.

[0018] The short contact time reaction step can be achieved using any of the known processes. For example, in one embodiment, a close coupled cyclone system effectively separates the catalyst from the reacted hydrocarbon to quench the cracking reaction. See, for example, Exxon's United States Patent Number 5,190,650, Tammera, et al., of which the detailed description is incorporated herein by reference.

[0019] Short contact time can be achieved in another embodiment by injecting a quench fluid directly into the riser portion of the reactor. The quench fluid is injected into the appropriate location to quench the cracking reaction in less than one second. See, for example, United States Patent Number 4,818,372, Mauleon, et al., of which the detailed description is incorporated herein by reference. Preferred as quench fluids are such examples as water or steam or any hydrocarbon that is vaporizable under conditions of injection, and more particularly the gas oils from coking or visbreaking, catalytic cycle oils, and heavy aromatic solvents as well as certain deasphalted fractions extracted with a heavy solvent.

[0020] In yet another embodiment, short contact time can be achieved using a downflow reactor system. In downflow reactor systems, contact time between catalyst and hydrocarbon can be as low as in the millisecond range. See, for example, United States Patent Numbers 4,985,136, Bartholic; 4,184,067, Cress; and 4,695,370, Galtier, et al., of which the detailed descriptions of each are incorporated herein by reference. Very short contact times can also be achieved by contacting feed with a falling curtain of catalyst as disclosed in United States Patent Number 4,985,136, Bartholic, which is incorporated herein by reference.

[0021] Short contact time conditions include riser outlet temperatures from 482°C to 621°C (900 to 1150°F), pressures from 0 to 100 psig (101 to 790 kPa) and residence times from 1 to 5 seconds.

[0022] The cracked products and catalyst exiting from the riser are typically passed through cyclones to separate cracked product from catalyst. The separated cracked product is then passed to a fractionation zone. Conventional fractionation involves the use of distillation columns. In the distillation process, cracked products are usually separated into light and medium distillates including light olefins and naphtha including fractions suitable for blending into gasoline. Other distillates include products such as light cycle oils. Normally gaseous products are taken off the top of the column as overheads. The bottoms from the fractionation zone are heavier boiling components and have a boiling point greater than 343°C. These bottoms are passed to a separation zone.

[0023] The separation zone involves the use of known separation techniques including solvent extraction, membrane separation, and liquid chromatography. The purpose of the separation zone is to separate saturates and 1- and 2-ring aromatics in a first fraction from 3+ ring aromatics in a second fraction. It is

possible that there will be some overlap between the two fractions from the separation. The concentration of saturates plus 1- and 2-ring aromatics in the first fraction ranges from 25 wt.% to 85 wt.%, the balance being 3+ ring aromatics and other polar species. By the term "saturates" is meant saturated ring compounds as well as paraffins and isoparaffins. This first fraction is either combined with fresh feed or separately injected into the reactor. The second fraction containing predominantly aromatics having three or more rings is the net bottoms stream from the process. By predominantly is meant the concentration of 3+ ring aromatics and polar species is greater than 75 wt.% and preferably greater than 95 wt.% of the stream.

[0024] One means of separating saturates and 1- and 2-ring aromatics from 3+ ring aromatics is solvent extraction. Preferred solvents include dimethyl sulfoxide, dimethyl formamide, n-methyl pyrrolidone (NMP), phenol and furfural, especially dimethyl sulfoxide. The solvent extraction process selectively dissolves the aromatic components in an extract phase while leaving the more paraffinic components in a raffinate phase. Naphthenes are distributed between the extract and raffinate phases. By controlling the solvent to oil ratio, extraction temperature and method of contacting the stream to be extracted with solvent, one can control the degree of separation between the extract and raffinate phases. In carrying out the extraction process with NMP, water may be added to the extraction solvent in amounts ranging from 1 to 10 vol.% such that the extraction solvent to the extraction tower contains from 3 to 10 vol.% water, preferably 4 to 7 vol.% water. In general, feed to the extraction tower is added at the bottom of the tower and extraction/water solvent mixture added at the top and the feed and extraction solvent contacted in counter-current flow. The extraction solvent containing added water may be injected at different levels if the extraction tower contains multiple trays for solvent extraction. The use of added water in the extraction solvent

permits the use of low quality feeds while maximizing the paraffin content of the raffinate and the 3+ multi-ring compounds content of the extract. Solvent extraction conditions include a solvent to oil ratio of from 0.5 to 5.0, preferably 1 to 3 and extraction temperatures of from 40°C to 120°C., preferably 50°C to 100°C.

[0025] Another means for separating saturates and 1- and 2-ring aromatics is chromatography. Chromatography is used to separate the constituents of a mixture based on one or more properties for the particular chromatography technique. A sample of the mixture is placed in the top of a chromatography column that contains a chromatographic medium, or matrix, that is capable of fractionating the mixture. While there are different chromatographic techniques that may be used, for practical reasons involving the high throughput demands in a FCC process, the preferred chromatographic method is simulated moving bed (SMB) liquid chromatography which may be applied on a commercial scale and is described in United States Patent Number 2,985,589, Broughton, et al.. By intermittent switching of inlet and outlet ports in a fixed, high pressure packed column the SMB approximates countercurrent flow of solid adsorbent and the hydrocarbon stream in an actual moving bed. Various embodiments of the SMB concept have evolved over the years. For a packed silica column of this application the "raffinate" stream is enriched in saturates and 1-and 2-ring aromatics, while the "extract" is predominantly 3+ ring aromatics and polars. Suitable desorbent streams are lower boiling single or two-ring aromatics with final boiling points less than 290°C. Feed for the SMB separation process is the hot fractionator bottoms stream with initial boiling point greater than 343°C.

[0026] Another separation technique for performing the separation of saturates and 1- and 2-ring aromatics is membrane separation. Polymeric membranes operating in the pervaporation or perstraction mode may be used to concentrate

aromatics in the permeate phase and produce a saturates-rich retentate. The highly aromatic permeate is then further separated using for example a second "functionalized" membrane system wherein the single and two ring aromatics species are preferentially removed and blended into the saturates-rich retentate for either blending with fresh feed or separately recracking in the riser reactor. The polymeric membranes may be supported on porous metals or ceramics which provide necessary stability for high pressure operation.

[0027] Referring now to the Figure which exemplifies an embodiment of the invention, FCC reactor 10 includes a riser 12 and stripper 14. Fresh feed is fed through line 16 to the bottom portion designated as 18 of riser 12. Fresh feed contacts fluidized hot catalyst from regenerator 20 which is connected to riser 12 through regenerated catalyst standpipe 22, slide valve 24 and J-bend 26. Spent catalyst from riser 12 enters reactor 10 where spent catalyst is separated from product vapors by means of cyclones (not shown). Spent catalyst is then stripped of additional product and returned to regenerator 20 through stripper 14, spent catalyst standpipe 28, slide valve 30 and spent catalyst J-bend 32. Product vapors exit reactor 10 through line 34 and are conducted to fractionator 40 which is a distillation unit. Fractionator 40 separates product from reactor 10 into a C_4- gas fraction that exits fractionator 40 through line 42. Another fraction from the fractionator is a naphtha fraction 44 that may be used as a fuel such as gasoline or blended into a fuel. A further cut from fractionator 40 is a light cycle oil 46. The bottoms from fractionator 40 is sent through line 48 to a separation unit 50. Separation unit 50 separates the bottoms into a paraffins and 1- and 2-ring aromatics rich stream and a 3+ ring aromatics rich stream. The paraffins and 1- and 2-ring aromatics rich stream is sent for recycle through line 54 where it is combined with fresh feed 16. The 3+ ring aromatics rich stream is removed through line 52.

The recycle stream in line 54 is shown in the Figure as being combined with fresh feed. The recycle stream may, however, be separately added to riser 12.

[0028] The invention is further illustrated by the following example.

EXAMPLE

[0029] This example relates to a short contact time FCC riser reactor. Vapor residence times in the riser are in the range of 3 to 4 seconds. Process performance data collected during a commercial monitoring test period were used to tune FCC model software program for representing base case operation. Selected feedstock inspections are listed in Table 1:

Table 1

Feedstock Properties

Gravity	API	19
Aniline Point	°F (°C)	165 (74)
Sulfur	WT%	2.921
Nitrogen	PPMW	1897
Basic Nitrogen	PPMW	728
Total P/N/A	WT%	
Total Paraffins		11.64
Total Naphthenes		26.46
Total Aromatics		61.90
HPLC-2 Analysis	WT%	
Saturates		38.43
Aromatic Ring Class 1		23.97
Aromatic Ring Class 2		13.92
Aromatic Ring class 3		11.70
Aromatic Ring Class 4		6.78
Polars		5.20
GC Distillation	°F (°C)	
IBP		511 (266)
5% Off		622 (328)
10% Off		668 (353)
30% Off		749 (398)
50% Off		813 (434)
70% Off		882 (472)
90% Off		976 (524)
95% Off		1019 (548)
EP		1093 (589)

[0030] Commercially available equilibrium catalyst properties are listed in Table 2. Using these feed and catalyst properties, the FCC model software was used to develop a base product yield slate for the existing unit configuration and operating conditions. These are listed in Table 3.

[0031] To demonstrate this invention the separation step in the model was operated with 100 vol.% dimethylsulfoxide (DMSO) on fractionator bottoms to remove 40 wt.% 3-ring aromatics, 85 wt.% 4-ring aromatics, and 85 wt.% polars. At steady state, the bottoms stream rate to extraction is 30 wt.% of fresh feed, and with a 1.3 raffinate/extract (wt.) ratio, there is a 17% increase in total feed rate to the riser. The basis for comparison was constant coke yield.

Table 2

Equilibrium Catalyst Properties

Catalyst Activity	MAT	68
Total Surface Area	M2/G	183
Matrix Surface Area	M2/G	63
Zeolite Surface Area	M2/G	120
Unit Cell Size (UCS)	A	24.25
Rare Earth (RE203)	WT%	0.56
Alumina (Al_2O_3)	WT%	40
Nickel	PPMW	190
Vanadium	PPMW	1660
Sodium	WT%	0.22
Bulk Density (ABD)	G/CC	0.88
Pore Volume	CC/G	0.33
Particle Size	MICRONS	72

[0032] With higher feed rate of poorer quality (vs. fresh) feed, the present process experienced a decrease in catalyst/oil ratio. Heat balance was achieved by switching from partial burn to full burn conditions in the regenerator, thereby raising catalyst temperature 42°F (6°C). Air blower rate was increased from 37.9 to 43.7 MSCFM. Riser top temperature was decreased 10°F (-12°C) to match base coke yield. Riser vapor velocity only increased 7% over the base case with a corresponding decrease in residence time. If a feed preheat furnace were available

the feed temperature could have been raised with less impact on regenerator temperature. Heat integration was assumed to increase raffinate temperature to blend with total feed at 330°F (166°C).

[0033] Referring to Table 3, the comparative yields show a dramatic (30 relative %) decrease in bottoms yield with 343°C- conversion increased from 82 to 87.3 wt.% for this invention. Raffinate recracking resulted in a 25 (rel) % increase in light cycle oil and a 3 (rel) % increase in gasoline yield. The light gas yields are comparable with slightly higher dry gas (5 rel%), a minor increase in LPG olefins, but a decrease (8 rel%) in LPG sats.

[0034] Gasoline octane quality is marginally lower for this invention , with (R+M)/2 decreased from 88 to 87.9. Conversely, cetane quality is incrementally increased by 0.8 CI. Aromaticity of the bottoms stream is significantly increased, with aromatic carbon content elevated from 70.9 to 83.3 wt.% and API gravity decreased from -2.3 to -9.7.

[0035] This example demonstrates that even with high base 221°C- conversion levels, crackable molecules remain in the 343°C+ bottoms stream which if segregated and recycled with fresh feed can boost conversion to lighter, more valuable products. This phenomenon is judged applicable to a spectrum of SCT FCC units which have been analyzed with respect to this invention using the same model framework as in this example.

Table 3

Comparative Process Predictions

		BASE	INVENTION	
Riser Conditions				
Riser Top Temperature	°F (°C)	990 (532)	980 (527)	
Mix Point Temperature	°F (°C)	1029 (554)	1022 (550)	
Cat to Oil	RATIO	9.22	7.29	
Catalyst Circulation Rate	TPM	17.9	16.6	
Fresh Feed (Weight)	TPM	1.94	1.94	
Total Feed (Weight)	TPM	1.94	2.27	
Combined Feed Temperature	°F (°C)	330 (166)	330 (166)	
Riser Vapor Velocity	ft./sec. (m/sec.)	47.6 (14.5)	51.0 (15.5)	
Vapor Residence Time	SEC	3.6	3.4	
Catalyst Residence Time	SEC	7.1	6.0	
Regenerator Conditions				
Air Rate (Wet)	MSCFM	37.9	43.7	
Flue Gas Oxygen (Dry)	VOL%	0.0002	1.00	
Flue Gas CO_2 (Dry)	VOL%	14.73	16.75	
Flue Gas CO (Dry)	VOL%	4.32	0.00	
Dense Bed Temperature	°F (°C)	1225 (373)	1267 (386)	
Detailed Yields				**Delta (Inv - Base)**
Hydrogen	WT%	0.05	0.05	0.00
Ammonia	WT%	0.01	0.01	0.00
H_2S	WT%	0.60	0.62	0.02
Methane	WT%	0.78	0.82	0.04
Ethane	WT%	1.29	1.36	0.07
Ethene	WT%	0.91	0.97	0.06
Propane	WT%	1.11	1.01	-0.10
Propene	WT%	3.84	3.88	0.04
N-Butane	WT%	0.56	0.49	-0.07
Iso-Butane	WT%	2.27	2.11	-0.16
Iso-Butenes	WT%	1.75	1.78	0.03
N-Butenes	WT%	3.91	3.94	0.03
C_4 Butadienes	WT%	0.02	0.02	0.00
Gasoline (C_{5+})	WT%	42.60	43.94	1.35
Light Cycle Oil	WT%	15.71	19.65	3.94
Main Column Bottoms	WT%	17.98	12.73	-5.25
Coke Yield	WT%	6.62	6.63	0.00
Conversion @ 430°F (221°C) (Wt.)	WT%	66.31	67.62	1.31
650°F- (379°C-) Conversion	WT%	82.02	87.27	5.25
Gasoline				
Gravity	API	58.4	58.3	
Research Octane (Clear)	R+0	93.5	93.4	
Motor Octane (Clear)	M+0	82.6	82.4	
Light Cycle Oil				
Gravity	API	17.6	17.2	
SimDis D2887 50% Off	°F (°C)	516 (157)	529 (161)	
Cetane Index	CI	23.0	23.8	
Net Bottoms Stream				
Gravity	API	-2.3	-9.7	
Aromatic Carbon (Ca)	%	70.9	83.3	

[0036] The model has also demonstrated that simply increasing intensity parameters to drive down bottoms yield results in loss of gasoline yield due to overcracking and an increase in coke yield. With the process according to the invention, FCC coke yield can be held constant with a large decrease in net bottoms yield. Various modifications of this basic concept are possible within the scope of the instant invention. For example, modifying catalyst formulation and tuning process variables (e.g. catalyst circulation, riser top temperature, etc.) may be used to optimize this invention for site-specific applications. Alternatives to solvent extraction of the bottoms stream such as selective molecular separation techniques may provide the preferred effectiveness ratio [3R+ Aromatics/(Sats + 1R +2R Aromatics)] target in extract of 10-20. While the above example is based upon blending raffinate with fresh feed, in some instances it may be advantageous to recycle raffinate at an alternate location, either upstream or downstream of fresh feed, or potentially segregating raffinate in one or more injectors located on the feed injector ring at the same elevation in the riser.

CLAIMS:

1. A fluid catalytic cracking process which comprises:
   (a) contacting hydrocarbon feed with regenerated catalyst in a fluid catalytic cracking reactor operating under short contact time cracking conditions including a contact time of less than 5 seconds to form a cracked product;
   (b) separating cracked product from catalyst;
   (c) passing the separated cracked product into a fractionating zone and separating from the cracked product a bottoms stream;
   (d) passing the bottoms stream to a separation zone and separating the bottoms stream into a stream containing saturates and 1- and 2-ring aromatics and a stream containing predominantly aromatics having 3 or more rings;
   (e) combining the stream containing saturates and 1- and 2-ring aromatics with hydrocarbon feed in step (a) to form a combined feed and contacting combined feed with catalyst of step (a).

2. A fluid cracking process which comprises:
   (a) contacting hydrocarbon feed with regenerated catalyst in a fluid catalytic cracking reactor operating under short contact time cracking conditions including a contact time of less than 5 seconds to form a cracked product;
   (b) separating cracked product from catalyst;
   (c) passing the separated cracked product into a fractionating zone and separating from the cracked product a bottoms stream;
   (d) passing the bottoms stream to a separation zone and separating the bottoms stream into a stream containing saturates and 1- and 2-ring

aromatics and a stream containing predominantly aromatics having 3 or more rings;

(e) passing the stream containing saturates and 1- and 2-ring aromatics to step (a) wherein hydrocarbon feed and the stream containing saturates and 1- and 2-ring aromatics are separately contacted with regenerated catalyst.

3. The process of claims 1 or 2 wherein the catalyst is amorphous, crystalline or a mixture thereof.

4. The process of claim 3 wherein the amorphous catalyst is silica-alumina.

5. The process of claim 3 wherein the crystalline catalyst is a zeolite which is isostructural to zeolite Y.

6. The process of claim 3 wherein the catalyst includes an additive which is a medium pore zeolite having a Constraint Index of 1 to 12.

7. The process of claims 1 or 2 wherein short contact time conditions include riser outlet temperatures from 482°C to 621°C, pressures from 0 to 100 psig (101 to 790 kPa) and residence times from 1 to 5 seconds.

8. The process of claims 1 or 2 wherein the separation zone comprises solvent extraction, membrane separation or liquid chromatography.

9. The process of claim 8 wherein the separation zone comprises solvent extraction.

10. The process of claim 9 wherein solvent for solvent extraction comprises at least one of dimethyl sulfoxide, dimethyl formamide, n-methyl pyrrolidone, phenol and furfural.

11. The process of claim 8 wherein membrane separation contains supported polymeric membranes.

12. The process of claims 1 or 2 wherein the stream containing saturates and 1- and 2-ring aromatics contains from 25 to 85 wt.% saturates and 1- and 2-ring aromatics.

13. The process of claims 1 or 2 wherein the stream containing predominantly aromatics having 3 or more rings comprises greater than 75 wt.% of aromatics having 3 or more rings.